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Techno-Economic Evaluation of CO₂ Capture in LNG Plants

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This report consists of the following three parts:

- 1. Technical Evaluation of CO₂ Capture Technologies for LNG Plants
- 2. Economic Evaluation of CO, Capture in LNG Plants
- 3. CO₂ Capture for Small Scale and Floating LNG Plants

The report should be cited in literature as follows:

'IEAGHG, "Techno-economic evaluation of CO₂ capture in LNG plants, 2019-07, October, 2019.'

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IEAGHG Technical Report

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TECHNO-ECONOMIC EVALUATION OF CO₂ CAPTURE IN LNG PLANTS

Key Messages

Natural gas (NG) is projected to play a vital role in the energy mix of the 21^{st} century. Its demand is forecasted to grow 2.5% a year for the next 10 years, ranking it second in the global energy mix in 2030. This study was commission to provide a technical evaluation and cost assessment of capturing CO₂ contained in produced natural gas and also CO₂ emitted by fuel combustions for power generation for LNG trains and refrigerant cycle compressors in liquefied natural gas (LNG) plants, including small scale (SSLNG) and floating (FLNG) plants. Prior to this study there was a lack of information on CO₂ capture in LNG plants. Consequently, the results of this study will be of direct interest to developers of LNG projects, the related capture technology, as well as vendors and policy makers.

- Although pre-combustion and oxyfuel options are available for capturing fuel related CO₂ emissions, post-combustion CO₂ capture, using well proven chemical absorption technology, will be the preferred route for baseload LNG (with a liquefaction capacity of 4.6 mtpa, 2mol% feed gas, state-of-the-art C3MR refrigeration process, proprietary amine CO₂ capture and located on the US Gulf Coast), as it can be installed without affecting the performance of the core process. This option reduces technical risks and process complexity.
- The cost of CO₂ captured for a baseload LNG plant (as described above) was estimated at €47.3/tCO₂ with the cost of CO₂ avoided at €55.2/tCO₂. The levelised cost of LNG for the baseload LNG plant without CCS is €1.18/MMBtu (or €54.5/tLNG), with CO₂ capture this cost increases by ~20% to €1.41/MMBtu (or €65.4/tLNG).
- The total range of cost encountered during the sensitivity analysis was €13 €57/tCO₂ for the capture cost and €14 €78/tCO₂ for the avoidance cost. A CO₂ emissions price of at least €129/tCO₂ would be required to make the base case LNG plant with CCS economically feasible.
- A CO₂ capture design that is incorporated into an exclusive acid gas removal unit (AGRU), instead of capturing the fuel related emissions as well, could bring down costs significantly to about €30/tCO₂. This figure is more in line with current CO₂ prices in certain countries, for example Norway and Finland, which indicates the potential for the implementation of CCS.
- Both SSLNG and FLNG plants have comparatively limited global capacity and therefore limited global CO₂ emissions. Global CO₂ emissions from SSLNG are an order of magnitude smaller in comparison with emissions from baseload LNG plants and three orders of magnitude smaller than global CO₂ emissions from power plants (8 10 mtpa vs 75 100 mtpa vs 10,000 mtpa). In addition, application of currently available CO₂ capture technologies face severe technical as well as economic challenges in these plants. Thus, efforts should focus on CCS in baseload LNG plants with capture capacity plants of around 3,000 t/CO₂ day equivalent to 1 mtpa.
- Large scale LNG trains (such as those found in Qatar with a capacity of 7.8 mtpa LNG) may provide greater benefits for CO₂ capture as a result of economies of scale. The total capture cost for plants this size is reduced by 12% with respect to the base case to about €41.6/tCO₂ and avoidance cost reduced to about €48.4/tCO₂.
- Recommendations for further work include:



- Pursuing general efforts to make CO₂ capture systems more efficient
- Assessment/development of other capture technologies as suitable for LNG
- Developing strategies to reduce compression power requirement
- Improving thermal efficiency of liquefaction process, e.g. through use of electric motor drives
- Developing exhaust gas recycle (EGR) technology with particular focus on gas turbines in LNG
- Demonstrating CCS in LNG on the fuel gas combustion processes

Background to the Study

Natural gas demand is forecasted to grow continuously for the next 10 years, playing a vital role in the global energy mix in 2030. In the specific case of liquefied natural gas (LNG), projections indicate a continued upward growth. The majority of near-term growth in liquefaction capacity is likely to happen in North America and Australia, although a number of other projects have the potential to add significant liquefaction capacity in the long term as well.

As a key contributor to the global energy supply, the LNG supply chain is expected to be subject to global requirements on reduction to greenhouse gas (GHG) emissions. While the majority of emissions are at the gas end user combustion processes, the LNG liquefaction plants produce a significant proportion of the total supply chain CO_2 emissions. The majority of the CO_2 emissions from LNG liquefaction plants arise from combustion of fuel for cycle compressors drive and power generation purposes. It is also necessary to remove and dispose of CO_2 from the natural gas feed, to avoid solidification in the liquefaction process.

The CO_2 emissions depend on the liquefaction plant configuration (e.g. feed gas pre-treatment, liquefaction technology, heat and power integration) and also local ambient conditions. The LNG industry has already improved the overall thermal efficiency of the LNG supply chain to reduce the associated CO_2 emissions. LNG plants may provide an early opportunity for CO_2 capture and storage (CCS) subject to the availability of high-concentration CO_2 streams and vicinity to storage reservoirs.

Based on this, IEAGHG considered it necessary to evaluate the techno-economics of CO_2 capture technologies in LNG plants in detail.

Scope of Work

IEAGHG commissioned Costain to develop a technical evaluation and cost assessment of CO₂ capture technologies for LNG plants including small scale (SSLNG) and floating (FLNG) plants.

The scope of work comprises three main tasks:

- 1) Technical evaluation of different CO2 capture technologies for LNG plants
- 2) Cost assessment of select CO2 capture technologies for LNG plants
- 3) Feasibility study of CO₂ capture technologies at SSLNG and FLNG plants

The contractor produced three separate reports for each of the above tasks.



Findings of the Study

Technical evaluation of CO₂ capture technologies for LNG plants

LNG baseload liquefaction plants use well established liquefaction technology, with a large proportion of these plants using a propane precooled mixed refrigerant process. Most of the plants use gas turbines to drive the refrigerant compressors, with recent plants using industrial Frame 5 and Frame 5 and their aeroderivative equivalents, resulting in typical train capacities in the order of 4-5 mtpa of LNG. A step change is seen with large trains having a capacity in the order of near 8 mpta using Frame 9 gas turbines. Electric motor drives are not widely used at the moment. They offer improvements in terms of overall efficiency and CO_2 emissions, so are expected to get more consideration in future designs. Figure 1 shows a typical block flow diagram of an LNG plant.



Figure 1 Typical block flow diagram of an LNG plant

 CO_2 emissions from combustion equipment depend on the thermodynamic efficiency of the processes, dictating the fuel gas that needs to be burned to produce the required power. This is a function of a number of variables including feed gas pressure and composition (dictating pre-treatment requirements), ambient conditions, liquefaction technology, gas turbine technology and process equipment. Liquefaction specific power requirement, considering technology and local factors, is typically in the range of 0.30 to 0.40 kWh/kg for baseload liquefaction technology. This leads to specific CO_2 emissions in the range of 0.20 to 0.28 tCO₂/tLNG for industrial gas turbines and 0.15 to 0.21 tCO₂/tLNG for aeroderivative gas turbines. Combined emissions, i.e. including from the acid gas removal unit (AGRU), increase this range to 0.30 to 0.40 tCO₂/tLNG for low CO₂ feed gas (2 mol%) and 0.70 to 0.90 tCO₂/tLNG for high CO₂ feed gas (14 mol%). Figure 2 shows the potential for CO₂ emissions captured and avoided for some representative LNG plants, while Figure 3 shows reported CO₂ emissions from existing LGN plants.





Figure 2 Potential of CO₂ captured and avoided for representative LNG plants



Figure 5. Reported CO₂ emissions from representative LNG plants (Ref. 50)

Figure 3 Reported CO₂ emissions (incl. power generation and other sources) from operating LNG plants

 CO_2 is also emitted by acid gas removal processes in the pre-treatment of feed gas prior to liquefaction. CO_2 from this process can be considered to be already captured. The separated CO_2 is typically vented, particularly if the content in feed gas is relatively low (i.e. 2 mol% or less). Where feed gas contains significant CO_2 content, considerations may need to be made for sequestration rather than venting in response to environmental constrains. The Snøhvit project is capturing and storing about 0.7 mtpa of



 CO_2 from the AGRU since 2008, with an LNG plant capacity of 4.3 mtpa and a feed gas content of 5-8 mol% CO_2 .

Based on the nature of the LNG industry, only well proven technologies are expected to be considered for CO_2 capture in LNG plants, with schemes that minimise the risks of disruptions to LNG production. Post-combustion CO_2 capture using well proven chemical absorption technology is likely to be the preferred route.

When compared to pre-combustion and oxyfuel combustion, post-combustion capture is considered to provide comparable performance with reduced technical risk and process complexity. Post-combustion can be installed without affecting the performance of the core liquefaction process. It requires a minimum number of modifications to existing equipment, hence reducing risk. This makes post-combustion attractive for new LNG plants or as retrofit to existing plants. However, space for the capture equipment is needed and this can be an issue, especially in SSLNG and FLNG plants (see related section later in this overview).

Table 1 summarises the advantages and disadvantages of the three CO_2 capture routes in the context of LNG plants.

Route	Advantages	Disadvantages
Post- combustion	Typically removes 85-90% of the CO_2	Flue gas is close to atmospheric pressure and the concentration of CO_2 is lower than in pre-combustion, so a more energy intensities characteristic required.
	Technology can be retrofitted with minimal modifications to liquefaction plant Applicable to any combustion system	Post-combustion CO_2 capture plants use more solvent and produce more solvent residue than pre-combustion plants
	Large experience for CO_2 capture applications in power generation (pilot and full scale commercial plants)	The major source of energy requirement is the use of heat to regenerate the solvent
	Downtime of CO_2 capture plant does not result in loss of LNG production if this can continue with untreated flue gas vented to atmosphere (current practice)	Absorption processes require large plot areas; large columns (both in height and cross sectional area), large solvent regeneration duties and cooling duties
Pre-	Typically removes 85-90% of the CO_2	Using H ₂ rich gas directly in large scale power production is a new application, involving complex gas turbine technology
combustion	Based on well-known and proven technologies used in chemicals and refinery industries to produce syngas	The fuel gas system producing H ₂ rich gas involves significant additional complexity and process equipment
	Gas turbine technology has been tested and developed to use hydrogen as fuel as an alternative to hydrocarbon fuel	Loss of H ₂ rich gas production would cause a loss of LNG production.
	Produced synthesis gas is rich in CO_2 and at high pressure, making separation of the CO_2 less energy intensive (using physical solvents) and the separation equipment more compact Requires less energy for CO_2 compression as some of the CO_2 is recovered at elevated pressure	Loss of flexibility in fuel gas system Economic feasibility may depend on economies of scale and
		the possibility to export $H_2\ rich\ gas\ produced\ in\ excess\ of\ the\ liquefaction\ plant\ requirements$
		Main issues are the integration, operability and reliability of plants
		Efficiency reduction as a result of energy losses due to shift conversion

Table 1 Advantages and disadvantages of different CO2 capture routes in LNG plants



		Oxyfuel combustion of natural gas in gas
Oxyfuel	Typically removes 90-97% of the CO ₂	turbines is under development and has not been tested on
on ji dol	- 2	a commercial scale
	Produces CO_2 at high purity from the combustion process	The fuel gas system and gas turbines involve significant
	No need for CO_2 capture process and solvent regeneration	additional complexity and process equipment
	Air separation plant for oxygen generation is a mature technology based on standard designs	Existing gas turbine drives may be difficult to retrofit, and feasibility may only be possible in the long term, but certainly involving a technical risk
	Oxyfuel boilers could be applied to generate steam used for power generation (including electric drive of cycle compressors) or heating (mainly solvent regeneration)	Efficiency losses are due to the electricity used by the oxygen production unit, with the cost of oxygen production in sufficient quantity representing a major operating cost
		Impact on thermal efficiency is potentially larger than post- combustion
		Loss of oxygen production would cause a loss of LNG production
		A higher volume of gas is fed to the CO_2 compressors due to the presence of impurities
		Purification of the CO_2 may be required to remove excess oxygen and impurities (nitrogen, argon)



The post-combustion technologies with the highest potential for immediate implementation are chemical absorption processes. These are proven technologies with the main disadvantage being the energy requirements to regenerate the solvent. However, there is potential for the heating duty to be provided by waste heat recovery, which in an LNG plant is expected to be available in excess. Perceived problems such as solvent degradation, solvent volatility and losses, corrosion, etc., represent operational challenges that can be managed within acceptable limits using solvent formulations that are commercially available. Table 2 contains a comparative assessment of all post-combustion capture technologies.

Technology		Advantages	Disadvantages	Applicable
Absorption	Chemical	 Applicable to selectively remove CO₂ from gas with low CO₂ content, at low pressure (near atmospheric) and delivering product with high purity 90% capture achievable at acceptable plant size and costs Mature technology commercially available, extensively tested for post-combustion CO₂ capture at pilot scale and implemented in large scale power generation plants, with capture plants having similar capacity to that expected for LNG plants Proprietary solvent formulations are commercially available with reduced regeneration energy demands, solvent degradation and corrosion LNG plants familiar with amine-based processes Availability of heat (from waste heat recovery) and power generation (from spare installed capacity) expected in LNG plants 	 Need for solvent handling (storage, pumping, reclaiming) Thermal regeneration of absorbent required Solvents can be subject to thermal degradation and in the presence of oxygen Relatively large equipment count and high energy requirements for regeneration of solvent Corrosion of carbon steel require use of stainless steel Waste produced from solvent reclamation needs disposal CO₂ product delivered at low pressure (near atmospheric) and water saturated, which requires dehydration and compression Applicable solvents are amine and ammonia based. Hot potassium carbonate and caustic wash excluded 	Yes (short term)
	Physical	 For high CO₂ partial pressures, solvent regeneration energy requirements for some physical solvents can be low in comparison to chemical absorption processes Use of ionic liquids with modified structure to suit CO₂ removal duty 	 Need for solvent handling Thermal regeneration of absorbent required Most suited to high partial pressure of CO₂ in feed gas above 3.5 bara. This would require compression of large flue gas volumes Low CO₂ loading capacity in low CO₂ partial pressure High viscosity of ionic liquids and lack of maturity and commercial availability 	No
	Hybrid	 Can potentially offer the benefits of both chemical solvents (higher CO₂ removal at low partial pressures) and physical solvents (lower energy requirements) 	 Thermal regeneration of absorbent required Need for solvent handling Typically more expensive than conventional amines CAPEX and OPEX tend to be higher than amine processes 	No
Adsorption		 Used for CO₂ removal in gas processing Widely available adsorbents, good thermal stability and low sensitivity to moisture Amine based adsorbents have lower heat of regeneration compared with aqueous amines (absorption) Fluidised bed processes (potassium carbonate) being tested at pilot scale for the specific application in post-combustion CO₂ capture 	 Regeneration of adsorbent required at low pressure (PSA) or high temperature (TSA) PSA would require compression of flue gases TSA requires handling of rich CO₂ regeneration gas Low CO₂ removal capacity, which makes them best suited to low CO₂ content and gas flowrates Not available commercially. It would require demonstration at full scale 	Yes (long term)

Table 2 Advantages of different post-combustion technologies in LNG plants



Technology	Advantages	Disadvantages	Applicable
Membranes	 Faster start-up, fewer equipment items and lower maintenance in comparison with solvent systems No regeneration, handling or disposal of solvents required CO₂ produced is not water saturated so no need for downstream dehydration Suited to plants in remote locations due to complexity in delivery of solvents and possibility of unmanned operation 	 Pre-treatment of the feed may be required to prevent damage of expensive membrane It would require compression of large flue gas volumes at first step High CO₂ recovery involves high recompression power requirements and large membrane areas (and associated footprint), increasing both the CAPEX and OPEX Best suited for bulk CO₂ removal to feed gas with high CO₂ concentrations of 10 mol% or more Membranes cannot produce high CO₂ purity on their own 	No
Cryogenic	 A liquid CO₂ stream at high pressure can be produced, reducing power requirements for CO₂ product compression Well known technology within other applications, particularly processing of gas with high CO₂ content 	 Need for pre-treatment to remove water and hydrocarbons that could freeze Process efficiency is low at low feed gas CO₂ concentrations, best suited to gas with >20 mol% CO₂ content Elevated feed gas pressures are required to improve performance High energy consumption for refrigeration system, provided as electricity Operating conditions and removal limitations to avoid CO₂ freezing 	No

The range of composition of flue gases produced by gas turbines can be considered to be in a relatively limited range. This is due to the large air excess volumes drawn into the gas turbines, with variability of fuel gas composition having a relatively minor impact on exhaust gas compositions. CO_2 content is in the order of 3 mol%, with oxygen being around 14 mol%.

For typical baseload liquefaction plants with capacities of around 5 mtpa of LNG, the required capacity of the CO_2 capture plant will be in the order of 3000 t CO_2 /day, equivalent to about 1 mtpa of CO_2 . Thus, relatively large flue gas flows need to be processed, see Table 3.

LNG Liquefaction Plant Capacity (mtpa of LNG)	CO ₂ Capture Plant Capacity (tonnes CO ₂ /day)	CO ₂ Capture Capacity (mtpa of CO ₂)
3.5	1900 - 2500	0.8 - 1.1
4.5	2400 - 3200	1.1 - 1.4
5.5	2900 - 3900	0.3 – 1.7
8.0	4500 - 5900	1.5 - 2.0

Table 3 CO₂ capture plant capacities for given LNG plant capacity

This is comparable to existing full-scale capture plants (e.g. Boundary Dam and Petra Nova), so a similar plant size and associated investment is expected. Means to reduce capital cost, strengthen the business case, minimise project risks and in general to improve the economics of the project will contribute to the feasibility of the CCS scheme.



Economic evaluation of CO₂ capture technologies in LNG plants

Description of base case

An economic assessment was done for the capture, compression, purification, transport and storage of CO_2 streams from two main sources of a typical baseload LNG plant:

- 1. CO₂ from the AGRU
- 2. CO_2 from combustion of fuel gas for producing power for the LNG process

The following Figure 4 shows the block flow diagram for the CO_2 capture plant of the base case.



Figure 4 Block flow diagram of the base case CO₂ capture plant

The following Table 4 summarises the main parameters of the base and the sensitivity cases.

Parameter	Base case	Sensitivity cases
Location	Gulf Coast, USA	Gulf Coast, USA
LNG capacity	4.6 mtpa	7.8 mtpa
Liquefaction technology	C3MR	AP-X
Refrigerant compressor drive	2x Frame 7	3x Frame 9
Specific power	0.3 kWh/kgLNG	0.3 kWh/kgLNG
Plant availability	95%	95%
Natural gas price	€6/GJ(LHV)	€3/GJ(LHV), €12/GJ(LHV)
Discount rate	8%	5%, 10%
CO ₂ transport and storage cost	€10/tCO ₂	€0/tCO ₂ , €20/tCO ₂

Table 4 Description of base case and sensitivity cases



Plant life	25 years	40 years
Steam production emissions	N/A	0.32 mtpa
Plant design	FOAK	NOAK
CO ₂ feed gas content	2 mol%	6 mol%, 14 mol%
CO ₂ sources captured	AGRU and combustion	AGRU only, CCS related emissions
CO ₂ capture process	Proprietary amine 90% capture rate 85% availability 3 GJ/tCO ₂ reboiler duty 0.2 kg/tCO ₂ make up rate €5/kg solvent cost 1 mtpa	Proprietary amine 90% capture rate 85% availability 3 GJ/tCO ₂ reboiler duty 0.2 kg/tCO ₂ make up rate €5/kg solvent cost 0.23 - 1.76 mtpa

The total specific costs of CO₂ were calculated as follows:

$$Total Specific Cost of CO_2 Captured = \frac{Total Capture Plant Lifetime Costs}{CO_2 Captured over Capture Plant Lifetime}$$
$$Total Specific Cost of CO_2 Avoided = \frac{Total Capture Plant Lifetime Costs}{CO_2 Avoided over Capture Plant Lifetime}$$

Results of base case assessment

Figure 5 shows the CO_2 emissions for the base case, with and without CCS. Applying CCS will lead to the capture of 1.06 mtpa and avoidance of 0.90 mtpa of CO_2 .



Figure 5 CO₂ emissions for base case with and without CCS

The Total Plant Cost (TCP) including contingency and contractor's fee is estimated as €597.5 million and the Total Capital Requirement (TCR) is approximately €704.6 million.



Total OPEX requirement (discounted Net Present Value (NPV)) for the project over the life of the plant (i.e. 25 years) is estimated as €567 million.

The specific cost of CO₂ captured for the base case is \notin 47.3/t CO₂. The main cost component for the base case is CAPEX which accounts for 56% of the total cost. OPEX accounts for the remaining cost; the key components are operation and maintenance, power generation and CO₂ transport and storage. The cost of CO₂ avoided for the base case is \notin 55.2/t CO₂.

The levelised cost of LNG for the base case is €65.4/tLNG (€1.41/MMBtu) whereas the estimated levelised cost of LNG for an LNG plant without capture is €54.5/tLNG (€1.18/MMBtu). These costs are for the production of LNG only and do not include costs incurred for transport, regasification, etc. The minimum cost of CO₂ emissions would have to be at €129/tCO₂ to justify the installation of CCS (based on CO₂ emissions cost against CO₂ capture costs), see Figure 6.



Figure 6 Levelised cost of LNG vs CO₂ emissions cost

Sensitivity analysis

The sensitivity analyses show a total range in capture costs of $\notin 13/tCO_2$ to $\notin 57/tCO_2$ and avoidance costs of $\notin 14/tCO_2$ to $\notin 78/tCO_2$. The lowest specific cost applies to CO_2 capture of AGRU emissions only, with a high feed gas CO_2 content (14 mol%).

Natural gas price impacts specific cost due to the requirement for natural gas to produce power which constitutes a large portion of the total cost (see above). Varying natural gas price to extremes of €3/GJ



(lower heating value (LHV)) and $\notin 12/\text{GJ}$ (LHV) result in capture costs of about $\notin 44.5/t\text{CO}_2$ and $\notin 52.8/t\text{CO}_2$ and avoidance costs of $\notin 52.0/t\text{CO}_2$ and $\notin 61.7/t\text{CO}_2$ respectively. This represents an increase of 12% for the high NG price case and a decrease of 6% for the low price case. Steam production is assumed to be provided by waste heat recovery. While there may be costs associated with adapting the existing facilities for the CCS scheme, the cost of steam production is assumed to be negligible.

Lower discount rates result in 22% higher specific cost while increasing the discount rate has the opposite effect, a reduction of 9%. At discount rates of 5% and 10% the capture costs are about $\notin 57.5/tCO_2$ and $\notin 43.0/tCO_2$, avoidance costs are $\notin 67.2/tCO_2$ and $\notin 50.3/tCO_2$ respectively.

Adjusting the cost for CO₂ transport and storage affects the total specific cost, with higher costs for transport and storage resulting in 9% higher total costs, while in case of no transport and storage cost the cost are 9% lower. At CO₂ transport and storage costs of $\notin 0/t$ CO₂ and $\notin 20/t$ CO₂ the capture costs are about $\notin 42.8/t$ CO₂ and $\notin 51.7/t$ CO₂, avoidance costs are $\notin 50.0/t$ CO₂ and $\notin 60.5/t$ CO₂ respectively. All cases do not include for the CAPEX requirement for the infrastructure associated to CO₂ transport and storage.

Increasing plant life to 40 years reduces capture costs by about 32% with respect to the base case to about \notin 32.1/tCO₂ with avoided cost reduced to about \notin 37.5/tCO₂. This assumes that there are no additional significant costs associated with the design, construction or operation of the plant as a result of the extended plant life.

It is assumed in the base case that there will be no running costs associated with the production of steam since the excess heating is expected to be available in LNG facilities through waste heat recovery. For a case where additional energy is needed and thus additional costs are incurred, additional emissions associated with 0.15 mtpa for power generation and 0.17 mtpa for steam generation are assumed. Capturing those emissions increases capture cost by 13% to $€53.6/tCO_2$ and avoidance cost by 41% to $€77.6/tCO_2$.

Large scale LNG trains (such as those found in Qatar with capacity of 7.8 mtpa LNG) may provide greater benefits for CO₂ capture as a result of economies of scale. The total capture cost for plants this size is reduced by 12% with respect to the base case to about \notin 41.6/tCO₂ and avoidance cost reduced to about \notin 48.4/tCO₂.

Capturing the emissions associated with CCS (i.e. burning fuel to provide power to the capture, compression and purification processes) results in a 2% reduction in specific capture costs to about \notin 46.5/tCO₂, and avoidance cost of \notin 54.5/tCO₂. The selection of this option may be dictated by increased complexity in the design and operation of the plant.

NOAK (nth of a kind) plant designs increase attractiveness of CCS due to the ability to reduce CAPEX costs compared to FOAK (1st of a kind) design. An expected CAPEX reduction of up to 30% is generally achievable, resulting in a reduction of capture cost to about \notin 36.6/tCO₂, and avoidance cost to \notin 42.8/tCO₂.

The base case considers feed gas with 2 mol% CO₂ content. For a higher feed gas concentration of 14 mol%, the capture cost reduces by 46% to \notin 25.6/tCO₂ and the avoidance cost to \notin 27.3/tCO₂. CAPEX, OPEX and specific cost of capture are all significantly reduced when the CCS scheme only considers capture of the CO₂ that is separated from the feed gas in the AGRU. Three different cases with varying



feed gases for CO_2 capture from the AGRU only were investigated (2 mol%, 6 mol% and 14 mol%). The costs are decreasing as follows:

- 2 mol%: capture cost is $\in 16.5/tCO_2$ and avoidance cost is $\in 17.9/tCO_2$
- 6 mol%: capture cost is $\in 13.0/tCO_2$ and avoidance cost is $\in 14.1/tCO_2$
- 14 mol%: capture cost is €12.5/tCO₂ and avoidance cost is €13.6/tCO₂

These results show the potential for implementation of CCS schemes could be limited to capture of CO_2 from the feed gas over post-combustion schemes that also include combustion related emissions to power the LNG plant.

The CO₂ emission price (i.e. CO₂ tax) required to justify implementation of post-combustion CCS schemes in LNG plants purely on economics, i.e. reduction of costs paid for emitted CO₂, is in the order of $\notin 100/t$ CO₂. Current world emission policies set CO₂ tax at a relatively low value (if any), with the majority of emissions currently priced at less than about $\notin 10/t$ CO₂. This indicates that implementation of post-combustion CCS would only occur for either significant CO₂ tax increases or by drivers other than plant economics. When the CCS scheme only considers capture of the CO₂ that is separated from the feed gas in the AGRU, the minimum CO₂ tax required is in the order of $\notin 30/t$ CO₂. This level of CO₂ tax is within current environmental policies in some regions (e.g. Norway and Finland), which indicates the potential for the implementation of CCS.

Figure 7 and Figure 8 summarise the CO_2 capture cost and CO_2 avoidance cost for all investigated sensitivity cases in comparison to the base case.



Figure 7 Summary of CO2 capture cost for base case and sensitivity analysis





Figure 8 Summary of CO₂ avoidance cost for base case and sensitivity analysis



CO₂ capture for small scale (SSLNG) and floating (FLNG) plants

Small-scale LNG (SSLNG)

SSLNG plant usually have a capacity in the order of 1 mtpa LNG and below. The global installed production capacity of SSLNG is of the order of 20 mtpa distributed across more than 100 plants. SSLNG Global capacity is expected to increase towards 30 mtpa by 2020. This constitutes only a fraction of the total global capacity from baseload LNG plants, which currently stands at over 300 mtpa distributed in about 100 liquefaction trains in over 40 export terminals.

SSLNG plants use single mixed refrigerant or nitrogen expander cycle liquefaction technologies, both are expected to have very similar CAPEX with main differences in OPEX and plant operability. Thermal efficiencies are smaller than those achieved by baseload liquefaction processes, but this is generally less relevant to the overall economics of SSLNG plants.

As per baseload LNG plants, it is likely that post-combustion capture would remain the most viable CCS route for SSLNG plants. However, post-combustion capture using solvents requires large plot areas, large column heights and high energy requirements all of which are likely to be unattractive for SSLNG unless sufficient plots of land and provision of power and heating are available. On the other hand, CCS in SSLNG plants might provide a gateway for demonstration of capture technologies for LNG plants in general.

CCS schemes on SSLNG plants would not benefit from economies of scale. Therefore, the specific cost (\notin/tCO_2) of a capture plant on a SSLNG plant will be higher than the cost of a capture plant on a baseload LNG plant, see Table 5. This will make financing and economic feasibility of SSLNG plants with CCS more challenging than baseload LNG plants with CCS. Although the CO₂ capture plant capacity for SSLNG is only about 30% of that for a baseload LNG, the CAPEX is still considerable.

		Baseload LNG	Small
			Scale LNG
LNG Plant Capacity	mtpa	4.6	0.8
Specific CO ₂ Emissions	tCO ₂ /tLNG	0.24	0.36
CO ₂ Capture Plant Capacity	mtpa	1.0	0.3
Total Capital Requirement	€ million	755	300-500

Table 5 CO₂ emissions and cost comparison for SSLNG

SSLNG plants located in geographically stranded regions are likely to experience difficulty in finding a suitable destination for captured CO_2 unless transport infrastructure (i.e. CO_2 pipelines) and suitable geological storage or usage application exist.

Simplicity, availability, reliability and quick project implementation are key selling points for SSLNG plants, therefore it is likely that the introduction of a CCS scheme, which adversely affects these attributes, would be unfavourable.

SSLNG plants are not likely to be a main priority for implementation of CCS schemes as a means to achieve global decarbonisation targets, at least in the short or medium term. The key reason is that the



global capacity of SSLNG plants is low and as such, associated net emissions from the key emission sources are small and distributed across a large number of plants. Global CO_2 emissions from SSLNG are an order of magnitude smaller in comparison with emissions from baseload LNG plants and three orders of magnitude smaller than global CO_2 emissions from power plants (8-10 mtpa vs 75-100 mtpa vs 10000 mtpa).

It is likely that other CO_2 abatement strategies will be the focus of attention for SSLNG plants before the decision is made to implement CCS schemes.

Floating LNG (FLNG)

Typical capacities of FLNG plants are about 1 mtpa per train, with capacities of current projects ranging from 0.5 to 3.6 mtpa. 24 FLNG proposals have been announced with a total capacity of 157 mtpa, and four FLNG projects have been sanctioned and moved to the construction phase, with a total liquefaction capacity of 8.7 mtpa (see Table 6). Thus, the FLNG industry is still at an early stage of development, with uncertainties regarding technical and economic feasibility at production scale.

	Project	Start	Capacity	Liquefaction	Refrigerant
Country		Year	(mtpa)	Technology	Compressor
Australia	Prelude FLNG	2018	3.6	Shell DMR	Steam
Cameroon	Cameroon FLNG	2017	2.4	B&V PRICO (SMR)	PGT25+G4
Malaysia	PFLNG Satu	2017	1.2	Air Products AP-N	PGT25+G4
Malaysia	PFLNG 2	2020	1.5	(dual N ₂ expander)	LM6000-PF+

Table 6 Sanctioned FLNG projects

Dual mixed refrigerant (DMR) and nitrogen expander cycle processes have generally been evaluated and selected in FLNG projects. DMR has been selected due to its high efficiency (comparable to baseload LNG technologies) and large production capacity in a single train (as dictated by reserves and project economics), whilst nitrogen expander cycle processes have been selected based on safety considerations (elimination of flammable refrigerants) and insensitivity to FLNG vessel motion.

Post-combustion capture based on chemical absorption has been deemed the most appropriate for onshore baseload plants. However, it is unlikely to be feasible for FLNG plants.

- Chemical absorption based capture has a significant footprint for the treatment of large quantities of low pressure flue gas. For a FLNG plant, there are space and weight limitations, optimised for the gas pre-treatment, liquefaction and utilities. There is no spare space for further topsides equipment, which makes integration unfeasible.
- Additional power is likely to be needed to meet the requirements of the capture plant which will impact FLNG vessel design further in terms of space availability, layout and weight constraints, etc.
- The options to have either a dedicated floating capture plant or an onshore capture plant treating flue gas for a FLNG unit located at shore appears to be unfeasible, too, particularly due to the interfaces required.

Oxyfuel and pre-combustion capture do not offer prospects in term of feasibility for FLNG plants either.



Capture of reservoir CO_2 separated in the pre-treatment process appears to be the only feasible CCS scheme for FLNG plants. The scheme may still pose significant technical challenges for design and operation of FLNG plants, and will impact the overall project economics.

- Fields with high CO₂ content will provide the greatest potential for CCS. However, from the project perspective, it is likely that fields with a low CO₂ content would be selected for development with a FLNG scheme, which reduces pre-treatment duties, size and associated costs.
- Transfer of CO₂ to a suitable storage site imposes significant challenges and has financial and environmental implications. Introducing the requirement to inject CO₂ will increase complexity (e.g. interfaces with subsea systems) and cost of design. Storage located far from the FLNG plant will require extensive transport infrastructure (pipelines etc.) which will add complexity, cost and environmental risks.

As per SSLNG plants, FLNG plants are not likely to be a main priority for implementation of CCS schemes as means to achieve global decarbonisation targets due to the small contribution to global emissions. It is likely that other CO_2 abatement strategies (e.g. improving the process efficiency and reducing the specific emissions of gas turbines) will be the focus of attention.

Expert Review

Three experts from the LNG industry took part in the review of the draft report and provided detailed comments on the techno-economic assessment.

Regarding the technical assessment of CO_2 capture in LNG, the reviewers stated their agreement with the selection of post-combustion capture for fuel related emission and the current non-feasibility of oxyfuel combustion. One reviewer mentioned pre-combustion would not be applied in LNG but outside the LNG boundary (if at all), which was added to the related section. The inclusion of inlet facilities and condensate stabilisation as potential CO_2 sources was also suggested. Consequently the related section was expanded to include emissions associated with heating of liquids in distillation processes. To be more precise, the term syngas was replaced with H₂ rich gas throughout the report where applicable. Comments on the overview table that is showing the advantages and disadvantages of the three capture options in LNG were addressed, too. A request to name the technology licensors for the non-liquefaction technologies was not implemented in order to maintain commercial neutrality.

There were several comments by all reviewers on the economic assessment, some of which have been used by the contractor to update the cost numbers where required. As the base case results are based on typical data and assumptions, one suggestion was to change it to a more detailed simulation of a specific LNG plant. This was not changed, as the base case definition was intended to be representative, rather than very specific. One reviewer noted, the overall investment cost seemed high. However, another reviewer found they were reasonable. Total plant cost estimated in this study (i.e. ϵ 600m) is in line with those of other CCS projects (e.g. ϵ 640m for Peterhead and ϵ \$800m for Boundary Dam). One reviewer questioned the selection of 85% availability for the AGRU capture as too low. 85% was used as per standard design specification for IEAGHG TEAs, the reason being that most FOAK projects usually show a lower availability during their first year(s). In addition to the updated cost numbers, definitions and formulas for calculating cost of CO₂ captured and avoided were added or presented more clearly. Overall, the reviewers positively acknowledged the number and variety of sensitivity cases in the TEA, especially the sensitivity for different CO₂ feed gas content.



Comments for the assessment of CO_2 capture in SSLNG and FLNG were also received. Clarification and information was added regarding the construction and production schedule for FLNGs. The sections about Prelude FLNG were updated with more information and clarifications and a disclaimer was added to note that CO_2 emissions have been allocated to the LNG production, ignoring LPG and condensate.

Conclusions

It was the aim of this study to develop a techno-economic evaluation and cost assessment of CO_2 capture technologies for LNG plants including small scale (SSLNG) and floating (FLNG) plants.

LNG baseload liquefaction plant use well established liquefaction technology, with the majority relying on a propane precooled mixed refrigerant process. The typical train capacities are in the range of 4 -5 mtpa LNG. CO_2 emissions from combustion equipment depend on a number of variables, e.g. feed gas pressure and composition, ambient conditions, liquefaction and gas turbine technology, process equipment, and are typically in the range of 0.20 - 0.28 tCO₂/tLNG for industrial gas turbines and 0.15 – 0.21 tCO₂/tLNG for aeroderivative gas turbines. Post-combustion CO₂ capture using well proven chemical absorption technology will likely be the preferred route for baseload LNG, as it can be installed without affecting the performance of the core process, thus reduces technical risks and process complexity. Chemical absorption processes tend to have high energy requirements for solvent regeneration, however, in an LNG plant waste heat is usually available in excess. For a typical baseload LNG with a capacity of around 5 mtpa, the required CO₂ capture plant will be in the order of 3000 tCO₂/d, or 1 mtpa, which is comparable to existing full scale capture plants.

The cost of CO₂ captured for a baseload LNG plant of 4.6 mtpa and 0.24 tCO₂/tLNG emissions intensity is \notin 47.3/tCO₂ and the cost of CO₂ avoided is \notin 55.2/tCO₂. The levelised cost of LNG for the baseload LNG plant without CCS is \notin 54.4/tLNG (\notin 1.18/MMBtu), with CO₂ capture increasing this cost to \notin 65.4/tLNG (\notin 1.41/MMBtu). There are several sensitivities that can affect the costs of CO₂ captured and avoided and the total cost range encountered during the analysis is \notin 12.5 – \notin 57.5/tCO₂ captured and \notin 13.6 – \notin 77.6/tCO₂ avoided. Cost can be lowered by, e.g. decreasing NG prices, lower CO₂ transport and storage costs, increased plant life, larger scale LNG trains, NOAK plant designs, and capturing CO₂ from feed gas in the AGRU only. The CO₂ price required to justify CCS implementation in LNG plants is about \notin 129/tCO₂. A design only considering CO₂ from the AGRU could bring this down to about \notin 30/tCO₂. The latter figure is more in line with current CO₂ prices in certain regions.

For SSLNG plants, post-combustion capture will likely be the preferred options as well. However, such capture plants might be unattractive, due to the large area requirements. The lack of economies of scale will result in a higher cost of CO₂ captured/avoided than in a baseload LNG plant. An SSLNG with 0.8 mtpa has specific CO₂ emissions of 0.36 tCO₂/tLNG (i.e. about 50% higher than a typical baseload plant) and could require up to \notin 500 million of total capital, compared to \notin 755 million of the reference baseload plant. However, for SSLNGs, capture and storage of CO₂ from the AGRU might be a feasible option.

Similar to SSLNGs, implementation of CCS in FLNGs faces comparable, if not more, issues. Safe and robust operation generally takes priority over efficiency. Only four FLNG projects are currently in the construction phase. Thus, it is likely that other CO_2 reduction approaches, such as improving process efficiency and reducing emissions of gas turbines, will be the focus of attention. Both SSLNG and



FLNG plants have in common, that their global capacity and related CO₂ emissions are low, so efforts should focus on CCS in baseload LNG plants.

Recommendations

Based on the results and conclusions of the assessment it is suggested that future work should focus on:

- Pursue general efforts to make CO₂ capture systems more efficient
- Assessment of other capture technologies that are suitable for LNG
- Develop strategies to reduce compression power requirement
- Improve thermal efficiency of liquefaction process, e.g. through use of electric motor drives
- Develop exhaust gas recycle (EGR) technology with particular focus on gas turbines in LNG
- Demonstrate CCS in LNG on the fuel gas combustion processes

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COSTAIN

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1.0 ABBREVIATIONS

AGRU	Acid Gas Recovery Unit
ALNG	Atlantic LNG
aMDEA	Activated Methyl Diethanol Amine
APCI	Air Products and Chemicals Inc.
APLNG	Australia Pacific LNG
Ar	Argon
bara	Bar (atmospheric)
BOG	Boil Off Gas
C3MR	Propane Precooled Mixed Refrigerant
CAD	Canadian Dollar
CAPEX	Capital Expenditure
CCGT	Combined Cycle Gas Turbine
CCS	Carbon Capture and Storage
CHP	Combined Heat and Power
CO	Carbon Monoxide
CO ₂	Carbon Dioxide
COS	Carbonyl Sulphide
CS_2	Carbon Disulphide
CWHE	Coil Wound Heat Exchanger
DEA	Diethylene Glycol
DIPA	Di-isopropanol Amine
DLE	Dry Low Emissions
DMR	Dual Mixed Refrigerant
EFG	End Flash Gas
EGR	Exhaust Gas Recycle
EGLNG	Equatorial Guinea LNG
ELNG	Egypt LNG
EOR	Enhanced Oil Recovery
FGR	Flue Gas Recycle
FLNG	Floating LNG
GE	General Electric
GHG	Greenhouse Gas
GJ	Gigajoule
GLNG	Gladstone LNG
GPSA	Gas Processors Suppliers Association
GT	Gas Turbine
H ₂ O	Water
H_2S	Hydrogen Sulphide
HHV	Higher Heating Value
HP	High Pressure
HRSG	Heat Recovery Steam Generator
IEA	International Energy Agency
IGCC	Integrated Gasification Combined Cycle
ISO	International Organization for Standardization
K_2CO_3	Potassium Carbonate
kg	Kilogram

KHCO₃	Potassium Bicarbonate
kW	Kilowatt
kWh	Kilowatt-hour
LHV	Lower Heating Value
LNG	Liquefied Natural Gas
LP	Low Pressure
LPG	Liquefied Petroleum Gas
MDEA	Methyl Di-ethanol Amine
MEA	Mono-ethanol Amine
MFC	Mixed Fluid Cascade
MJ	Megajoule
MLNG	Malaysian LNG
MMSCFD	Million Standard Cubic Feet per Day
mol	Molar
MP	Medium Pressure
MR	Mixed Refrigerant
mtpa	Million Tonnes Per Annum
MW	Megawatt
MWh	Megawatt-hour
N ₂	Nitrogen
NGL	Natural Gas Liquids
NO	Nitric Oxide
NO ₂	Nitrogen Dioxide
NOx	Nitrogen Oxides
O ₂	Oxygen
OPEX	Operational Expenditure
PFHE	Plate Fin Heat Exchanger
PNG	Papua New Guinea
ppmv	Parts Per Million (volume)
PSA	Pressure Swing Adsorption
QCLNG	Queensland Curtis LNG
R&D	Research and Development
RH	Relative Humidity
ROAD	Rotterdam Opslag en Afvang Demonstraieproject
RSH	Mercaptans
SO ₂	Sulphur Dioxide
SO ₃	Sulphur Trioxide
SOx	Sulphur Oxides
SRD	Specific Reboiler Duty
TEA	Triethanolamine
tpa	Tonnes Per Annum
TSA	Temperature Swing Adsorption
UK	United Kingdom
US	United States (of America)
VSA	Vacuum Swing Adsorption
WHRU	Waste Heat Recovery Unit

wt Weight (mass)

COSTAIN

2.0 EXECUTIVE SUMMARY

IEAGHG commissioned Costain to develop a techno-economic evaluation and cost assessment of different CO_2 capture technologies for LNG plants. This report summarises the technical investigations and evaluations developed for the technical evaluation of different CO_2 capture technologies for LNG plants. A summary of the key findings of this report is given below:

- LNG baseload liquefaction plants use well established liquefaction technology, with a large proportion of these plants using a propane precooled mixed refrigerant process. Most of the plants use gas turbines to drive the refrigerant compressors, with recent plants using industrial Frame 5 (and their aeroderivative equivalents) and Frame 7 resulting in typical train capacities in the order of 4-5 mtpa of LNG. Large trains with a capacity near 8 mpta use Frame 9 gas turbines.
- CO₂ emissions from combustion equipment depend on the thermodynamic efficiency of the processes, dictating the fuel gas needed to produce the required power. Liquefaction specific power considering technology and local factors is typically in the range 0.3 to 0.4 kWh/kg. This leads to specific CO₂ emissions in the range 0.20 to 0.28 tonne of CO₂ per tonne of LNG for industrial gas turbines and 0.15 to 0.21 tonne of CO₂ per tonne of LNG for aeroderivative gas turbines
- Based on the nature of the LNG industry, only well proven technologies are expected to be considered for CO₂ capture in LNG plants, with schemes that minimise the risks of disruptions to LNG production. Post-combustion CO₂ capture using well proven chemical absorption technology is likely to be the preferred route.
- When compared to pre-combustion and oxyfuel combustion, post-combustion capture is considered to provide comparable performance with reduced technical risk and process complexity. Post-combustion can be installed without affecting the performance of the core liquefaction process. It requires a minimum number of modifications to existing equipment, hence reducing risk. This makes post-combustion attractive for new LNG plants or as retrofit to existing plants.
- The post-combustion technologies with the highest potential for immediate implementation are chemical absorption processes. These are proven technologies with the main disadvantage being the energy requirements to regenerate the solvent. However, there is potential for the heating duty to be provided by waste heat recovery, which in an LNG plant is expected to be available in excess. Perceived problems such as solvent degradation, solvent volatility and losses, corrosion, etc., represent operational challenges that can be managed within acceptable limits using solvent formulations commercially available.
- The range of composition of flue gases produced by gas turbines can be considered to be in a relatively small range. This is due to the large air excess volumes drawn into the gas turbines, with variability of fuel gas composition having a relatively minor impact. CO₂ content is in the order of 3 mol%, with oxygen around 14 mol%.
- For typical baseload liquefaction plants with capacities of around 5 mtpa of LNG, the required capacity of the CO₂ capture plant will be in the order of 3000 tonnes of CO₂ per day, equivalent to about 1 mtpa of CO₂. This is comparable to existing full-scale capture plants (e.g. Boundary Dam and Petra Nova), so a similar plant size and associated investment is expected.

- CO₂ capture remains an energy-intensive process despite incremental performance improvements being delivered by using proprietary capture technologies. LNG plants have a significant potential in this regard as there is usually scope for additional waste heat recovery at the exhaust of gas turbines in mechanical drive or power generation service that would provide the required process heating at minimal cost (assuming waste heat recovery is installed).
- Sequestration of CO₂ vented in the AGRU will play a major role in the implementation of CCS in LNG plants in consideration of the higher potential in terms of technical feasibility, footprint, cost and impact on overall project feasibility particularly on financing, compared to post-combustion capture. Project costs (excluding transport and storage infrastructure) are one order of magnitude lower than the full scale postcombustion capture costs.
- New build plants are expected to have greater potential for CCS due to the ability to optimise the design to facilitate the incorporation of the CCS scheme. Capture-ready plants, while allowing for reduction in costs, will also encourage implementation of CCS by providing design allowances (tie-ins, allocated plot space, spare power generation capacity) to facilitate for installation of future CO₂ capture plants. Implementation of CCS schemes as retrofit on existing (non-capture ready) plants appears to be difficult due to technical challenges and the impact on the LNG production economics.
- The technical and commercial feasibility leading to successful implementation of a CCS scheme is expected to be considered early when the economics of the LNG production scheme are developed. The potential implementation of CCS in LNG plants could be realised by a phased development, with sequestration of CO₂ vented from the AGRU being a precursor for full scale CCS to make financing feasible and to manage technical and commercial risks. Post-combustion CCS would be then implemented via capture-ready plant designs.

3.0 BACKGROUND

Natural gas demand is forecasted to grow continuously for the next 10 years, playing a vital role in the global energy mix in 2030. In the particular case of Liquefied Natural Gas (LNG), projections indicate a continued considerable growth. The majority of near-term growth in liquefaction capacity is likely to happen in North America and Australia, although a number of other projects have the potential to add significant liquefaction capacity in the long term as well.

As a key contributor to the global energy supply, the LNG supply chain is expected to be subject to global requirements on reduction to greenhouse gas (GHG) emissions. While the majority of emissions are at the gas end user combustion processes, the LNG liquefaction plants produce a significant proportion of the total supply chain CO_2 emissions. The majority of the CO_2 emissions from LNG liquefaction plants arise from combustion of fuel for cycle compressors drive and power generation purposes. It is also necessary to remove and dispose of CO_2 from the natural gas feed, to avoid solidification in the liquefaction process.

The CO_2 emissions depend on the liquefaction plant configuration (e.g. feed gas pretreatment, liquefaction technology, heat and power integration) and also local ambient conditions. The LNG industry has already improved the overall thermal efficiency of the LNG supply chain to reduce the associated CO_2 emissions. LNG plants may provide an early opportunity for CO_2 capture and storage (CCS) subject to the availability of highconcentration CO_2 streams and vicinity to storage reservoirs.

Based on this, the International Energy Agency (IEA) GHG programme considered necessary to evaluate the techno-economics of different CO₂ capture technologies in LNG plants in detail.

4.0 SCOPE

IEAGHG commissioned Costain to develop a techno-economic evaluation and cost assessment of different CO₂ capture technologies for LNG plants including for small scale and floating (FLNG) plants.

The scope of work comprises three main tasks:

- 1) Technical evaluation of different CO₂ capture technologies for LNG plants
- 2) Cost assessment of different CO₂ capture technologies for LNG plants
- 3) Feasibility study of different CO₂ capture technologies at small scale and floating (FLNG) Plants

This report summarises the technical investigations and evaluations developed for Task 1.



5.0 LIQUEFACTION PROCESSES AND PLANTS

5.1 GENERIC PROCESS SCHEME

An overview of the scope of facilities included in this study is presented below. This is based on a generic scheme representative of LNG liquefaction baseload plants (export terminals). Identification of sources of CO_2 emissions will be based on this scope of facilities.

5.1.1 Inlet Facilities

Gas received from the gas field is processed in the usual manner as any other onshore gas processing facility, with bulk condensate liquids separated from the gas in slug catchers and two-phase separators. Condensate is routed to the Condensate Stabilisation unit.

5.1.2 Pre-Treatment

The natural gas feed is pre-treated in a series of process units to remove CO_2 (in the AGRU) and water (in a molecular sieve dehydration unit) that would otherwise freeze in the downstream liquefaction unit and cause blockages. H_2S is also removed with CO_2 mainly to meet LNG product and fuel gas specifications, also reducing risk of corrosion.

Where aluminium is used as a material of construction for equipment items such as brazed aluminium or Coil-Wound heat exchangers, mercury removal units (solid bed adsorption) are used to avoid damage to the heat exchangers due to mercury deposition leading to loss of structural integrity due to metal embrittlement.

5.1.3 Feed Gas Cooling

The treated feed gas is pre-cooled and partially condensed in the front-end part of the liquefaction process in the first refrigerant cycle.

5.1.4 Heavy Hydrocarbons Removal

The cooled feed gas is passed to a refluxed scrub column to separate valuable natural gas liquids (NGL) and to remove heavy components such as benzene and aromatics that may freeze up in the downstream cryogenic exchangers. The liquid stream from the scrub column bottoms is passed to the fractionation unit to provide liquefied petroleum gas (LPG), NGL, refrigerant make-up and fuel.

5.1.5 Liquefaction, Sub-cooling and End Flash

The liquefaction process consists of one or a series of refrigerant cycles, with refrigerant compressors driven either by gas turbines or electric motors. The feed gas is liquefied and subcooled against the evaporating refrigerants. The LNG is then flashed to near atmospheric pressure for atmospheric storage in cryogenic tanks at around -160°C. The flash gas generated is recompressed and used for fuel. Nitrogen rejection can be integrated in the end flash gas process depending on the fuel gas specification requirements, with removed high purity nitrogen vented to atmosphere. A portion of the available cycle power is required to provide sufficient sub-cooling of the LNG to ensure that the quantity of flash gas generated meets the plant fuel gas requirements. To further enhance LNG production, hydraulic expansion turbines may be used.



5.1.6 Storage and Export

LNG is stored in cryogenic tanks at a pressure slightly above atmospheric. LNG is pumped into LNG carriers and transferred using loading arms. Boil Off Gas (BOG) generated during this process as a result of overall heat–in leak is recompressed and used as plant fuel.

5.1.7 Fractionation / Condensate Stabilisation

The fractionation unit consists of a number of columns and separators designed to separate the scrub column bottoms into ethane, propane, butane and stabilised condensate, which may be used as refrigerant make-up and/or exported as separate products. Excess ethane, propane and butane may be re-injected into the natural gas stream in the main cryogenic heat exchanger to increase LNG production and calorific value if required.

Where stabilisation of condensate from various sources is required, this is achieved in a reboiled column.

5.1.8 Power and Heat Generation

Power generation is required where this is not imported from the local grid to meet the electrical demand of electric motors driving pumps, fans, gas compressors (including liquefaction cycle refrigerant compressors if these are fitted with electric motors), control systems, electric heating, etc.

The main heating duty is required for regeneration of the solvent in the acid gas recovery unit (AGRU), regeneration of adsorber beds media and reboiling duties in the fractionation and condensate stabilisation units. The power and heating demand can be provided in an efficient manner by a combined cycle gas turbine (CCGT). Additional heat may be produced from waste heat recovery installed in gas turbines. Depending on the heat balance and recovery, additional heating requirements can be provided by steam boilers (for reboiling duties) or fired heaters (for adsorber media regeneration gas heating).



Figure 1. Typical Block Flow Diagram of LNG Liquefaction Plant



5.2 LIQUEFACTION TECHNOLOGIES AND LNG PLANTS

The main liquefaction technologies applicable to baseload LNG liquefaction plants are:

- Propane Pre-cooled Mixed Refrigerant (AP-C3MR[™]) licenced by Air Products and Chemicals Inc. (APCI), is the leading liquefaction process accounting for approximately 87% of the world's baseload LNG capacity
- AP-X[®] licenced by APCI is essentially the AP-C3MR[™] process in combination with a nitrogen expansion cycle and has been developed in response to demand for increased LNG train capacity and lower unit cost
- **Optimized Cascade**[®] licensed by ConocoPhillips uses three pure refrigerant cycles i.e. propane, ethylene and methane, that are optimally cascaded in sequence
- **Mixed Fluid Cascade (MFC[®])** licensed by Linde is essentially a three-cycle mixed refrigerant cascade process using a combination of PFHEs and CWHEs
- **Dual Mixed Refrigerant (DMR)** licensed by Shell is a development of the AP-C3MR[™] process but uses two separate mixed refrigerant cycles which are cascaded to achieve a better match to the natural gas cooling curve
- Liquefin currently licensed by Air Liquide consists of two mixed refrigerant cycles, one for pre-cooling (using ethane and propane) and the other for liquefaction and sub-cooling (using a lighter blend of methane, ethane, propane and nitrogen).

See Appendix 1 for a more detailed overview of these liquefaction technologies.

The vast majority of plants in operation use the AP-C3MR[™] process, with the global average capacity for an LNG liquefaction train of approximately 3.8 mtpa. LNG production capacity per train peaks at 7.8 mtpa in Qatar which was brought online in 2009 using the AP-X[®] process.

As of January 2017 the total global LNG production capacity stood at about 340 mtpa (Ref 5). A number of projects started commercial operations during 2016 increasing the total global LNG production capacity by approximately 35 mtpa. A further 115 mtpa is scheduled to be online over the next few years. APCI's market share of liquefaction technologies is set to remain relatively unaffected based upon the plants currently in construction with over half opting for the APCI technology.

See Appendix 2 for further details on LNG baseload plants.

5.3 REFRIGERATION CYCLE COMPRESSOR DRIVES

The selection and configuration of the driver for the refrigerant compressors are key considerations in the design of a baseload LNG plant. Compressors and drivers account for a large proportion of the liquefaction process equipment CAPEX. With the increase in size of available refrigerant compressor drivers, plant designs have sought to lower the specific cost of LNG production by benefiting from economies of scale. As a result, the capacity of plants has grown with the development of heavy industrial gas turbines such as the Frame 5, 6 and 7 and more recently Frame 9 gas turbines.

Lighter duty higher efficiency aeroderivative drivers have been used extensively in power generation applications. They have been considered in numerous studies and their use has been implemented in Optimized Cascade[®] plants in Australia. The use of aeroderivative gas



turbines has the potential to reduce fuel consumption and associated CO₂ emissions compared to industrial gas turbines.

The thermal efficiency and output power of industrial and aeroderivative gas turbines considered for LNG liquefaction service (mechanical drive of refrigeration cycle compressors) is shown in Figure 2. Typically, industrial type gas turbines will offer thermal efficiencies at ISO conditions of approximately 29% to 35% whereas aeroderivative type gas turbines offer thermal efficiencies at ISO conditions above 40%.

It is noted that the power output of gas turbines is dependent on the mass flow through the combustion air compressor. The available power produced by the gas turbine driver depends on the local conditions i.e. ambient temperature, altitude and air humidity as these affect the density of air. If the air density at local ambient conditions is lower than the air density at ISO conditions, the power output and efficiency of the gas turbine will reduce with respect to the quoted value at ISO conditions (Ref 15). Figure 2 shows the indicative loss of performance when the gas turbines are operated in hot climates (40°C).





Notes:

1. Performance points shown at ISO ambient conditions (15°C, 60% RH, 1.013 bara) and power output and efficiency de-rated to 40°C. De-rating based on typical gas turbine performance in function of ambient temperature.



Electric motors are used as starter-helper drives for gas turbines and to boost gas turbine power output as needed. Large electric motors have been considered as an alternative to mechanical drive gas turbines. Electric motors offer greater flexibility than gas turbines as they can be engineered to match the actual compressor speed and power requirements allowing better plant optimisation. In contrast to gas turbines, performance of electric motors is practically unaffected by ambient conditions. Electric motors are not constrained by the number of starts and can restart the compressors under load. The use of electric motors also allows decoupling of the individual sizing constraints between liquefaction compressors and gas turbines associated to available compression-driver trains. The plant availability is also expected to be increased due to the higher reliability of electric motors and reduction of extended outages for periodic maintenance.

5.4 LIQUEFACTION EFFICIENCY

Liquefaction efficiency is normally measured either in terms of specific power, which is a measure of the process thermodynamic efficiency, or as thermal efficiency.

The thermodynamic efficiency includes only the liquefaction process and does not consider any energy consumed elsewhere e.g. for pre-treatment. Nor does it consider the refrigeration compressor driver efficiency. It is influenced by feed conditions, ambient conditions, the refrigeration process and equipment design.

Specific power is calculated as the ratio of shaft power absorbed by the liquefaction compressors to the LNG product routed to storage and is usually quoted in kWh/kg of LNG. In the context of CO_2 emissions and excluding the thermal efficiency of driver, increased emissions are expected on a process with high specific power as a result of increased fuel combustion requirement.

Typical specific liquefaction power figures are reported for each liquefaction technology. Liquefaction specific power considering technology and local factors is typically in the range of 0.25 to 0.4 kWh/kg for baseload liquefaction technology. Specific power around 0.3 kWh/kg is reported as typical. The lowest specific power quoted is 0.23 kWh/kg for the Snøhvit plant (Ref 22). This low figure is a result of an optimised thermal and electrical efficiency with centralised power generation based on aeroderivative gas turbines operating in a very cold environment. Liquefaction technology using single refrigerant expander cycles (e.g. in small scale liquefaction) has typical specific power in the order of 0.4-0.5 kWh/kg.

Details on design strategies focussing on reducing the specific power are included in Appendix 3.

The thermal efficiency of a process is defined as the proportion of feed gas fuel value that is contained in the products (LNG and NGL). This parameter covers both the process efficiency and the efficiency of the gas turbine drivers. Reported thermal efficiencies for liquefaction process are in the range 90-93%. The balance 7-10% represents the fuel consumption. These figures are highly dependent upon the heat rate or efficiency of the gas turbine selected.

Energy used to remove inert components and contaminants reduces thermal efficiency. Increased NGL recovery will increase the power requirement overall. The main factor affecting the thermal efficiency of the liquefaction process is the thermal efficiency of the cycle refrigerant compressor driver.

6.0 EVALUATION OF CCS POTENTIAL

6.1 SOURCES OF CO₂ IN LNG PLANTS

There are various operations within an LNG liquefaction plant that produce a stream containing CO_2 which may be targeted by capture schemes. Sources of CO_2 emissions will vary from LNG plant to LNG plant dependent on the specific plant configuration. A discussion and identification of all potential sources of CO_2 emission sources is given below.

6.1.1 AGRU

 CO_2 is removed from the feed gas typically in an acid gas removal or sweetening unit typically consisting of a chemical absorption process using amine-based solvents which are tailored to simultaneously remove H₂S if present. Practically full removal is achieved, with CO_2 content in the feed gas being reduced to less than 50 ppmv to prevent freezing in the cryogenic section of the liquefaction plant. The CO_2 is absorbed in the solvent and then released by thermal regeneration of the solvent. H₂S and sulphur compounds in the feed gas will also be absorbed in the solvent. In the absence of significant H₂S content in the feed gas, the CO_2 stream has a high purity and is saturated with water. The CO_2 stream may also contain some hydrocarbons that are dissolved in the solvent. When the feed gas contains H₂S and other sulphur compounds in moderate to high content, these will leave the acid gas stripper together with the CO_2 and further processing may be required prior to venting. If the stream has low sulphur content, it may be further processed in a thermal oxidiser to combust any hydrocarbons and convert H₂S to SO_2 . The residual stream is generally vented to atmosphere, constituting a direct source of emissions.

Indirectly, CO₂ emissions are associated with the burning of fuels to produce power required in electric equipment (machinery, control systems, etc.) and to produce heat (steam) required for the solvent regeneration reboil duty (supplementary to any waste heat recovery). Also, supplementary fuel is used in the thermal oxidiser.

The quantity of CO_2 emitted from the feed gas pre-treatment depends on a number of factors including quantity of feed gas processed, CO_2 content in the feed gas (for direct emissions) and thermal efficiency of the heat and power generation systems (for indirect emissions). CO_2 concentrations are variable and have been seen to be up to 14% at the Gorgon gas field.

Based on the currently operational LNG liquefaction facilities worldwide the Snøhvit LNG plant carries out sequestration of the 0.7 mtpa of CO_2 removed in the AGRU. A similar sequestration scheme is included for the Gorgon LNG plant to store 4 mtpa of CO_2 removed from the feed gas.

Sources of CO₂ emissions associated with the AGRU:

- Vented CO₂ removed from process stream.
- Fuel gas combustion to provide power generation for machinery (pumps, cooler fans), control, and other functions requiring electricity supply.
- Fuel gas combustion in thermal oxidiser.
- Fuel gas combustion to provide heating (steam) for solvent regeneration reboiler duty (steam demand reduced by waste heat recovery).

6.1.2 Dehydration

Treated gas leaving the AGRU is water saturated. The water content is typically reduced to a maximum of 1 ppmv to prevent freezing in the cryogenic section of the plant. This is often achieved by firstly cooling the gas with a refrigerant for bulk water removal and then passing the gas though a molecular sieve adsorption system.

Indirect sources of CO₂ emissions associated with dehydration are as follows:

- Fuel gas combustion to provide power generation for machinery (refrigerant compressors, regeneration gas compressor, cooler fans, ancilliary pumps) and regeneration gas heating (if electric), control, and other functions requiring electricity supply.
- Fuel gas combustion to provide regeneration heating (e.g. on a fired heater)

6.1.3 Mercury Removal

Mercury in the feed gas can be removed with a non-regenerative solid adsorbent bed. Feed gas pre-heating may be required to prevent heavy hydrocarbon condensation leading to adsorbent degradation.

Indirect sources of CO₂ emissions associated with mercury removal are as follows:

• Fuel gas combustion to provide gas heating

6.1.4 Liquefaction

Power required to liquefy the feed gas stream is mainly provided via the refrigerant compressor drivers, either gas turbine or electric motors. Gas turbine drives are a key source of CO_2 emissions, burning fuel to generate power used to propel rotating equipment. Where electric motors (including Variable Speed Drives) are used for the main mechanical drive of compressors or as starter-helper motors, the associated CO_2 emissions are produced from burning fuel for power generation.

Sources of CO₂ emissions associated with liquefaction:

- Fuel gas combustion to provide power for gas turbine refrigerant cycle compressor drivers (direct mechanical drive).
- Fuel gas combustion to provide power generation for alternative electric motor refrigerant cycle compressor drivers and starter-helper motors.
- Fuel gas combustion to provide power generation for other machinery (LNG pumps, BOG compressors, cooler fans, ancilliary pumps), control, and other functions requiring electricity supply.

6.1.5 Nitrogen Removal

Removal of nitrogen, usually integrated (when required) in the end flash gas system requires power primarily for pumping and re-compression of treated gas. Compression power can be provided either by using a gas turbine in mechanical drive service or an electric motor.

Sources of CO₂ emissions associated with Nitrogen Removal:


- Fuel gas combustion to provide power for gas turbine drives or fuel gas combustion to provide power generation for electric motor drives used in recompression of treated gas
- Fuel gas combustion to provide power generation for machinery (process pumps, starter-helper motors, cooler fans, ancilliary pumps), control, and other functions requiring electricity supply.

6.1.6 NGL Recovery and Liquids Processing

The NGL separated in the scrub column is separated into Propane, Butane and Condensate in a series of distillation columns.

Condensate separated in the inlet facilities is subject to stabilisation in a reboiled column.

If hydrate inhibitor (e.g. methanol or MEG) is injected in upstream systems, this is recovered by distillation of the aqueous stream separated in the inlet facilities.

Indirect sources of CO_2 emissions associated with the operation of distillation processes:

- Fuel gas combustion to provide power generation for machinery (process pumps, cooler fans, gas compressors, ancilliary pumps) control, and other functions requiring electricity supply.
- Fuel gas combustion to provide heating for de-ethaniser, depropaniser, debutaniser and condensate stabilisation reboiler duties.

6.1.7 Flaring

Flaring operations at an LNG plant may arise from a wide range of events during commissioning, during a process upset or emergency, when starting-up the process or in preparation of a shutdown (e.g. prior to maintenance activities). Sources of CO₂ emissions associated with flaring are as follows:

- Ocasional combustion of flammable gases in flare from relief and depressurisation
- Routine vented emissions from storage facilities
- Continuous burning of fuel gas in flare pilots

No CO₂ capture scheme is expected to be associated to flare stacks as emissions occur only in exceptional events, and the annual CO₂ emissions can be considered as marginal when compared to the emissions from other sources.



6.2 CCS SCHEMES IN LNG PLANTS

Potential exists for sequestration of CO_2 contained in the reservoir gas and capture and sequestration of CO_2 produced by combustion of fuel gas in gas turbines (in mechanical drive and power generation service).

1. CO₂ from reservoir gas

 CO_2 with high purity is already captured and typically vented to atmosphere in the AGRU. Infrastructure required is limited to CO_2 compression and purification. The sequestration scheme is particularly attractive where CO_2 in large quantities is contained in the feed gas.

2. CO₂ from fuel gas combustion

The gas turbines driving the refrigerant cycle compressors are typically the major emmitters of CO_2 . Gas turbines in power generation service also contribute to the CO_2 emissions. These depend on the power generation capacity dictated by specific requirements of each plant and location.

Quantification of the potential CO_2 emissions associated to generation of heat depends on a large number of factors, including site ambient conditions (impacting efficiency of combustion equipment, boilers, heaters, etc.); heat integration schemes installed such as waste heat recovery; heating requirements in function of feed gas processing (feed gas flow, CO_2 content, NGL content). It is expected that a large proportion (or all) of the heating demand from the LNG train will be met by waste heat recovery.

A CO₂ capture process either by pre-combustion (replacing natural gas as fuel gas for gas turbines) or post-combustion (capturing CO₂ from flue gases) can be implemented (see Appendix 5). Where electric motors are used as refrigeration cycle compressor drivers (as in the Snøhvit plant), the capture scheme would be applied to the centralised power generation.

It is noted that the CO_2 capture process itself will require additional power and heating, and additional fuel gas will be required to provide with this extra demand. Therefore, the capture process will produce additional CO_2 emissions.



6.3 GAS TURBINE EMISSIONS

The type of gas turbine technology and configuration for power generation plays a role in the amount of CO_2 emitted. Based on natural gas for fuel, for a simple cycle about 0.5 kg CO_2 per kWh (aeroderivative turbines) to 0.6 kg CO_2 per kWh (industrial turbines) will be released. Gas turbines in combined cycle typically emit as low as about 0.4 kg CO_2 per kWh (Ref 13).

CO₂ emissions will also be affected by ambient conditions as the performance (output power and thermal efficiency) of the gas turbine will reduce, particularly with increasing air temperature leading to use of more fuel (and hence more emissions) per unit of output power.

Gas turbines in mechanical drive applications are reported to have a higher efficiency compared to the efficiency when the same gas turbine models are used for power generation service. However, taking advantage of the increased efficiency of a combined cycle for power generation against simple cycle, the use of electric motors to drive the liquefaction cycle compressors may yield lower CO_2 emissions due to the improved efficiency overall.

The fuel type used in the gas turbines affect CO_2 emissions. Table 1 shows the CO_2 emissions for gas turbines in LNG service based on natural gas fuel (see Appendix 7 for representative basis of design). Propane fuels will yield 15 % higher emissions than natural gas and fuel oil will yield 50 % more (Ref 13).

Table 2 shows an indicative CO_2 production based on typical LNG train configurations including an allowance for power generation. This shows that for a typical baseload train capacity the potential for CO_2 capture is in the order of 1 mtpa.

		Mechanical drive		Power Generation (Note 5)	
Gasi	Gas rurbine		De-rated (Note 4)	ISO (Note 3)	De-rated (Note 4)
	Frame 5	0.57	0.61	0.58	0.63
Industrial	Frame 6	0.61	0.66	0.64	0.68
(Note 1, 2)	Frame 7	0.62	0.66	0.62	0.67
	Frame 9	0.59	0.63	0.60	0.65
	PGT25+	0.49	0.53	0.51	0.55
Aero-	LM6000	0.48	0.52	0.49	0.53
derivative	RB211 DLE	0.50	0.54	0.52	0.56
	Trent 60 DLE	0.47	0.50	0.48	0.52

Table 1. CO ₂ emiss	sions (kg/kWh) fo	r gas turbines	in LNG service
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Notes:

- 1. Estimates based on natural gas fuel (LHV=48.9 MJ/kg, see Appendix 7 for representative fuel gas basis considered).
- 2. Uses industrial turbine data for MS5002E, MS6001B, MS7001EA and MS9001E gas turbines.
- 3. ISO ambient conditions 15°C, 60% RH, 1.013 bara.
- 4. Power output and efficiency de-rated to 40°C. Indicative de-rating based on typical gas turbine performance in function of ambient temperature.
- 5. Gas turbines in simple cycle.



Driver	Number of drivers	Train Shaft Power (MW, ISO)	Power Generation Demand (MW) (Note 1)	LNG Train Capacity (mtpa) (Note 2)	CO ₂ annual emissions (mtpa) (Note 4)
PGT25+	6	188	46	5.1	1.0
LM6000	6	262	64	7.1	1.4
Frame 5C	6	170	42	4.6	1.2
Frame 5D	6	195	48	5.3	1.4
Frame 7	2	175	43	4.7	1.1
Frame 9	3	276 (Note 3)	114 (Note 3)	7.5 (Note 3)	1.9

Table 2. Indicative CO ₂	production for re	presentative lique	efaction train o	onfigurations
	pi • • • • • • • • • • • • • •			garatione

Notes:

- 1. Assumed power generation based on typical/representative 9 MW per mtpa of LNG. Power generation will depend on specific plant electrical load.
- 2. Indicative LNG capacity estimated based on representative specific power 0.3 kWh/kg, 93% plant availability and gas turbine performance (output power and efficiency) at ISO conditions.
- 3. The LNG capacity reported for the Frame 9 configuration is as per AP-X trains in Qatar, where about 70% of the power output (276 MW) is used to drive refrigerant compressors while the rest is exported as power via a generator in the same shaft.
- CO₂ emissions from gas turbine drives and power generation based on use of natural gas as fuel gas (LHV=48.9 MJ/kg, see Appendix 7 for representative fuel gas basis considered) and 93% plant availability.



6.4 SPECIFIC CO₂ EMISSIONS OF LNG LIQUEFACTION

Based on the typical range of baseload liquefaction specific power (0.3-0.4 kWh/kg) and estimation of natural gas fuel gas requirements for typical gas turbine drives, CO_2 emissions for LNG liquefaction plants are typically in the range 0.20 to 0.28 tonne of CO_2 per tonne of LNG for industrial heavy duty gas turbines. This range can be reduced to about 0.15 to 0.21 tonne of CO_2 per tonne of LNG if more efficient aeroderivative gas turbines are used.





Notes:

- 1. CO₂ emissions based on use of natural gas as fuel gas (LHV=48.9 MJ/kg, see Appendix 7 for representative fuel gas basis considered).
- 2. Graph shows emissions from gas turbines used for mechanical drive of liquefaction refrigerant cycle compressors only (excludes power generation and other fuel combustion sources).
- 3. Power output and efficiency de-rated to 40°C. De-rating based on typical gas turbine performance in function of ambient temperature.
- 4. Representative gas turbine performance data based on PGT25+ efficiency (~41% ISO, ~38% derated) for aeroderivative gas turbines. Representative performance of industrial gas turbines is based on averaged efficiency (~30% ISO, ~28% de-rated) of Frame 5 (MS5002C and MS5002D) and Frame 7 (MS7001EA). This is applicable for typical baseload plant capacities in the range of approximately 3.5 5.5 mtpa. For Frame 9 (MS9001E) gas turbines applied to larger plants ~8 mtpa, a reduction in the specific CO₂ emissions with respect to the industrial gas turbine figures quoted is achievable due to the improved efficiency (~35% ISO, ~32% de-rated), but aeroderivative gas turbines still have the lowest emissions.

Figure 4 shows combined CO_2 emissions (representative) from gas turbine flue gases (in function of liquefaction efficiency and type of gas turbine used) and CO_2 separated from the feed gas in the acid gas removal unit (for representative extremes of CO_2 content in feed gas). Combined emissions are in the approximate range 0.3-0.4 tonne of CO_2 per tonne of LNG for typical plants with relatively low CO_2 content in feed gas (<2 mol%) and up to around 0.7 tonne of CO_2 per tonne of LNG for plants liquefying gas with a high CO_2 content.



Figure 4. Representative CO₂ emissions from LNG plant from liquefaction cycle compressor gas turbines and acid gas removal



Notes:

- 1. Notes on Figure 3 also apply. Emissions calculated based on de-rated GT performance.
- 2. Power generation for baseload LNG plants based on typical/representative 9 MW per mtpa of LNG, resulting in power generation duty being about 25% of the refrigeration cycle power.

Figure 5 shows the reported total plant emissions (including power generation and other sources) for representative LNG plants.



Figure 5. Reported CO₂ emissions from representative LNG plants (Ref. 50)



An indicative CO_2 production based on typical LNG train configurations is shown in Table 3. This analysis has been extended to cover a range of specific power (within the typical figures for baseload plants) and plant capacities as shown in Figure 6. The data includes the CO_2 removed from feed gas in the AGRU and CO_2 produced by combustion of fuel in gas turbines in mechanical drive and power generation service.

The use of aeroderivative gas turbines has been used as a basis, and the additional CO_2 production (about +30%) associated to the use of industrial gas turbines is included. It is noted that for an LNG plant capacity of 8 mtpa only Frame 9 gas turbines are considered as per current implementation. The required plant capacity for post-combustion capture of CO_2 from gas turbine drives for the given reference liquefaction plant capacities is given in Table 3.

LNG Liquefaction Plant Capacity (mtpa of LNG)	CO₂ Capture Plant Capacity (tonnes CO₂/day) (Note 1)	CO ₂ Capture Annual Capacity (mtpa of CO ₂) (Note 1,2)
3.5	1900 – 2500	0.8 – 1.1
4.5	2400 – 3200	1.1 – 1.4
5.5	2900 – 3900	0.3 – 1.7
8.0	4500 – 5900	1.5 – 2.0

Table 3. CO₂ Capture Plant Capacities for given LNG Liquefaction Plant Capacity

Notes:

- Based on post-combustion capture with 90% CO₂ recovery from flue gases from gas turbines (mechanical drive and power generation). Figures are for industrial gas turbines, which produce about 30% more CO₂ emissions than aeroderivative gas turbines. The range of CO₂ plant capacity shown is for the consideration of liquefaction power specific in the range 0.3 to 0.4 kWh/kg.
- 2. Based on 93% annual plant availability

Two graphs are shown in Figure 6 to show the difference in the potential CO_2 sequestration when the feed gas contains 2 mol% CO_2 (representative of typical feed gas) and 14 mol% (representative of feed gas with relatively high CO_2 content, as in the Gorgon LNG development). If the CO_2 removed from feed gas in the AGRU is combined with the CO_2 captured from the gas turbines for sequestration, the proportion of the former with respect to the total CO_2 sent to storage (depending on the specific power assumed and gas turbine type) is:

- Around 20% of the total CO₂ for 2 mol% CO₂ content in feed gas
- 60 to 70% of the total CO₂ for 14 mol% CO₂ content in feed gas

The supply of power required to capture, compress and purify the CO_2 generated by gas turbines will produce additional emissions as shown in Figure 6. The compression and purification of CO_2 separated from the feed gas also incur in additional emissions associated to the generation of power to operate these systems. However, the specific emissions (per unit of captured CO_2) of CO_2 separated from feed gas are much lower as CO_2 is considered to be already captured and the stream available at high purity (see Figure 7). The lower associated emissions and the lower costs (both capital and operating) required to sequestrate CO_2 from reservoir gas (feed gas) has more potential than the capture and sequestration of CO_2 from combustion processes (gas turbines).





Figure 6. Potential CO₂ Capture and Sequestration Capacity for LNG Plants





Liquefaction Specific Power (kWh/kg) and LNG plant capacity (mtpa)

Notes:

- 1. Notes on Figure 3 and Figure 4 also apply. For large plants (8 mtpa) 10% of the cycle power is assumed to be the power generation required.
- CO₂ capture capacity is based on post-combustion capture removing 90% of the CO₂ in the flue gases from all of the gas turbines in mechanical drive and power generation service. CO₂ from AGRU is based on reduction of CO₂ content to 50 ppmv (i.e. practically full removal).
- 3. Emissions are based on 10% of the CO₂ in flue gases (i.e. the proportion not captured). The CO₂ associated to the generation of power required to operate the capture plant (assumed post-combustion capture plant based on chemical absorption) plus compression and purification of the full CO₂ stream to storage (including CO₂ separated from feed gas) is assumed to be emitted (not captured). It is assumed that heating duty for solvent regeneration can be supplied by waste heat recovery.







Note 1. CO₂ associated to the operation of the CCS scheme assumed to be emitted (not captured)

The additional emissions associated with the operation of the CCS schemes may be captured in integrated schemes (e.g. if power for the capture plant is generated by the power generation gas turbines of the liquefaction plant) or not captured (e.g. if power is produced by gas turbine dedicated to the capture plant and exhaust gas not routed to the capture plant, or if electricity is imported from the grid). Figure 8 shows the CO₂ captured and avoided for representative plant capacities and for the option to exclude and to include capture of the emissions associated with the operation of the CCS scheme.





Figure 8. CO₂ Captured and Avoided for Representative LNG Plants



Notes:

- 1. Notes on Figure 3, Figure 4 and Figure 6 also apply. CO₂ associated to the operation of the capture, compression and purification process is assumed to be captured as noted.
- 2. Reference plant is an LNG plant with capacity indicated and no CO₂ capture (representative specific liquefaction power 0.3 kWh/kg of LNG is used). Plant with CCS is the equivalent LNG plant with post-combustion capture from gas turbines (both in mechanical drive and power generation service) and compression and purification of captured CO₂ together with CO₂ removed from feed gas.
- 3. If emissions associated to the operation of the capture plant are considered to be captured, it will increase the size and power requirements of the capture plant, and emissions are therefore increased.



6.5 LNG PLANTS WITH POTENTIAL FOR CCS

 CO_2 content in the reservoir gas will likely play a major role in LNG developments with CCS potential. CO_2 needs to be removed from the liquefaction plant feed gas as part of the pretreatment requirements, therefore no further capture processing steps (other than compression and purification of the produced CO_2 stream) is required. Compared to a capture scheme from combustions processes, sequestration of CO_2 removed from feed gas reduces overall plant complexity and footprint, capital and operating costs and additional emissions associated with combustion capture.

Reservoir gas fed to liquefaction terminals typically contains around 2 mol% of CO_2 (e.g. Pluto, Rasgas LNG, Qatargas LNG, Nigeria LNG). Following CO_2 removal, the produced stream is vented and contributes to the total plant emissions, but as shown in Figure 6, the emissions from combustion processes are much higher. Higher CO_2 concentrations will result in substantially higher CO_2 emissions from the treatment of the reservoir gas in the AGRU and will provide a good basis upon which a case for CCS implementation may be built. A major benefit may be obtained by sequestrating the CO_2 removed from the feed gas only. This is the case for the only two CCS schemes on LNG liquefaction facilities: Snøhvit (8.0 mol%) and Gorgon LNG (up to 14 mol%).

Based on CO_2 composition in the reservoir gas, plants may show potential for CCS due to high CO_2 concentration in the reservoir gas, such as the case of gas fields feeding Darwin LNG (6 mol%), Prelude LNG (9 mol%), Browse LNG (12 mol%) and Ichthys LNG (8-17 mol%). CCS will play a major role for gas fields with high CO_2 content (i.e. above 20 mol%) with gas development schemes likely to include CCS as part of the economics of the project. However, it may be expected that development of gas reserves with low to moderate CO_2 (less than 20 mol%) will take priority, and that gas fields with high CO_2 content will still be considered uneconomical to develop in the near future.

A number of additional drivers need to be present for a CCS scheme to be realised, including relatively high penalties on emissions (i.e. CO_2 tax), the availability of local storage, availability of funding and/or other financial incentives and other conditions contributing to the commercial feasibility of the overall scheme.

The LNG industry projects are characterised by the use of proven designs where technology risks are minimised, so the use of conventional technology is always preferred. Only well proven technologies will be considered, with schemes that minimise the risks of disruptions to LNG production. A number of CO_2 capture technologies are commercially available, supported by extensive testing and recent operational experience in large scale projects. Post-combustion CO_2 capture using conventional and well proven chemical absorption technology is likely to be the preferred route.

Although incremental performance improvements have been delivered by using proprietary capture technologies, CO_2 capture remains as an energy-intensive process. For example, chemical absorption post-combustion capture processes have a significant heating demand for solvent regeneration. The operating costs can be significantly reduced for processes where heating or power generation excess capacity exists. This may allow a portion of the CO_2 capture plant requirements to be provided at negligible cost. LNG plants have a significant potential on this regard as there is usually scope for additional waste heat recovery at the exhaust of gas turbines in mechanical drive or power generation service.

In terms of power generation, considerable potential exists for liquefaction plants where multiple gas turbines are installed with total power output in excess of the liquefaction train

requirements, motivated by future requirements of a phased development or by machinery selection. In this scenario, power generation gas turbines are operated at part load and mechanical drive gas turbines operated at full load with excess power exported. If operation of power generation gas turbines at increased load or use of surplus power otherwise exported allows the power requirements of the capture plant to be met, it will greatly improve the prospects of CCS.

Similarly, it is expected that LNG liquefaction plants will offer the potential to meet the heating requirements of the capture plant via heat recovery from gas turbines. LNG facilities where heat recovery systems have not been installed on all turbines, could provide heat for the CCS plant by installing heat recovery system on the turbines whose waste heat is not utilised. This would incur capital expenditure for the installation of additional equipment related to the new heat recovery system. However, the additional capital cost is relatively low (when compared to the overall cost of the CCS project) and expected to be justified by the savings made in operating costs.

The suitability of available geological formations for storage located in the vicinity of LNG plants may restrict implementation of CCS due to lack of experience for CO_2 storage (uncertainties on expected interaction of CO_2 and plume development) or limited storage capacity against the life time of the plant. Injection of CO_2 into unknown geological formations may increase cost and perceived risk of the scheme. Significant work may be required to gain confidence in the characterisation of the geological formations and need for extensive monitoring may lead to cost increases. Common CO_2 storage across a network of capture schemes may reduce overall costs and improve the feasibility of the CCS scheme.

The location of LNG liquefaction plants may play a role to facilitate the implementation of CO_2 capture technology due to delivery of equipment and consumables. LNG liquefaction plants generally have the advantage over other industries in which CCS could be adopted due to their locality to marine loading/unloading for transportation. On the other hand, the location of the LNG liquefaction plant may limit the feasibility of CCS projects due to availability of infrastructure for plants located in remote locations.

The technical and commercial feasibility leading to successful implementation of a CCS scheme is expected to be defined when the LNG production scheme is developed. Implementation of CCS at a later stage may only occur as a part of a phased development already identified. Otherwise CCS implementation may be unlikely, unless the local CO₂ tax system changes significantly in the future.

New build plants are expected to have greater potential for CCS due to the ability to optimise the design in order to help facilitate the incorporation of the CCS scheme. Movement towards capture-ready plants, while allowing for reduction in costs, will also encourage implementation of CCS in LNG by providing design allowances (tie-ins, allocated plot space, spare power generation capacity) to facilitate for installation of future CO₂ capture plants. This is important in terms of plant plot allocation (where space may be a constraint), considering that comparable to full scale CO₂ capture plants are of a significantly large size. Implementation of CCS schemes on existing (non-capture ready) plants may lead to difficulties in locating a suitable plot for installation of equipment, the need to install additional power and steam generation capacity and tie-ins which may impact budget and schedule of the project.



6.6 INTEGRATION OF CCS WITH LIQUEFACTION PLANT

The LNG industry projects are characterised by the use of proven designs where technology risks are largely mitigated, so the use of conventional technology is always preferred. The implementation of a CCS scheme will have to take into account the stakeholder interactions and the risk-averse nature of the industry. Only well proven technologies will be considered, with schemes that minimise the risks of disruptions to LNG production.

A number of CO_2 capture technologies are commercially available, supported by extensive testing and recent operational experience in large scale projects. Post-combustion CO_2 capture using conventional and well proven chemical absorption technology is therefore likely to be the preferred route. CO_2 capture technologies may be more attractive by reducing the perceived risks if current plant operations have experience in using a similar process, i.e. using amine processes for both the AGRU and the CO_2 capture plant.

In a post-combustion capture scheme, an interface between the gas turbines flue gas exhaust ductwork and the inlet to the CO_2 capture plant is required. Considerations regarding this interface include:

- Gas Turbine Exhaust System. Tie-ins in gas turbine exhaust stacks are needed to
 route flue gases to the capture plant, with valves/dampers included to allow venting of
 flue gases when the capture plant is not available. In a retrofitted plant, modifications
 to exhaust ducting will be necessary to accommodate tie-ins and potentially a WHRU
 where this is not installed but required to provide heating to the capture plant.
- Flue gas ducting. Ducts from the exhaust of each gas turbine (either in mechanical drive or power generation service) will need to be combined into a substantially large flue gas duct through which the total flue gas flow will be transported. It is noted that the gas turbines may be installed in different plant locations and gas gathering may be challenging. For typical post-combustion capture schemes, a blower on the inlet to the capture plant will boost the pressure of the flue gas.
- Plant layout. It is not unusual for the LNG liquefaction train layout to be compact, and installation of tie-ins and the large interconnecting flue gas ductwork may represent a challenge, potentially requiring substantial modifications to pipework and structures of existing plants to be retrofitted or considerable space allowances for new build capture-ready plants.
- Performance of Gas Turbines. In a retrofitted plant, increased pressure drop through new ducting and WHRU will lead to increased backpressures in the gas turbine outlet, which may detrimentally impact on performance of the gas turbine, ultimately impacting on efficiency of the liquefaction process, ultimately reducing the LNG production capacity.
- Control complexity. It is anticipated that additional control devices may be required as well as including for the ability to alternate between routing flue gas to the capture plant or when the capture plant is unavailable, routing the flue gas as originally intended to the existing stack in bypass mode of operation. Operation of shared power generation facilities at part load during periods when the capture plant is not operated needs to be considered in the design (machinery selection).
- Tie-ins. Scheduling and phasing of tie-ins for interconnecting pipework and to facilitate the installation of the capture plant would need to be accounted for in order to minimise the impact on LNG production in existing plants being retrofitted. A



capture-ready plant is expected to have allowances (space and valves) for these connections.

Installation of a large duct is also required to route the CO_2 stream from the AGRU to the CO_2 product compressor, and considerations for tie-ins (with AGRU CO_2 vent stack) and layout may represent a challenge, particularly in retrofitted plants. The possibility of a CO_2 compressor dedicated to for the CO_2 from AGRU may be considered to expand operation flexibility. This may be the particular case of phased schemes where capture from AGRU is initially implemented as a project separate to post-combustion capture.

For new build applications, the issues associated with the interface between the source of emissions and capture plant should be considered in design of the overall facility and there is potential for the impacts to be reduced through consideration of CO_2 capture in the layout and configuration of the plant, readiness of key tie-ins, location of emission sources relative to the capture plant, selection of gas turbine type, etc. Nevertheless, these factors will add complexity and costs to the LNG plant design when compared to the design of a facility without design allowances for the CO_2 capture plant.

An important integration aspect for new build (capture-ready) LNG plants is the possibility for provision of power generation and waste heat recovery capacity covering the requirements of both LNG liquefaction train and CO_2 capture plant (see Section 6.5). This includes the potential to include a number of gas turbines operating efficiently at part load to meet the power demand of the liquefaction train only, whilst having sufficient installed output capacity to deliver the power needed for the capture plant. As an alternative, large efficient gas turbines (e.g. Frame 9) may be selected as drivers for the refrigerant compressors, with the surplus power being used for the capture plant or exported when the capture plant is not operating. Waste heat recovery capacity can be installed to meet the overall demands of the LNG plant the CO_2 capture and purification, or at least space allowances left in exhaust systems for future WHRU installation.

The following options for configuration of a post-combustion CO_2 capture plant (chemical absorption) can be envisaged. These options are dictated by operational availability considerations (requirement to maximise availability of CO_2 capture plant with parallel trains or stand-by equipment), physical construction limitations of equipment, plot area available and phased development.

a) One large centralised CO₂ capture plant

Flue gases from all individual CO_2 major emitters are transported to a centralised CO_2 capture plant. Construction of a flue gas gathering system of significant size (but low design pressure) is required. The size of the CO_2 capture plant will be considerable and may be constrained by available space. The capture plant size may be limited by physical construction constraints, with current experience for an equivalent post-combustion capture of 1.4 mtpa of CO_2 from flue gas at a flowrate in the order of 1600 t/h.

b) Common stripping column with multiple absorber columns

Absorbers are the largest equipment expected due to the relatively large volumes of gas to be treated, with cross sectional areas in the order of 400 m². In full size capture plants, absorbers have been designed as towers constructed in concrete or steel with a rectangular cross section. Multiple absorbers of reduced size (footprint only) with a centralised stripping column could be considered as a concept. However,

feasibility of this scheme appears to be difficult to implement due to the complexity involved, e.g. the need to install individual or centralised direct contact coolers and booster fans. This scheme is expected to have higher capital costs than the single absorber option and assumes the availability of space.

Parallel absorbers could be considered when the combined flue gas flow is such that construction of large towers involves significant risks potentially impacting costs and performance.

c) Multiple small-scale CO₂ capture plants

Consideration for a small-scale CO_2 capture plant treating a slip-stream of the flue gas may be the preferred route in demonstration schemes or where a phased development is required, particularly regarding access to funding. This enables confidence to be gained in the scheme before investing further. However, consideration of multiple CO_2 capture plants treating gas from individual emitters (i.e. on each gas turbine) in an LNG train is expected to result in large capital costs if implemented as full scale project.

The size of the CO_2 capture plant expected for a typical baseload liquefaction train is comparable to the largest CO_2 capture plants currently installed, with equipment near physical construction limits (particularly cross-sectional area of columns). Therefore, a scheme in which a single CO_2 capture plant processes the full flue gas from multiple LNG trains is not expected.

Phased installation of additional CO₂ capture capacity may be beneficial in order to match storage development. Also, reduction in capital and operating costs can be realised for installation of additional future capacity (i.e. additional CO₂ capture trains) if it is possible to share facilities and resources (e.g. pipelines, storage, buildings and roads, utilities, power generation capacity, labour, etc.). Installed spare capacity in power generation and utilities and availability of plot space needs to be considered if this has been reserved for future LNG trains.

 CO_2 from the capture plant and AGRU will be combined, compressed and purified (dehydration and oxygen removal etc.) as required prior to being transported to storage. The configuration and size of the CO_2 product compression, dehydration and transport system depends on the relative and the total volumes of gas to be processed from the two sources (AGRU and post-combustion CO_2 capture plant). The flowrate of CO_2 to be sequestrated may be limited by constrains such as compressor size, subsea pipeline size, injection rates limits, etc.

In the AGRU a significant proportion of H_2S and sulphur compounds plus some hydrocarbons may be separated in addition to CO_2 . It is expected that the content of impurities will typically be below the specification for CO_2 for storage, particularly on hydrocarbon content. However, for some LNG plants, depending on the composition of the natural gas feed to the LNG plant, the CO_2 stream produced in the AGRU may undergo further processing steps (e.g. in an oxidiser or in a Claus plant) to remove the sulphur components and burn the hydrocarbons prior to venting. This will result in the residual CO_2 stream being delivered at an elevated temperature and containing high levels of oxygen and SO_2 . This may prevent this stream to be considered for sequestration without significant processing required to purify the stream.



7.0 CCS PROJECTS IN LNG PLANTS

7.1 OVERVIEW

LNG plants provide a good opportunity for CCS because of the presence of CO_2 streams already generated by CO_2 removal from feed gas pre-treatment in the AGRU. Sequestration schemes have already been put in place for plants processing feed gas with a high CO_2 content. There are currently no LNG liquefaction terminals operating with CO_2 capture from fuel gas combustion processes.

Potential exists for significant CO_2 capture from combustion of fuel in gas turbine drives and power generation units. Liquefaction cycle power for a typical baseload LNG train is in the order of 150-200 MW, which is similar to the output of Coal Power Plants where large scale CCS projects have been implemented. In addition to the potential for CO_2 capture, the feasibility of CCS schemes is supported by the potential availability for storage reservoirs for capture of the CO_2 being located in the geographical vicinity.

To date, there are only two LNG liquefaction plants in the world that have CCS schemes implemented: Snøhvit in Norway and Gorgon in Australia. The Browse project in Australia was also proposed but has been put on hold. An overview of these three projects is given below.

An overview of other relevant full-scale CCS projects is included in Appendix 4. Although these projects are related to the wider natural gas processing and post-combustion capture in power generation, relevant considerations are applicable to the potential implementation of CCS in LNG liquefaction plants.

7.2 SNØHVIT

The Snøhvit project routes gas from offshore gas fields off the coast of Norway at Hammerfest to an LNG liquefaction plant at Melkøya. The Snøhvit area is subject to strict environmental standards enforced by Norwegian authorities to protect the marine ecosystem and local fishing industry. The Norwegian State mandated CCS as a condition of the license to operate for Snøhvit.

The plant has the capacity to produce 4.3 mtpa LNG and has been producing LNG since 2007.

The gas arriving at the plant contains approximately 5-8 mol% CO_2 , which is higher than typical LNG liquefaction plant feed gas. The CO_2 capture scheme is limited on the removal of CO_2 from the feed gas to the LNG plant. A chemical solvent-based (amine) absorption process captures 0.7 mtpa of CO_2 from the feed gas to prevent it freezing in the downstream liquefaction process. The treated gas then contains less than 50 ppmv of CO_2 .

The captured CO_2 is dried in order to reduce corrosion of piping and equipment and compressed to before being piped offshore in an 8" inner diameter pipeline and injected at pressure of 80-140 bara and a flowrate of approximately 2000 ton/day into dedicated geological reservoirs in the large Tubåen sandstone for long term storage (Ref 3). The Tubåen formation is a saline aquifer lying 2600 m below sea level and around 100-200 metres below the gas cap at Snøhvit.

An estimated 0.9 mtpa of CO_2 is produced each year by the combustion of fuel gas to power the liquefaction process. This CO_2 is not captured, as there was deemed to be a lack of economically feasible technologies for the treatment of exhaust (flue) gases. It has been reported that CO_2 injection was halted because of operational challenges at the LNG plant (Ref 3) and unexpected rises approaching rock fracture pressure within the original storage reservoir. Injection had to be re-routed to a fallback reservoir following unsuccessful attempts to remedy the problem.

During the 30 years lifetime of the Snøhvit LNG project, approximately 23 million tonnes of CO_2 are expected to be captured and stored. The potential storage capacity of Tubåen formation has not yet been determined, as this depends on the unknown nature of the formation in terms of connectivity. In principle, it is expected that the compressibility of the formation would allow a storage capacity for between 8 and 15 million tonnes of CO_2 . Further development considering well designs or multiple wells would be needed to achieve the required storage capacity throughout the life of the Snøhvit plant (Ref 3).





7.3 GORGON

The Gorgon project in Australia is the world's largest CO_2 sequestration project on an LNG liquefaction plant. Feed gas containing up to 14 mol% CO_2 is routed from the Greater Gorgon Fields to a processing facility on Barrow Island, Western Australia. The facility consists of two liquefaction trains which have the capacity to produce 10.4 mtpa LNG (5.2 mtpa per train). A third processing train is scheduled to come online in 2017 adding a further 5.2 mtpa LNG production capacity.

The dominant driving force behind the decision to implement CCS as part of the Gorgon LNG project is expected financial penalty on carbon emissions, often referred to as carbon tax. This was the basis upon which a business case could be built to justify the cost of approximately \$2 billion to implement the sequestration scheme (Ref. 51). The feasibility of the project was largely benefited by the agreement of the Australian government to accept long term responsibility and liability from the storage of CO_2 in underground geological reservoirs in relation to the Gorgon LNG project.



The project received funding of approximately \$60 million from the Australian Government's Low Emissions Technology Demonstration Fund (Ref. 4). The fund aims to help Australian companies develop technologies to reduce GHG emissions at commercial scale in the long term.

As in the Snøhvit project, the CO_2 capture is limited to storing the CO_2 already removed from the feed gas in the existing ARGU, rather than venting the CO_2 to atmosphere as was the case prior to the implementation of the CO_2 sequestration project. CO_2 emissions from combustion of fuel gas to meet power and heating demands within the liquefaction process are not captured and are vented to atmosphere.

 CO_2 capture is achieved through the use of three CO_2 removal trains, each of which uses a chemical solvent-based (activated methyldiethanolamine, aMDEA) absorption process to remove the CO_2 . Once removed from the process the CO_2 is compressed in a four-stage compression system to a supercritical state, and then transported by a 12 km pipeline to the injection site for storage.

Due to the presence of water in the CO_2 stream there are inherent safety and operability concerns around the formation of carbonic acid leading to corrosion of the piping and equipment. Consideration was given in the design of the CO_2 compressors to maximise the amount of water removal. This is achieved by operating the third compression stage within a fixed pressure range (Ref 4).

The CCS scheme will capture between 3.4 and 4 mtpa of CO_2 and store it at a depth of 2 km below Barrow Island, in the Dupuy Formation. It is expected that 100 million tonnes of CO_2 will be injected over the life of the Gorgon Project.

Production of LNG from the first liquefaction train commenced in March 2016 and as such there is little operational experience available.

7.4 BROWSE

The Browse FLNG project was planned for the development of three gas fields in the Browse Basin, Western Australia. A key component of the project involved removing CO_2 from the feed gas in the FLNG plant with capacity to produce 3.9 mtpa of LNG. The CO_2 would then be transported and stored. The project was developed up until Front End Engineering Design stage where the project was put on hold due to commercial and technical viability under the prevailing energy market conditions.



8.0 CO₂ CAPTURE ROUTES

8.1 CAPTURE FROM COMBUSTION PROCESSES

Emissions can be reduced by improving heat integration and/or efficiency of the liquefaction process, (see Appendix 3) as less fuel gas is used in combustion processes. While this approach has a positive impact on reducing the emissions from an LNG facility, the amount of emissions reduction is not comparable to that achieved by a CO₂ capture scheme.

In addition of the gains achieved by efficiency improvements, decarbonisation of technologies requires the application of a CO_2 capture scheme targeted at reducing CO_2 emissions from combustion processes, typically by 90%. There are currently three main routes for CO_2 capture:

- Pre-combustion
- Oxyfuel combustion
- Post-combustion capture

An overview of each of the CO₂ capture routes is provided in Appendix 5.

8.2 ROUTE SCREENING AND SELECTION

A summary outlining the advantages and disadvantages of post-combustion, pre-combustion and oxyfuel CO_2 capture routes are summarised in Table 4, particularly considering implementation of CO_2 capture schemes in LNG plants for retrofit and new build scenarios.

Pre-combustion capture involves a number of technical complexities, including additional equipment and modifications for combustion equipment to use hydrogen rich fuel, and the dependency of the liquefaction operations on fuel gas production. Also, pre-combustion schemes involve a loss of flexibility in the fuel gas system as EFG and BOG still would need to be used as fuel gas (by another user on site, e.g. boiler, gas engine driven generator, etc.). In any case, it could be considered that given than an external process (e.g. reformer) is required to produce decarbonised hydrogen rich gas, the capture is carried out outside of the LNG plant boundary. Consequently, for the purpose of this study, it will not be considered.

Oxyfuel combustion will not be considered further for the purposes of this study on the basis that this technology is still under development and not yet available or tested on a commercial scale. Consideration of availability of the oxygen generation plant impacting the availability of the LNG plant regardless of whether installed as retrofit or new build also supports this route being ruled out.

CO₂ capture via post-combustion is the most adequate route due to the technical maturity and adequate performance in either a new build or retrofit scenario. Post-combustion can be installed without increasing risks in design or affecting the performance of the core liquefaction process. The possibility to use conventional gas turbine technology and proven capture processes reduces risks. The post-combustion capture plant could be designed to be bypassed when unavailable such that there is minimal impact on LNG production availability. This scheme requires a minimum number of modifications to existing equipment, which makes this route particularly attractive as a retrofit to existing plants. In a new build, postcombustion capture technologies have the potential to allow the overall liquefaction plant (process and plant layout) to be optimised e.g. by better power and heat integration and still allowing liquefaction and capture processes to operate independently.



Route	Advantages	Disadvantages
	Typically removes 85-90% of the CO ₂ Technology can be retrofitted with minimal modifications to liquefaction process	Flue gas is close to atmospheric pressure and the concentration of CO ₂ is lower than in pre-combustion, so a more energy intensive chemical solvent is required
Post- combustion	Applicable to any combustion system Large experience for CO ₂ capture applications in power generation (pilot and full scale commercial plants) Downtime of CO ₂ capture plant does not result in loss of LNG production if this can continue with untreated flue gas vented to atmosphere (current practice) Route of choice for current CCS projects	Post-combustion CO ₂ capture plants use more solvent and produce more solvent residue The major source of energy requirement is the use of heat to regenerate the solvent Absorption processes require large plot areas; large columns (both in height and cross-sectional area), large solvent regeneration duties and cooling duties
Pre- combustion	Typically removes 85-90% of the CO ₂ Based on well-known and proven technologies used in chemicals and refinery industries to produce hydrogen rich gas Gas turbine technology has been tested and developed to use hydrogen as fuel as an alternative to hydrocarbon fuel Produced synthesis gas is rich in CO ₂ and at high pressure, making separation of the CO ₂ less energy intensive (using physical solvents) and the separation equipment more compact Requires less energy for CO ₂ compression as some of the CO ₂ is recovered at elevated pressure	Using of hydrogen rich gas directly in large scale power production is a new application, involving complex gas turbine technology The fuel gas system producing hydrogen rich gas involve significant additional complexity and process equipment directly impacting the design, operation and performance of gas turbines and liquefaction process Loss of hydrogen rich gas production would cause a loss of LNG production. Loss of flexibility in fuel gas system to handle EFG and BOG Economic feasibility may depend on economies of scale and the possibility to export hydrogen rich gas produced in excess of the liquefaction plant requirements Main issues are the integration, operability and reliability of plants Efficiency reduction as a result of energy losses due to shift conversion Requirement to add steam to the shift converter feed, and the need for catalysts

Table 4. Evaluation of CO₂ Capture Routes



Route	Advantages	Disadvantages
		Oxyfuel combustion of natural gas in gas turbines is under development and has not been tested on a commercial scale
		The fuel gas system and gas turbines involve significant additional complexity and process equipment
Oxyfuel	Typically removes 90-97% of the CO ₂ Produces CO ₂ at high purity from the combustion process No need for CO ₂ capture process and solvent regeneration Air separation plant for oxygen generation is a mature technology based on standard designs Oxyfuel boilers could be applied to generate steam used for power generation (including electric drive of cycle compressors) or heating (mainly solvent regeneration)	 Existing gas turbine drives may be difficult to retrofit, and feasibility may only be possible in the long term, but certainly involving a technical risk Efficiency losses are due to the electricity used by the oxygen production unit, with the cost of oxygen production in sufficient quantity representing a major operating cost Impact on thermal efficiency is potentially larger than post-combustion Loss of oxygen production would cause a loss of LNG production A higher volume of gas is fed to the CO₂ compressors due to the presence of impurities Purification of the CO₂ may be required to remove excess oxygen and impurities (nitrogen, argon) Additional power is necessary to drive the (low temperature) separation unit, removing impurities

Table 4. (Continued)

Techno-economic analyses comparing the three capture routes in power generation plants have been found to be remarkably similar. Capital costs, cost of electricity and thermal efficiencies of the power plant with capture schemes have been found to be comparable. These are subject to a number of uncertainties, including the fact that limited or no information is available on plants being built and operated in full scale commercial plants, particularly for pre-combustion and oxyfuel routes.

When compared to pre-combustion and oxyfuel combustion, post-combustion capture is considered to provide comparable performance with reduced technical risk and process complexity. The significant experience from pilot tests, front end engineering and design studies, and construction and operation of large-scale commercial plants provide certainty in the feasibility and economic evaluation.

The overall CCS scheme would see the CO_2 captured in the post-combustion route (from gas turbine flue gases) to be compressed and purified together with CO_2 that is removed from the reservoir feed gas in the pre-treatment part of the liquefaction train.



9.0 CO₂ CAPTURE TECHNOLOGIES

9.1 BASIS FOR TECHNICAL EVALUATION OF APPLICABLE TECHNOLOGIES

For initial implementation of CCS schemes in LNG plants, post-combustion has been identified as the most appropriate route to capture the CO_2 produced by fuel gas combustion. The technical evaluation, selection and design of the post-combustion CO_2 capture technology appropriate for LNG liquefaction plants would require consideration of the particular flue gas flows and composition, CO_2 removal and product purity. The following defines the basis of design for a potential functional specification.

9.1.1 Flue Gas Pressure

Flue gas available at atmospheric pressure at the flue gas stack.

9.1.2 Flue Gas Temperature

Exhaust temperatures reported for gas turbines is in the order of 500°C. It is expected that waste heat recovery is in place, reducing the temperature of flue gases to approximately 200°C which will require further cooling based on required temperature for the capture plant.

9.1.3 Flue Gas Composition

The range of composition of flue gases produced by gas turbines can be considered to be relatively small, particularly when gas turbine technology is limited to those typically used in LNG liquefaction plants (see Figure 2). This is due to the large air excess volumes drawn into the gas turbines (see Table 6) in excess and well above the minimum stoichiometric ratio required for combustion of fuel gas. Therefore, variability of fuel gas composition has a relatively minor impact on exhaust gas compositions. This results in oxygen content typically in the order of 12-18 mol%. The excess air dilutes the combustion products, with CO₂ content in the order of 1-5 mol%. Final water content depends on the humidity of ambient air.

Expected flue gas compositions based on fuel gas combustion calculations have been produced (see Figure 10) on the following basis:

- Fuel gas composition considering representative feed gas (see Appendix 7)
- Representative air humidity and ambient temperature (see Appendix 7)
- Gas turbine performance, with fuel gas requirement based on output power and efficiency (for turbine type) and air flow from reported gas turbine mass flows.

As noted before, CO_2 in the exhaust gas is diluted due to the excess air drawn into the gas turbine compressor. The calculated CO_2 composition is in the order of 3 mol%, with oxygen being around 14 mol%. This composition is taken as representative of typical baseload LNG liquefaction plants and can be used as the basis for technology selection.





Figure 10. Range of Flue Gas Compositions

Notes:

- 1. Representative ambient conditions considered (see Appendix 7): average ambient conditions 21°C and 74.7% RH, maximum ambient 43°C and 90% RH.
- Representative gas turbines for estimate of air flow are PGT25+ and MS5002D. These have the minimum and maximum flow of excess air respectively, out of those gas turbines typically considered for LNG service.
- 3. Representative fuel gas compositions are included in Appendix 7.

Typical gas turbine exhaust emissions from combustion of conventional gas fuels (taken from GE Power Systems Specification GER-4211) is shown in Table 5.

Table	5.	Typical	Gas	Turbine	Exhaust	Trace	Emission	IS

Component	Range
Nitric Oxide (NO)	20 – 220 ppmv
Nitrogen Dioxide (NO2)	2 – 20 ppmv
Carbon Monoxide (CO)	5 – 330 ppmv
Sulphur Dioxide (SO ₂) Note 1	Trace – 100 ppmv
Sulphur Trioxide (SO3) Note 1	Trace – 4 ppmv
Unburned Hydrocarbons	5 – 300 ppmv
Particulate Matter Smoke	Trace – 25 ppmv

Notes:

1. Sulphur oxides depend on sulphur content in fuel gas



9.1.4 Flue Gas Flows

Flue gas flows and compositions are dictated by the particular ambient and process conditions and plant configuration, which are different for each liquefaction plant and include:

- Feed gas conditions and composition
- Feed gas pre-treatment requirements
- Liquefaction process technology
- Gas turbine technology selection
- Compressor technology selection
- Refrigerant compressor train configuration
- Location of LNG plant and ambient conditions
- Heat and power integration
- Fuel gas composition and variability

All of these factors have an ultimate effect in the overall thermal efficiency of the liquefaction train, which defines the quantity of fuel gas that is required and the associated emissions. Flue gas flow rates are dependent on the achieved thermal efficiency (as a result of a combination of the factors above) and the capacity of the liquefaction train. Therefore, LNG plant capacity and efficiency will dictate the required size of a CO_2 capture plant.

For typical baseload liquefaction plants with capacities of around 5 mtpa of LNG, the required capacity of the CO_2 capture plant will be in the order of 3000 tonnes of CO_2 per day, equivalent to about 1 mtpa of CO_2 (see Table 2 and Table 3). Relatively large flue gas flows need to be processed, indicatively in the order of 1500-2000 MMSCFD (see Table 7).

The reported exhaust flows for gas turbines are shown in Table 6. It is noted that the exhaust flows for industrial turbines is about double the flows for the power equivalent aeroderivative gas turbines.

Coo Turbino	Exhaust Flow				
Gas Turbine	kg/h	kmol/h	MMSCFD		
PGT25+	303,480	10,677	214		
LM6000	455,400	16,004	321		
Frame 5C	444,240	15,606	313		
Frame 5E	367,200	12,913	259		
Frame 5D	507,600	17,830	358		
Frame 6	504,000	17,738	356		
Frame 7	1,087,200	38,232	768		
Frame 9	1,515,600	53,311	1,070		
LMS100	740,160	26,091	524		
RB211 DLE	337,680	11,875	238		
Trent 60 DLE	555,480	19,517	392		

Table 6. Gas Turbine Exhaust Flows



Table 7 shows the range of flue gas exhaust flows estimated for the given reference liquefaction plant capacities is given in Table 3 (only for gas turbine drives).

LNG Liquefaction	Flue gas flow (MMSCFD)			
(mtpa of LNG)	Aeroderivative	Industrial		
3.5	1000 – 1300	1500 – 2000		
4.5	1300 – 1700	1900 – 2500		
5.5	1600 – 2100	2300 – 3100		
8.0	-	2700 – 3600		

 Table 7. Flue Gas Exhaust Flows for given LNG Liquefaction Plant Capacity

Notes:

- 1. The range of flows is for liquefaction power specific in the range 0.3 0.4 kWh/kg.
- 2. Flue gas flows are calculated for combustion of sweet feed gas at average ambient conditions (see Appendix 7 for basis of design).

9.1.5 CO₂ Removal and Product Purity

CO₂ removal is expected to be maximised but without incurring in excessive costs. CO₂ removal efficiencies achieved by capture schemes are at typically 90%. This is to some extent a nominal specification rather than actual technology limits or dictated by environmental regulation or downstream requirements. CO₂ removal can be greater than 90% if the incremental cost is not considered to be excessive.

CO₂ product purity depends on technical factors related to CO₂ transport and storage, including corrosion, hydrate formation and regulatory requirements. A high purity, nominally greater than 99 mol% is expected. Impurity limits are given in Table 8. Water specification can be achieved by a downstream dehydration step. Considering the typical maximum gas turbine emissions (see Table 5) it is not expected that the specifications given in Table 8 will have an impact in technology selection as none of the processes are highly selective for these impurities.

Component	Limit	Notes
Water	50 ppmv	To avoid formation of free water phase and hydrates
Nitrogen and Argon	4 mol%	To reduce the volume for compression, transport and storage, avoid operation in two-phase region and to limit the increase in minimum miscibility pressure in EOR
Oxygen	100 ppmv	Preliminary/indicative, depending on the effects of oxygen in underground reservoirs
Carbon Monoxide	0.2 mol%	
H ₂ S	20 ppmv	Dictated by HSE considerations and pipeline
SOx	100 ppmv	embrittlement)
NOx	100 ppmv	
Hydrocarbons	4 mol%	
Total non-condensables	4 mol%	

Table 8. Impurity Limits in CO₂ Product



9.2 OVERVIEW OF CO₂ REMOVAL PROCESSES

Pre-combustion capture and oxyfuel combustion routes for CO₂ capture have been screened out. Therefore, the following sections will focus specifically on CO₂ removal processes suitable for post-combustion application.

Technologies for CO₂ removal have been widely used in natural gas processing due to the requirement to meet downstream specifications either for further gas processing or for export to pipeline gas networks.

 CO_2 removal from flue gas is aimed at removing a large proportion of CO_2 (typically 90%) for purposes of sequestration to reduce the environmental impact of CO_2 emissions. The estimated flue gas composition from combustion of natural gas in gas turbines contains a relatively low CO_2 content (in the order of 3 mol%) and considerable amounts of oxygen (about 14 mol%). Gas fired boilers use less air, so CO_2 is less diluted with composition in the range of about 5 to 10 mol%

A number of CO_2 capture schemes have been implemented at pilot and full scale in power plants utilising pulverised coal as fuel. These capture schemes are based on absorption on chemical solvents (amine-based). It is noted that the flue gas composition for coal power plants differ to that expected from CCGT power plants. Coal produces a larger amount of CO_2 per unit of fuel than natural gas combustion. In these plants, combustion uses a reduced volume of air compared to gas turbines, such that the oxygen content in flue gases is in the order of 6 mol%. CO_2 content in flue gases is about 11%.

Performance of absorption and adsorption processes improves at low temperatures. The feed gas to acid gas removal units is usually in a close approach to ambient conditions, and no feed cooling is usually required. On the other hand, flue gases are hot, typically around 500°C at the gas turbine outlet or about 100-200°C at the waste heat recovery outlet (if installed) and cooling to near ambient temperature or below is required before the gas is passed to the capture process.

Pressures of the order of 20 to 70 bar are typically seen in acid gas removal feed gas, so use of physical solvents may be feasible. Flue gases are at low pressure, near atmospheric. Although some form of pressure boost is typically installed, this is only sufficient to overcome the pressure drop in the capture process. The use of physical solvents that require high partial pressures of CO_2 may require further compression to achieve a performance comparable to that achieved by chemical absorption.

In addition to low pressure leading to relatively low flue gas densities, gas turbines use large volumes of excess air. Volumetric flowrates, therefore, are typically much higher for CO₂ capture from flue gas than in acid gas removal processes, which lead to significantly larger equipment sizes.

Technologies at various levels of maturity are available for post-combustion CO_2 capture schemes, including absorption, adsorption, cryogenic and membranes. A review of these technologies is included in Appendix 6. The technologies are screened based on their applicability to post-combustion CO_2 capture from flue gas, in particular the effect of CO_2 composition in flue gas, CO_2 product purity, process conditions and flows. This approach highlights relevant technologies and allows the merits of a number of technologies to be identified and evaluated against requirements to eliminate options that are not suited to this application or cannot be economically applied. Key factors that are considered in the initial screening include technical relevance/applicability, CAPEX, OPEX and track record of operating plants.



9.3 TECHNOLOGY SCREENING AND SELECTION

The merits of a number of technologies are identified and evaluated against requirements to eliminate options that are not suited to this application or cannot be economically applied. Key factors that are considered in the screening include applicability against the post-combustion capture process duty (flue gas flows, compositions and process conditions), CAPEX, OPEX and technology maturity. The majority of the considerations apply to technology selection for post-combustion CO₂ capture from gas fired power plants, already carried out elsewhere (e.g. Ref 49). Screening and selection also takes into account particular considerations for implementation in LNG baseload liquefaction plants. Key considerations for each technology are summarised in Table 9.

Chemical absorption processes are very well established in CO₂ removal from gases. These processes are well known in LNG plants as these are used for acid gas removal in feed gas pre-treatment. Amine-based and ammonia-based processes have been specifically developed, extensively tested and commercially available for post-combustion CO₂ capture.

Amine-based capture processes (either using generic aqueous solutions of MEA or proprietary amine-based solvent formulations) are concluded to be the most appropriate for the application, particularly due to the technical readiness for utilisation on CO_2 capture schemes of this scale. This technology has been extensively developed, tested at pilot scale and the only one demonstrated on large scale schemes (see Appendix 4) which are similar in size to this application.

The performance of amine-based capture processes is acceptable against the requirements, proven to be capable of selective removal of CO_2 from flue gas with low CO_2 concentrations to produce high purity CO_2 . The process operates at low pressures, so compression of flue gas is not required, and mechanical design of equipment allows reduction in capital costs. Large equipment is required but this is associated to the large volumes of flue gas in post-combustion capture (and hence applicable to all technologies).

For amine-based capture processes, a large energy requirement remains associated to solvent regeneration, but this can be minimised by the use of proprietary solvents. Furthermore, there is potential for the regeneration duty to be provided by waste heat recovery, which in an LNG plant is expected to be available in excess (see Appendix 3) when compared to a CCGT (where it has been already used for power generation). The impact of installing waste heat recovery in the performance of gas turbines and reduction in LNG capacity needs consideration (see Section 6.6). Solvent degradation occurs but this is minimised by using proprietary solvents and established reclamation processes are available. Solvent losses are minimised by water wash.

Ammonia-based capture processes also benefit from extensive development and testing, and the process is available on a commercial basis, with performance comparable to aminebased processes. Therefore, ammonia-based processes could be potentially considered for CO₂ capture in LNG plants. However, under the current scenario, implementation in LNG plants would see better probabilities of success after this technology is demonstrated in a full scale plant (likely to be a coal fired power plant) and clear benefits over amine-based processes are established and demonstrated. This demonstration would have to tackle concerns on ammonia losses and perceived process complexity. The hot potassium carbonate process is a mature technology widely used in gas processing, but it is generally more economical for CO_2 concentrations higher than what is anticipated in this application. Furthermore, the entire flue gas volume would need to be compressed from atmospheric pressure to a high pressure (~20 bara), which would require a large compression system which may result technically and economically unfeasible.

The caustic wash process is considered not to be ideally suited to this application due to the non-regenerable nature of the solvent which leads to very high caustic consumption for the estimated CO_2 capture rate.

Absorption using amino acids is an alternative method to avoid problems typically associated with amines like corrosion and solvent losses and degradation. However, it is considered that these cannot be compared with amines in terms of the technology maturity, and that a number of amine-based formulations are commercially available to minimise the impact of the perceived problems to an acceptable level.

Physical absorption relies upon a high partial pressure of CO_2 and solvents have low CO_2 loading capacity. Even with compression of the full flue gas flow (assuming that technical and economic feasibility can be confirmed), physical solvents would not be expected to achieve comparable performance to chemical absorption unless CO_2 concentration is increased (e.g. by EGR techniques).

Absorption processes using hybrid solvents and ionic liquids have a low maturity level (especially on post-combustion capture applications) and high cost compared to amine systems.

Physical adsorption technologies, for example using molecular sieves, have been used for CO_2 removal from natural gas. These processes are generally best suited for low gas flows, with low concentrations of CO_2 (in the order 2 mol%) with the purpose of achieving cryogenic specifications (50 ppmv of CO_2). TSA processes are feasible for plants such as peak-shaving facilities because the CO_2 rich regeneration gas is either used as fuel or re-injected in the gas pipeline network. Similar options for the handling/disposal of regeneration gas are not possible in the treatment of flue gases. PSA processes offer an alternative for adsorbent regeneration via desorption of CO_2 and production of a highly pure CO_2 product. However, these processes would require compression of the full flue gas flow, which impacts the capital and operating costs.

An adsorption process has been specifically developed for post-combustion CO_2 capture utilising fluidised beds. This process is under development at pilot scale (200 tonnes of CO_2 per day). This may represent an option with potential in the long term when the required technology maturity is reached. This would require successful demonstration (particularly at full scale) of its technical feasibility and techno-economic benefits against the established amine-based absorption processes.

This low concentration of CO_2 in the flue gas combined with the large volumes of flue gas to be treated provides the basis to discount cryogenic processes. These are not adequate as a very large refrigeration duty and associated size of equipment would be required.

The use of membranes can also be discounted on the basis of the low CO_2 concentration plus the need to compress the full flue gas flow. If used, the contribution of membranes would be in concentrating CO_2 at the expense of large membrane areas and high recompression duties, with an additional process (likely to be amine-based absorption) being required to produce CO_2 with high purity.



Technology		Advantages	Disadvantages	Applicable
Absorption	Chemical	 Applicable to selectively remove CO₂ from gas with low CO₂ content, at low pressure (near atmospheric) and delivering product with high purity. 90% capture achievable at acceptable plant size and costs Mature technology commercially available, extensively tested for post-combustion CO₂ capture at pilot scale and implemented in large scale power generation plants, with capture plants having similar capacity to that expected for LNG plants Proprietary solvent formulations are commercially available with reduced regeneration energy demands, solvent degradation and corrosion LNG plants familiar with amine-based processes Availability of heat (from waste heat recovery) and power generation (from spare installed capacity) expected in LNG plants 	 Need for solvent handling (storage, pumping, reclaiming) Thermal regeneration of absorbent required Solvents can be subject to thermal degradation and in the presence of oxygen Relatively large equipment count and high energy requirements for regeneration of solvent Corrosion of carbon steel require use of stainless steel Waste produced from solvent reclamation need disposal CO₂ product delivered at low pressure (near atmospheric) and water saturated, which requires dehydration and compression. Applicable solvents are amine and ammonia based. Hot potassium carbonate and caustic wash excluded. 	Yes (short term)
	Physical	 For high CO₂ partial pressures, solvent regeneration energy requirements for some physical solvents can be low in comparison to chemical absorption processes Use of lonic Liquids with modified structure to suit CO₂ removal duty 	 Need for solvent handling Thermal regeneration of absorbent required Most suited to high partial pressure of CO₂ in feed gas above 3.5 bara. This would require compression of large flue gas volumes Low CO₂ loading capacity in low CO₂ partial pressure High viscosity of lonic Liquids and lack of maturity and commercial availability. 	No
	Hybrid	 Can potentially offer the benefits of both chemical solvents (higher CO₂ removal at low partial pressures) and physical solvents (lower energy requirements) 	 Thermal regeneration of absorbent required Need for solvent handling Typically, more expensive than conventional amines CAPEX and OPEX tend to be higher than amine processes 	No
Adsorption		 Used for CO₂ removal in gas processing Widely available adsorbents, good thermal stability and low sensitivity to moisture. Amine based adsorbents have lower heat of regeneration compared with aqueous amines (absorption) Fluidised bed process (potassium carbonate) being tested at pilot scale for the specific application on post-combustion CO₂ capture 	 Regeneration of adsorbent required at low pressure (PSA) or high temperature (TSA) PSA would require compression of flue gases TSA requires handling of rich CO₂ regeneration gas Low CO₂ removal capacity, which makes them best suited to low CO₂ content and gas flowrates. Not available commercially. It would require demonstration at full scale. 	Yes (long term)
Membranes		 Faster start-up, fewer equipment items and lower maintenance requirements in comparison with solvent systems No regeneration, handling or disposal of solvents required CO₂ produced is not water saturated so no need for downstream dehydration Suited to plants in remote locations due to complexities in delivery of solvents and possibility for unmanned operation 	 Pre-treatment of the feed may be required to prevent damage of expensive membrane It would require compression of large flue gas volumes as a first step High CO₂ recovery involves high recompression power requirements and large membrane areas (and associated footprint), increasing both CAPEX and OPEX Best suited for bulk CO₂ removal to feed gas with high CO₂ concentrations of 10 mol% or more Membranes cannot produce high CO₂ purity on their own 	No
Cryogenic		 A liquid CO₂ stream at high pressure can be produced, reducing power requirements for CO₂ product compression. Well known technology within other applications, particularly processing of gas with high CO₂ content. 	 Need for pre-treatment to remove water and hydrocarbons that could freeze Process efficiency is low at low feed gas CO₂ concentrations, best suited to gas with >20 mol% CO₂ content Elevated feed gas pressures are required to improve performance High energy consumption for refrigeration system, provided as electricity Operating conditions and removal limitations to avoid CO₂ freezing 	No



10.0 SUMMARY AND RECOMMENDATIONS

10.1 SUMMARY

LNG baseload liquefaction plants use well established liquefaction technology, with a large proportion of these plants using a propane precooled mixed refrigerant process. Most of the plants use gas turbines to drive the refrigerant compressors, with recent plants using industrial Frame 7 (typically 2 per train) and Frame 5 (typically 6 per train) and their aeroderivative equivalents, resulting in typical train capacities in the order of 4-5 mtpa of LNG. Total power output is in the order of 200 MW. A step change is seen with large trains having a capacity in the order of near 8 mpta using Frame 9 gas turbines (3 per train).

 CO_2 emissions from combustion equipment depend on the thermodynamic efficiency of the processes, dictating the fuel gas that needs to be burned to produce the required power. This is a function of a number of variables including feed gas pressure and composition (dictating pre-treatment requirements), ambient conditions, liquefaction technology, gas turbine technology and process equipment. Liquefaction specific power considering technology and local factors is typically in the range 0.3 to 0.4 kWh/kg for baseload liquefaction technology (~0.3 kWh/kg reported as typical). This leads to specific CO₂ emissions typically in the range 0.20 to 0.28 tonne of CO₂ per tonne of LNG for industrial gas turbines and 0.15 to 0.21 tonne of CO₂ per tonne of LNG for aeroderivative gas turbines.

 CO_2 is also emitted by acid gas removal processes in the pre-treatment of feed gas prior to liquefaction. CO_2 from this process can be considered to be already captured. The separated CO_2 is typically vented, particularly if the content in feed gas is relatively low (indicatively in the order of 2 mol% or less). Where feed gas contains significant CO_2 content, considerations may need to be made for sequestration rather than venting in response to environmental constrains.

Based on the nature of the LNG industry, only well proven technologies are expected to be considered for CO_2 capture in LNG plants, with schemes that minimise the risks of disruptions to LNG production. Post-combustion CO_2 capture from gas turbine emissions using well proven chemical absorption technology is likely to be the preferred route.

When compared to pre-combustion and oxyfuel combustion, post-combustion capture is considered to provide comparable performance with reduced technical risk and process complexity. Post-combustion can be installed without affecting the performance of the core liquefaction process. It requires a minimum number of modifications to existing equipment, hence reducing risk. This makes post-combustion attractive for new LNG plants or as retrofit to existing plants.

The post-combustion technologies with the highest potential for immediate implementation are chemical absorption processes. These are proven technologies with the main disadvantage being the energy requirements to regenerate the solvent. However, there is potential for the heating duty to be provided by waste heat recovery, which in an LNG plant is expected to be available in excess. Perceived problems such as solvent degradation, solvent volatility and losses, corrosion, etc., represent operational challenges that can be managed within acceptable limits using solvent formulations commercially available.

The range of composition of flue gases produced by gas turbines can be considered to be in a relatively small range. This is due to the large air excess volumes drawn into the gas turbines, with variability of fuel gas composition having a relatively minor impact on exhaust gas compositions. CO_2 content is in the order of 3 mol%, with oxygen being around 14 mol%.



For typical baseload liquefaction plants with capacities of around 5 mtpa of LNG, the required capacity of the CO_2 capture plant will be in the order of 3000 tonnes of CO_2 per day, equivalent to about 1 mtpa of CO_2 . This is comparable to existing full scale capture plants (e.g. Boundary Dam and Petra Nova), so a similar plant size and associated investment is expected. Means to reduce capital cost, strengthen the business case, minimise project risks and in general to improve the economics of the project will contribute to the feasibility of the CCS scheme.

10.2 RECOMMENDATIONS FOR FURTHER WORK

The recommendations for further work presented here refer to knowledge gaps where research and development efforts can be directed in the context of facilitating technoeconomic evaluation, demonstration of feasibility and eventual implementation of CCS schemes in LNG baseload liquefaction plants.

- Potential options to tackle the challenge to reduce energy required by the CO₂ capture (particularly associated to solvent regeneration) should remain the main focus for research and development. These options include:
 - Development of new capture agents (e.g. solvents and adsorbents) with lower energy requirements for CO₂ desorption. These agents are not limited to aminebased solvent formulations and amine-impregnated adsorbents, but potentially including different reacting agents.
 - Development of highly efficient processes and equipment. This includes novel regeneration schemes and hybrid processes combining the advantages of technologies to carry out the right processing step (e.g. use of technologies concentrating CO₂ prior to capture). Also, energy integration schemes recovering waste heat whilst minimising the impact on existing production processes must be considered.
- The strategies above may provide with incremental improvements, and the benefits in terms of cost reductions may be lost when put in the overall context of the full CCS scheme economics. Therefore, priority should be given to the development of alternative technologies, particularly those able to provide with a step change in performance and overall economics.
- Some technologies reviewed show good potential for potential consideration in the long term, but at present, these cannot be considered to be fully suitable for LNG plants due to their relatively low level of maturity. For example, adsorption processes may have acceptable CO₂ capture performance with reduced energy requirements, but the technology remains at the demonstration scale. Some other technologies like ammonia-based processes have undergone extensive testing at pilot scale, and demonstration at full scale (in power generation applications) will be required for consideration for CO₂ capture in LNG plants. Demonstration not only refers to proving technical feasibility, but includes lessons learned in an extended context including project economics, financing, project execution and operating experience.
- Pre-combustion and oxyfuel combustion are alternatives to reduce the processing requirements associated with post-combustion capture (large plant size and energy requirements). For implementation in LNG plants, consideration of these options will require full demonstration of commercially available processes and equipment, particularly as these processes will be fully integrated with the LNG production process.

Key issues include reliability of equipment (particularly gas turbines), flexibility of gas turbines to handle a range of fuel gases (hydrogen rich gas and hydrocarbon gas fuels with different content of ethane+ components and nitrogen), operating availability of the whole process and control and operation complexity.

- The power required for compression of the CO₂ product is considerable. Strategies to reduce compression power (e.g. operation of stripper column in AGRU at elevated pressure) will have an impact in capital and operating cost. Reduction in compression duty may also facilitate compressor selection and confirm technical feasibility. Alternative schemes consisting of compression followed by liquefaction and pumping could be evaluated to match available machinery.
- There appears to be potential for improvement of the thermal efficiency of the overall liquefaction process through the use of electric motor drives with a centralised combined cycle power generation facility. This option has already been considered in evaluations and designs, but liquefaction technology remains linked to proven gas turbine drives. Consideration for electrical drives may be incentivised by the need to reduce emissions dictated by environmental permitting. Further consideration will be beneficial, particularly as implementation of post-combustion capture may be facilitated when this is implemented in the centralised power generation plant as opposed to numerous different locations.
- EGR technology presents an opportunity to provide incremental benefits to CO₂ capture, but at present this is considered not to be mature enough for use in this application. Development is required with particular focus in the suitability of gas turbines in LNG service operating with an EGR. It is noted that implementation of EGR has a direct impact on capital cost reduction, with benefits maximised for operation at part load. Furthermore, implementation involves a degree of integration with the LNG plants as gas turbine modifications are required, and the benefits need to be established in order to justify the risks involved. The feasibility for implementation, risks and realistic benefits for baseload LNG plants need to be established by further studies.
- Implementation of CCS in LNG will be benefited from detailed and up to date studies on issues that are applicable to CCS in general such as lessons learned from pilot plants and particularly full scale plants, cost reduction for nth of a kind plants, potential business case schemes, funding schemes and financing options, trends in energy prices and environmental policy, etc.
- To date, CCS schemes have only been implemented in LNG gas plants where the CO₂ content in feed gas is relatively high, so a significant gain in reducing emissions can be achieved by sequestrating the CO₂ that is already separated in the plant. No CO₂ capture plants from fuel gas combustion processes have been installed to date. In addition to demonstrating the technical feasibility, understanding of the possible scenarios and drivers that may make the CCS schemes economically feasible is also a key requirement and should be part of further studies. Particular focus should be put in the investigation of overall project economics scenarios where CCS is driven by a commercial justification beyond the delivery of CO₂ product i.e. oil revenues from EOR. Further studies should assess the potential (e.g. LNG plants in the vicinity of oil fields benefiting from EOR) and implications of extending the commercial scope (e.g. oil field development licensing, partner and commercial agreements and risks).



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APPENDIX 1. BASELOAD LIQUEFACTION TECHNOLOGIES

A1.1. PROPANE PRE-COOLED MIXED REFRIGERANT (AP-C3MR™)

The propane pre-cooled mixed refrigerant (AP-C3MR[™]) process licenced by Air Products and Chemicals Inc. (APCI), is the leading liquefaction process accounting for approximately 87% of the world's baseload LNG capacity.

Feed gas is pre-treated before being pre-cooled and partially condensed against three or four levels of propane in the propane evaporators. A MR typically consisting of nitrogen, methane, ethane and propane is used for the main liquefaction and sub-cooling duty, allowing an efficient matching of the overall feed gas cooling curve and optimisation of the production rate to maximise output for seasonal ambient temperature changes.

MR compressed to a high pressure is cooled against air or seawater and subsequently cooled and partially condensed against all the stages of propane before being separated into vapour and liquid fractions. The vapour fraction is condensed and sub-cooled in the main Coil-Wound Heat Exchanger (CWHE) and then flashed across a Joule-Thomson valve, before returning to the cold end of the CWHE to provide the sub-cooling duty. The low-pressure MR vapour from the warm end of the exchanger totally vaporised and slightly superheated is then recompressed over several stages with intercooling and aftercooling.

The AP-C3MR[™] process has been designed and operated with three and four Frame 5 gas turbines and combination of Frame 6 and Frame 7 with various starter/helper motor arrangements and all sizes of steam turbines. Process designs have also considered aeroderivative gas turbines. Use of large industrial gas turbines has been considered to achieve significant economy of scale gains. For this further increase in capacity the Split-MR drive configuration is used to allow full utilisation of two Frame 7 gas turbine shaft. The LP and MP mixed refrigerant compressors are on the same Frame 7 gas turbine shaft. The HP mixed refrigerant compressor is on driven by the other Frame 7 gas turbine on the same shaft as the propane compressor.

Generally, AP-C3MR[™] has a capacity of 4-5 mtpa (5 - 5.5 mtpa for a single train based on two Frame 7 drivers). This is increased by use of the Split MR configuration to 6 mtpa (based on two Frame 9 drivers).

A1.2. AP-X®

The AP-X[®] process (licenced by APCI) is essentially the AP-C3MR[™] process in combination with a nitrogen expansion cycle. The propane cycle is used for pre-cooling, single light MR for liquefaction and nitrogen for sub-cooling. In the nitrogen cycle at the back end of the liquefaction process, high-pressure nitrogen is cooled by MR and cold returning streams, then expanded in a turbo-expander to sub-cool the LNG. Whereas, nitrogen expander cycles have an efficiency disadvantage relative to pure or mixed hydrocarbon refrigerant cycles when providing with refrigeration for liquefaction, they can be efficient at providing refrigeration at sub-cooling conditions.
The AP-X[®] process has been developed in response to demand for increased LNG train capacity and lower unit cost. It could be used to expand capacity on existing AP-C3MR[™] plants from its current limit 5 mtpa (typical, based on size limits of existing cryogenic exchangers and compressor frame sizes) to 8 mtpa with the addition of the nitrogen cycle and without the need to use duplicate/parallel equipment.

The power balance between the propane, mixed refrigerant and nitrogen cycles, together with the Split MR machinery configuration allows flexibility in matching available compressor driver sets. A number of driver options has been considered including three Frame 7 gas turbines (two on AP-C3MR[™] with Split MR configuration, one on the nitrogen cycle), two Frame 9 gas turbines (the preferred option to minimise equipment count), three Frame 9 gas turbines (two on AP-C3MR[™] with Split MR configuration, one on the nitrogen cycle with export of surplus power) and electric motors.

The AP-X[®] process has a capacity in the range 6.5 - 10 mtpa (7.5 mtpa for a single train based on two Frame 9 drivers, 10 mtpa for a single train based on three Frame 9 drivers).

A1.3. OPTIMIZED CASCADE®

The Optimized Cascade[®] process (licensed by ConocoPhillips) uses three pure refrigerant cycles i.e. propane, ethane (propylene and ethylene may also be used) and methane that are optimally cascaded in sequence. The performance across the cooling curve is optimised in function of the number of cooling stages used and where they are located in the process.

After pre-treatment, the feed gas passes to the propane stage for pre-cooling and partial condensation in multi-stream Plate Fin Heat Exchangers (PFHE). Next the feed gas passes to the ethane cold box where it is further cooled, condensed and slightly sub-cooled. The propane and ethylene refrigerant cycles are arranged in closed loops. Propane is flashed to three different pressure levels whilst ethane is flashed at two or three. Increasing the number of levels would give a closer match to the cooling curve, but this may be offset by the increased capital costs (CAPEX) of an additional pressure stage.

The condensed feed gas is then sent to a scrub column to remove heavies and NGL from the stream. Finally, the stream is sent to the methane cold box. The methane cycle operates in an open loop arrangement, with the methane refrigerant supply being provided by the expansion of condensed gas to the final LNG storage pressure. The methane vapours compressed and mixed with the feed gas upstream of the ethane cold box or used as fuel.

The plants built to date make use of Frame 5 gas turbine drivers. Multiple gas turbines and compressors are configured in parallel in a single train. When one gas turbine or compressor in a refrigerant cycle trips, the parallel compressor continues to run, allowing the plant to maintain production at a reduced rate, with load being balanced between the three refrigeration loops to maintain 60-75 % of total plant capacity.



A1.4. MIXED FLUID CASCADE (MFC®)

The MFC[®] process (licensed by Linde) is essentially a three-cycle mixed refrigerant cascade process using a combination of PFHEs and CWHEs. The mixed refrigerant in the pre-cooling section allows lower temperatures than a propane cycle. The liquefaction and sub-cooling duties are split into two further independent mixed refrigerant cascade cycles, which are designed to give a closer match to the natural gas cooling curve. The pre-cooling cycle uses a blend of ethane and propane, the liquefaction cycle operates with a blend of methane, ethane and propane and the sub-cooling cycle uses a lighter mixture of methane, ethane and nitrogen.

Dry feed gas from the pre-treatment section is cooled against the pre-cooling mixed refrigerant. The vapour is routed via an expander to a scrub column separating a major part of the heavier hydrocarbons from the gas using predominantly the de-ethaniser overhead stream from the Fractionation unit. Scrubber bottoms are fed to the Fractionation unit. The gas stream is then recompressed, liquefied and sub-cooled in two separate CWHEs.

The refrigeration system design utilises five compressors, two on the pre-cooling circuit, one on the liquefaction circuit and two on the sub-cooling circuit. The Snøhvit plant uses electric motors as compressor drivers.

A1.5. DUAL MIXED REFRIGERANT (DMR)

The DMR process (licensed by Shell) is a development of the AP-C3MR[™] process but uses two separate mixed refrigerant cycles which are cascaded to achieve a better match to the natural gas cooling curve; one for pre-cooling and one for liquefaction and sub-cooling. It makes use of multi-stream heat exchangers, usually CWHEs, although the use of PFHEs is possible. Designs are usually offered with two Frame 7 to reduce the equipment count and cost.

A1.6. LIQUEFIN™

The Liquefin[™] technology (currently licensed by Air Liquide) consists of two mixed refrigerant cycles, one for pre-cooling (using ethane and propane) and the other for liquefaction and sub-cooling (using a lighter blend of methane, ethane, propane and nitrogen). Capacities up to 6 mtpa can be reached with two Frame 7 gas turbines. Parallel configurations with four Frame 6 or the equivalent aeroderivative gas turbines or large electrical motors have also been studied. Although the process is widely referenced and considered in technology evaluations, no plant has used this technology to date.

A2.1. LIQUEFACTION TECHNOLOGIES

As of January 2017 the total global LNG production capacity stood at about 340 mtpa (Ref 5). A number of projects started commercial operations during 2016 in Australia and the United States (US). Asia Pacific LNG Train 1 and 2 (Australia), Golar LNG Train 1 and 2 (Australia) and Sabine Pass Train 1 and 2 (US) all started production utilising the Optimized Cascade® liquefaction technology. Gorgon LNG Train 1 and 2 (Australia) achieved completion during 2016 using the AP-C3MR[™]/Split MR liquefaction technology. The completion of these projects added 35 mtpa to the total global LNG production capacity.

LNG production capacity per train peaks at 7.8 mtpa in Qatar. This large scale capacity was first brought online in 2009 as part of the Qatargas II (Train 1 and 2) and RasGas III projects which utilise the AP-X® process. The configuration uses three Frame 9 gas turbines with and the excess power being exported (Ref 10). Before this, the global maximum capacity per train stood at 5.2 mtpa at Atlantic LNG (ALNG) in Trinidad utilising the Optimized Cascade® process.

A number of other projects such as Petronas MLNG Train 9 in Malaysia scheduled to reach commercial operations by end 2016 added a further 9.75 mtpa capacity to the global LNG network. Based on announced start dates, by 2017 the total global LNG production capacity will stand at approximately 350 mtpa, giving approximately 17% increase in capacity compared to the start of the previous year. A further 95.5 mtpa is scheduled to be online by end 2019.

About 53% (based on nameplate capacity) of the additional capacity currently under construction (105.3 mtpa) is located in the US, with 26.6% located in Australia, the balance located in Russia (16.5%), Malaysia (6.3%), Cameroon (2.4%) and Indonesia (0.5%).

Liquefaction technologies implemented in current plants are provided by APCI, ConocoPhillips, Linde and Shell. Over 70% of plants currently in operation use technology provided by APCI of which over three quarters are based upon the AP-C3MR[™] process and the remainder use in roughly equal shares the AP-C3MR[™]/Split MR and AP-X® processes. Roughly 20% of liquefaction facilities use technology provided by ConocoPhillips and the remainder at present comprise of Linde and Shell technologies. The vast majority of technologies are able to provide a capacity in the range of approximately 2.5 to 5 mtpa, with the global average standing at approximately 3.8 mtpa.

The AP-C3MR[™] technology has been used since the 1970s providing capacity per train in the range of 1.2 to 5 mtpa with an average of 3.2 mtpa, while the AP-C3MR[™]/Split MR option is able to provide a slightly higher capacity at 5.2 mtpa with the average train capacity using this technology standing at 4.3 mtpa. The AP-X® technology was first implemented in Qatar in 2009 and is able to provide a capacity per train of 7.8 mtpa. It has been utilised six times in total, all of which on projects in Qatar such as Qatargas II/III/IV and RasGas III.

The ConocoPhillips Optimized Cascade® technology provides capacity per train from 1.5 mtpa up to a maximum of 5.2 mtpa with an average of about 4.3 mtpa.



APCI's market share of liquefaction technologies is set to remain relatively unaffected based upon the plants currently in construction. Roughly 55% of technologies selected for plants currently in construction are opting to utilise APCI's technology, while 23% will utilise ConocoPhillips technology.

Linde's MFC® technology is currently only in operation at one baseload LNG liquefaction plant, providing capacity per train of 4.2 mtpa at Snøhvit LNG.

Shell's DMR technology is only in use at Sakhalin 2 (Train 1 and 2) in Russia providing a capacity per train of 4.8 mtpa. A process developed by Shell based on the C3MR process is used at Pluto LNG in Australia giving a train capacity of 4.3 mtpa.

A2.2. REFRIGERATION CYCLE COMPRESSOR DRIVES

Kenai LNG, the world's first LNG liquefaction facility starting operations in 1969, utilised Frame 5 industrial gas turbines. LNG liquefaction facilities continued to utilise industrial type gas turbines such as Frame 5, 6, 7 and 9

In Qatar, the RasGas project which saw the AP-C3MR[™] technology utilised to provide a capacity per train of 3.3 mtpa utilised Frame 7E gas turbines. The RasGas II project which expanded the existing capacity used the AP-C3MR[™]/Split MR process to provide a capacity per train of 4.7 mtpa and utilised Frame 7 gas turbines.

The turbine selected for use in the Darwin LNG project in Australia using the Optimized Cascade[®] process producing 3.6 mtpa was a GE LM2500+ aeroderivative type (Ref 7). This was the first instance of an aeroderivative gas turbine being used in LNG production. Although this type of turbine provided similar horsepower output to a GE Frame 5D industrial gas turbine, the aeroderivative option offered much better efficiency (41.1% compared to 30.3%). Australia Gladstone LNG (GLNG) which uses the Optimized Cascade[®] process and has a capacity of 7.8 mtpa (3.9 mtpa per train) operates using GE PGT25+ G4 aeroderivative gas turbines.

The Wheatstone project, a two train LNG facility with a capacity of 8.9 mtpa (4.45 mtpa per train) using the GE LM6000 gas turbine, is the first commercial use of this type of engine in a mechanical drive application (Ref 8). Various studies have also considered using this gas turbine to drive a propane pre-cooling compressor and a mixed refrigerant (MR) compressor (Ref 6).

The Snøhvit LNG plant uses the Linde MFC[®] and provides a capacity of 4.2 mtpa was the first major LNG liquefaction plant to choose electric motor drives (Ref 9). The development incorporates central power generation using five LM6000 aeroderivative gas turbines providing 230 MW of power to an internal grid. A back-up from the electricity grid provides 50 MW. Three 65 MW electrical motors with variable speed drives act as refrigerant compressor drivers. The design intent is to improve overall thermal efficiency compared to the use of gas turbines in mechanical drive service (Ref 11).

The largest LNG liquefaction trains in the world in Qatar with a capacity per train of 7.8 mtpa. The configuration uses three Frame 9 gas turbines with and the excess power being exported (Ref 10).



A2.3. DATABASE OF BASELOAD LNG PLANTS

Country	Plant	Start-up Year	LNG Capacity (mtpa)	Liquefaction Technology
	Skikda GL1K (Rebuild)	2013	4.5	AP-C3MR™
	Arzew GL1Z Train 1-6	1978	7.9	AP-C3MR™
Algeria	Arzew GL2Z Train 1-6	1981	8.2	AP-C3MR™
	Arzew GL3Z (Gassi Touil)	2014	4.7	AP-C3MR™/Split MR
Angola	Angola LNG Train 1	2013	5.2	Optimized Cascade [®]
	North West Shelf Train 1	1989	2.5	AP-C3MR™
	North West Shelf Train 2	1989	2.5	AP-C3MR™
	North West Shelf Train 3	1992	2.5	AP-C3MR™
	North West Shelf Train 4	2004	4.4	AP-C3MR™
	North West Shelf Train 5	2008	4.4	AP-C3MR™
	Darwin LNG Train 1	2006	3.6	Optimized Cascade®
	Pluto LNG Train 1	2012	4.3	Shell C3MR
	QCLNG Train 1	2014	4.3	Optimized Cascade®
	QCLNG Train 2	2015	4.3	Optimized Cascade®
Australia	APLNG Train 1	2016	4.5	Optimized Cascade®
Australia	APLNG Train 2	2016	4.5	Optimized Cascade®
	GLNG Train 1	2016	3.9	Optimized Cascade®
	GLNG Train 2	2016	3.9	Optimized Cascade®
	Wheatstone LNG Train 1	2016	4.5	Optimized Cascade®
	Gorgon LNG Train 1	2016	5.2	AP-C3MR™/ Split MR
	Gorgon LNG Train 2	2016	5.2	AP-C3MR™/ Split MR
	Gorgon LNG Train 3	2017	5.2	AP-C3MR™/ Split MR
	Ichthys LNG Train 1	2017	4.5	AP-C3MR™/ Split MR
	Ichthys LNG Train 2	2018	4.5	AP-C3MR™/ Split MR
	Wheatstone LNG Train 2	2017	4.5	Optimized Cascade®
Brunei	Brunei LNG Train 1-5	1972	7.2	AP-C3MR™
	Damietta LNG Train 1	2005	5.0	AP-C3MR™/ Split MR
Egypt	ELNG Train 1	2005	3.6	Optimized Cascade [®]
	ELNG Train 2	2005	3.6	Optimized Cascade [®]
Equatorial Guinea	EGLNG Train 1	2007	3.7	Optimized Cascade [®]
	Bontang LNG Train 3	1983	2.7	AP-C3MR™
	Bontang LNG Train 4	1983	2.7	AP-C3MR™
	Bontang LNG Train 5	1989	2.9	AP-C3MR™
	Bontang LNG Train 6	1994	2.9	AP-C3MR™
Indonesia	Bontang LNG Train 7	1998	2.7	AP-C3MR™
	Bontang LNG Train 8	1999	3.0	AP-C3MR™
	Tangguh LNG Train 1	2009	3.8	AP-C3MR™/ Split MR
	Tangguh LNG Train 2	2009	3.8	AP-C3MR™/ Split MR
	Donggi-Senoro LNG	2015	2.0	AP-C3MR™
Libya	Marsa El Brega	1970	3.2	AP-C3MR™
Malaysia	MLNG Tiga Train 1-2	2003	6.8	AP-C3MR™
	MLNG Satu Train 1-3	1983	8.1	AP-C3MR™
	MLNG Dua Train 1-3	1995	7.8	AP-C3MR™
	MLNG Train 9	2016	3.6	AP-C3MR™/ Split MR
	NLNG Train 1	1999	3.3	AP-C3MR™
	NLNG Train 2	2000	3.3	AP-C3MR™
Nigeria	NLNG Train 3	2002	3.0	AP-C3MR™
	NLNG Train 4	2006	4.1	AP-C3MR™
	NLNG Train 5	2006	4.1	AP-C3MR™
	NLNG Train 6	2008	4.1	AP-C3MR™

Country	Plant	Start-up Year	LNG Capacity (mtpa)	Liquefaction Technology
Norway	Snøhvit LNG Train 1	2007	4.2	Linde MFC [®]
	Oman LNG Train 1	2000	3.6	AP-C3MR™
Oman	Oman LNG Train 2	2000	3.6	AP-C3MR™
	Qalhat LNG	2006	3.7	AP-C3MR™
Danua Naw Ouinas	PNG LNG Train 1	2014	3.5	AP-C3MR™
Papua New Guinea	PNG LNG Train 2	2014	3.5	AP-C3MR™
Peru	Peru LNG	2010	4.45	AP-C3MR™/ Split MR
	Qatargas I Train 1	1997	3.2	AP-C3MR™
	Qatargas I Train 2	1997	3.2	AP-C3MR™
	Qatargas I Train 3	1998	3.1	AP-C3MR™
	Qatargas II Train 1	2009	7.8	AP-X®
	Qatargas II Train 2	2009	7.8	AP-X [®]
	Qatargas III	2010	7.8	AP-X [®]
Ostar	Qatargas IV	2011	7.8	AP-X [®]
Qatar	RasGas I Train 1	1999	3.3	AP-C3MR™
	RasGas I Train 2	2000	3.3	AP-C3MR™
	RasGas II Train 1	2004	4.7	AP-C3MR™/ Split MR
	RasGas II Train 2	2005	4.7	AP-C3MR™/ Split MR
	RasGas II Train 3	2007	4.7	AP-C3MR™/ Split MR
	RasGas III Train 1	2009	7.8	AP-X®
	RasGas III Train 2	2010	7.8	AP-X®
	Sakhalin 2 Train 1	2009	4.8	Shell DMR
	Sakhalin 2 Train 2	2009	4.8	Shell DMR
Russia	Yamal LNG Train 1	2017	5.5	AP-C3MR™
	Yamal LNG Train 2	2018	5.5	AP-C3MR™
	Yamal LNG Train 3	2019	5.5	AP-C3MR™
	ALNG Train 1	1999	3.3	Optimized Cascade [®]
	ALNG Train 2	2002	3.4	Optimized Cascade [®]
Trinidad	ALNG Train 3	2003	3.4	Optimized Cascade [®]
	ALNG Train 4	2006	5.2	Optimized Cascade [®]
	ADGAS LNG Train 1-2	1977	2.6	AP-C3MR™
United Arab Emirates	ADGAS LNG Train 3	1994	3.2	AP-C3MR™
	Kenai LNG	1969	1.5	Optimized Cascade [®]
	Sabine Pass Train 1	2016	4.5	Optimized Cascade [®]
	Sabine Pass Train 2	2016	4.5	Optimized Cascade [®]
	Sabine Pass Train 3	2017	4.5	Optimized Cascade [®]
	Sabine Pass Train 4	2017	4.5	Optimized Cascade [®]
	Sabine Pass Train 5	2019	4.5	Optimized Cascade [®]
	Cove Point LNG	2017	5.3	AP-C3MR™/ Split MR
United States	Cameron LNG Train 1	2018	4.0	AP-C3MR™
	Cameron LNG Train 2	2018	4.0	AP-C3MR™
	Cameron LNG Train 3	2018	4.0	AP-C3MR™
	Freeport LNG T1	2018	4.4	AP-C3MR™/ Split MR
	Freeport LNG T2	2019	4.4	AP-C3MR™/ Split MR
	Freeport LNG T3	2019	4.4	AP-C3MR™/ Split MR
	Corpus Christi LNG Train 1	2019	4.5	Optimized Cascade [®]
	Corpus Christi LNG Train 2	2019	4.5	Optimized Cascade [®]
	Yemen LNG T1	2009	3.6	AP-C3MR™/ Split MR
Yemen	Yemen LNG T2	2010	3.6	AP-C3MR™/ Split MR

COSTAIN

APPENDIX 3. DESIGN STRATEGIES FOR EMISSIONS REDUCTION

A3.1. OPTIMISATION OF THERMODYNAMIC EFFICIENCY

Improvement of liquefaction efficiency by design optimisation reduces overall fuel gas consumption per tonne of LNG to achieve the commercial objectives above. In recent years, the increasing concerns regarding GHG emissions have further motivated process efficiency improvements.

Thermodynamic efficiency can by optimised using various techniques during design. Design efforts are focused to reduce the liquefaction specific power by optimising the heat transfer between the cold streams and the feed gas being liquefied by:

- Precise control of refrigerant composition and refrigerant evaporating conditions (pressures and temperature) matching the cooling curves of specific feed gas composition and process conditions.
- Use of heat exchanger technology allowing very small temperature approaches, i.e. Coil-Wound Heat Exchangers, brazed aluminium Plate-Fin Heat Exchangers and Printed Circuit Heat Exchangers.
- Optimum machinery selection, including compressor types (the use of an axial compressor in the MR compressor string enhances the process efficiency) compressor sizes and train configurations allowing operation with high mechanical efficiency. Also use of hydraulic turbines as a replacement for Joule-Thomson valves in refrigerant expansion allows increase of production for a given installed power.
- Overall heat integration optimisation.

It is noted that the achievable benefits of design optimisation strategies for a given LNG plant are dependent on a wide range of variables and considerations specific to the plant such as:

- Feed gas composition. Lean feed gas (including gas with high nitrogen content) increases the liquefaction specific power requirement. If feed gas contains ethane+ components, more feed gas can be liquefied for a given liquefaction capacity (i.e. installed refrigerant compression power).
- Feed gas conditions. Feed gas at high pressure reduces the specific power.
- Gas cooling temperature. Where air coolers are used to cool compressed refrigerant streams, the liquefaction specific power increases for increased ambient temperature. Availability of cold seawater for gas cooling allows significant reductions in the specific power, but high costs, environmental concerns due to water discharge temperatures and reliability issues associated to handling seawater need to be taken into account.
- Ambient conditions. The main impact of ambient air temperature is in the degradation of the performance of Gas Turbines. Cold environments allow higher Gas Turbine efficiency plus reductions of heat in-leak into the cryogenic equipment, hence reducing of the specific power.
- Fuel gas calorific value. Less fuel gas will be required if this has a relatively high calorific value



A3.2. HEAT RECOVERY AND INTEGRATION

Installing a heat recovery system on the gas turbine exhaust makes it possible to recover the otherwise wasted heat energy.

Environmental and economic benefits can be achieved through the conversion of simple cycle gas turbines used in power generation duties into combined cycle plants. A combined cycle can reduce CO_2 emissions by improving the energy efficiency of a gas turbine by recovering the waste heat from the flue gas in the form of power. This power can then be utilised for electrical equipment. It is reported that CO_2 emissions in power generation can be reduced by about 30% through the use of a combined cycle compared to simple cycle (Ref 13).

For mechanical drive, potential for CO_2 emission reductions of 40 to 50% are reported to be achievable when combined cycle power generation is used with electric motor driven refrigeration compressors, when compared to gas turbine driven compressors (Ref 20).

Table A3.1 shows how the improvement in power output and thermal efficiency for representative Frame 6, Frame 7 and Frame 9 gas turbines in a combined cycle configuration.

Power Generation Unit	Cycle Type	Net Plant Power Output (MW)	Cycle Thermal Efficiency (%)
Eramo 6B 03	Simple	44	34
Flame 0B.03	Combined	67	52
Frame 7E.03	Simple	91	34
	Combined	139	51
Frame 0E 02	Simple	132	35
	Combined	199	52

Table A3.1 Performance Comparison of Power Generation Gas Turbines by Cycle Type

Heat Recovery Steam Generator (HRSG) systems can be adopted to use the heat energy from the turbine exhaust to produce steam. Steam can be used to drive machinery (e.g. BOG compressor and to provide heating duties to the process substituting steam generation boilers. This reduces the CO_2 emissions associated to power generation for electric motors. Reported CO_2 emission reductions are in the order of up to 30% for an LNG plant operating with a HRSG system (Ref 13).

Waste heat recovery units (WHRU) can be installed in the exhaust of gas turbines to provide heating via a hot oil system or steam generation to other areas of the LNG plant such as for regeneration of the AGRU solvent or regeneration of molecular sieve adsorbent used in dehydration, thus reducing the fuel requirements of the plant as a whole and allowing the plant to move away from using other sources of heating such as fired heaters or steam boilers. It is reported that the CAPEX of a waste heat recovery system is generally higher than fired heating systems (Ref 20). However, waste heat recovery allows reductions in fuel requirement (otherwise required to generate heat) and associated emissions.

Steam demand and availability considerations need to be taken into account in the design of the overall plant, as production of steam is dependent upon the operation of the gas turbines. As a result, availability of the gas turbines should be taken into consideration during design to avoid shortfalls in energy provision from the WHRU. Steam demand for heating duty

depends on the particular plant, e.g. steam for AGRU solvent regeneration duties will depend on the CO₂ content in feed gas. Where the steam demand can only be met by WHRU in all gas turbines, spare steam generation capacity may need to be added in the form of steam generation boilers to reduce the dependency of steam generation on gas turbine operation.

Table A3.2 shows the potential heat duty from WHRU installed on gas turbine drive exhausts against estimated AGRU solvent regeneration duties for typical LNG train configurations. Potential for heat recovery from aeroderivative gas turbine exhausts is lower due to the higher efficiency, compared to industrial gas turbines. It is noted that excess heat is generally available. In the Pluto LNG plant, waste heat recovery is only implemented in two of the four Frame 6 gas turbines used for power generation and in one of the two Frame 7 gas turbines driving the refrigerant compressors. The WHRU capacity installed is sufficient to meet the process heating requirements for the AGRU and glycol regeneration systems (Ref. 50).

Driver	LNG Train Capacity (mtpa)	Thermal Regeneration Duty (GJ/h) (Note 2)		WHRU Heating Duty (GJ/h) (Note 3)	
	(Note 1)	2 mol% CO ₂	6 mol% CO ₂	T _{out} = 300°C	T _{out} = 200°C
6x PGT25+	5.2	220	560	420	630
6x LM6000	7.3	300	770	490	800
6x Frame 5C	4.7	200	520	650	950
6x Frame 5D	5.4	230	590	720	1060
2x Frame 7	4.8	210	530	580	820
3x Frame 9	7.7	310	820	1230	1740

 Table A3.2 Indicative AGRU and WHRU Duty for Representative Liquefaction Train

 Configurations

Notes:

- Indicative LNG capacity estimated based on representative specific power 0.3 kWh/kg, 93% plant availability and gas turbine performance (output power and efficiency) at ISO conditions (see Table 2). The LNG capacity reported for the Frame 9 configuration is as per AP-X trains in Qatar, where about 70% of the power output (276 MW) is used to drive refrigerant compressors while the rest is exported as power via a generator in the same shaft.
- 2. Based on the assumption of a representative 2 and 6 mol% CO₂ content in feed gas, amine regeneration reboiler duty is calculated as a function of solvent circulation rate based on 30 wt% MEA as per GPSA data book Section 21 Equation 21-6 and Figure 21-4. The thermal regeneration duty figures also include the dehydration gas regeneration heating duty, which is calculated to heat 10% of the feed gas (as regeneration gas) from 5°C to 280°C.
- 3. Waste heat recovery based on cooling of flue gas to 300°C and 200°C from all mechanical drive gas turbines in the liquefaction train.

A3.3. USE OF MORE EFFICIENT REFRIGERANT COMPRESSOR DRIVES

The selection of turbine type plays a key role in the reduction of emissions. Developments in gas turbine technology have resulted in both increased power output and increased efficiency. Aeroderivative type gas turbines have been used as an alternative to industrial gas turbines to increase overall fuel efficiency and reduce emissions as shown in Table 1.

Utilisation of various techniques on gas turbines is reported to provide further improvements to the performance of combined cycle power plants (Ref 25). These techniques broadly fall under two categories: cooling of gas turbine inlet air and intercooling of air during compression. Both techniques use the same principles in order to achieve performance enhancements. Cooling the air to a lower temperature causes the density of the air to increase and the mass flowrate of air increases, which leads to enhancement in gas turbine power output.

Gas turbine fuel heating can be used to reduce the amount of fuel energy used to bring the fuel up to combustion temperature, in doing so the cycle efficiency is increased, and emissions reduction achieved. Applying this principle is reported to give a marginal reduction in power output (Ref 25). This reduction in power output may be greater in combined cycle type applications in comparison to simple cycle, since the heat used for raising the temperature of the fuel gas in a combined cycle type would otherwise be utilised elsewhere.

In recent years there has been an increased interest towards the use of all electric motor driven refrigerant compressors. Electrically driven refrigeration compressors offer benefits compared to traditional industrial gas turbines of equivalent rating, including:

- Increased overall fuel efficiency
- Increased life in service including increased availability
- Reduced outage frequency and duration for periodic service
- Reduced maintenance costs
- Reduced mechanical sensitivity (small clearances etc.) and complexity of components
- Improved start-up capability
- Wide range of power output ratings
- Reduced (minor) influence of ambient conditions on performance
- Availability of a larger number of options (suppliers) for procurement of units, components and technical services

From an emissions perspective, electrically driven refrigeration compressors provide an opportunity for improved overall energy efficiency by using power generated in a centralised combined cycle plant. This offers a higher fuel efficiency overall than a gas turbine in a mechanical drive service, particularly with a power generation plant designed to operate at high load to meet the overall liquefaction plant power demand. Furthermore, the ability to utilise waste heat from all power generation gas turbine exhausts is retained, reducing fuel gas requirement to meet process heating demand.

It is claimed that electrical efficiencies (the ratio between shaft output power and electrical input power) in excess of 95% can be achieved for electric motor variable speed drives (Ref

24). Even considering the energy efficiency in the power generation as well as other key items such as transformers, motors and inverters, the overall efficiency per unit compressors driven by electric motors is reported to be higher compared to the efficiency of at compressor driven by as turbine.

Component	Efficiency (indicative)			
Component	Gas Turbine Driven	Electric Motor Driven		
Gas Turbine	30 % (Note 1)	-		
Power Generation	-	47 % (Note 1)		
Transformer	-	82 %		
Motor	-	97 %		
Inverter	-	99 %		
Overall	30 %	37 %		

Table A3.3. Comparison of Efficiencies of Gas Turbine vs Electrically Driven Compression(Ref 14)

Note 1. Representative de-rated cycle power generation efficiency.

While all-electric LNG plants appear to offer potential advantages over more traditional style plants using gas turbine drives, it is essential to ensure that LNG production is not adversely affected. The availability and reliability of the electric power source is an important factor to consider when investigating the use of an all-electric driven LNG plant. The adverse side effects of electric variable-speed drives must be well understood. Power-system harmonics and oscillating torques in the compressor strings may lead to varying levels of detrimental effects during operation. Events resulting in power instabilities (e.g. mechanical failures, electrical faults, transient events) may impact the availability of the LNG plant and lead to commercial implications.

A3.4. OPTIMISATION OF UNIT OPERATIONS

A number of design strategies can be considered to obtain efficiency gains. These include optimisation of flowsheets and process conditions and the use of more efficient equipment.

For standalone NGL recovery processes, feed gas is cooled by either using a mechanical refrigeration package or by letting down the feed gas pressure to cool the feed gas by Joule-Thomson effect, with high pressure feed gas cooling achieved by heat exchange with returning cold streams.

Depending on the optimum or required feed gas pressure for liquefaction, a booster compressor downstream of the NGL recovery unit may be installed. By optimising of the operating pressure of the NGL recovery column, a reduction in overall compression power (and associated fuel gas and emissions) in the order of 5% may be possible (Ref 13). High operating pressure in the NGL recovery column will lead to a reduction in the feed gas booster compressor size and power requirements. Maximisation of the pressure of the gas sent to liquefaction will increase the LNG production for a given installed liquefaction cycle power (Ref 13).

It is noted that the operating pressure of the NGL recovery column is dictated (limited) by the critical pressure of the hydrocarbon streams in the column on the basis that relative volatility decreases significantly as the column pressure approaches the critical pressure leading to reduced separation capacity.

There is potential for liquid expanders to be used instead of traditional pressure let down valves within the liquefaction process. Expanders produce lower outlet temperatures which allow improvements in the overall liquefaction efficiency. Expanders also generate power which can contribute to meeting power generation requirements which in turn may lead to reductions in both the overall fuel gas use and the associated CO_2 emissions.

Expanders can be installed in the End Flash Gas system to let down the LNG product to storage pressure. Expanders can be installed within the refrigerant circuits to reduce high pressure streams to a lower pressure. This results in LNG production increase reported to be in the order of 6% (Ref 20). Experience exists in the use of two- phase expanders on LNG service which removes the design and operational constraints and the power requirement to subcool fluids when using conventional liquid expanders (Ref 21).

Production efficiency improvements can also be achieved by the use of hybrid coolers or wet surface air coolers. In these, water is placed into contact with the exterior surface of the heat exchanger tube bundles in combination with a flow of air to provide cooling through evaporative heat transfer. It is reported that hybrid coolers have the potential to improve production efficiency when compared to standard air coolers as they are able to reduce the outlet temperature of the cooled stream further, and provide benefits when compared to a direct cooling water system as they use less water which can be of lower quality (Ref 22).

Hybrid coolers can be used as refrigerant compressor aftercoolers, with reduced refrigerant outlet temperature allowing reductions in the overall liquefaction cycle power requirements, thus reducing emissions per tonne of LNG produced. Benefits gained from the use of hybrid coolers vary depending on site location as they depend on ambient conditions, such as temperature, as they may encounter problems associated with freezing in cold climates, and relative humidity, since they are most beneficial in climates with low relative humidity (larger differential between dry and wet bulb temperature).

There is potential for a reduction in heating required in the condensate stabilisation unit potentially in the order of up to 20% through the use of heat integration and operating pressure optimisation (Ref 13). This can be achieved by introducing side reboilers to the stabiliser column and integrating them with the hot bottom product stream from the stabiliser which leads to a reduced reboiler duty and reduced power requirements. Adding a reflux to the stabiliser column reduces power demands of the overhead off gas compressors.

A4.1. SLEIPNER

The Sleipner platform is a gas production facility located in the North Sea (off the coast of Norway) which delivers 300,000 barrels/day of oil equivalent and 36 million standard cubic metres of natural gas. The feed gas from Sleipner West field ranges from 4 to 9 mol% CO₂ which has to be reduced to meet European gas specifications.

The driving force behind the CCS scheme was the introduction of CO_2 tax by the Norwegian government (\$35 per tonne of CO_2 in 1996, rising to \$65 in 2016 (Ref 52). Sleipner has operated with the CCS scheme since 1996, using an amine process to remove up to 1.2 mtpa of CO_2 from the feed gas which is then stored underground in the Utsira sandstone formation at a depth of about 1000 m below the sea level. Over the field's expected life of 25 years, injection of around 25 million tonnes of CO_2 was estimated, but the actual figure is expected to be below this due to a lower CO_2 content and a decreasing production profile.

The CO_2 is captured from the natural gas stream with a conventional amine process using methyl diethanolamine (MDEA) as the solvent. After capture, the CO_2 is compressed to a supercritical state in four compression trains to a pressure of 80 bara for injection. Stainless steel well casing and other hardware removed the need for dehydration of the nearly pure CO_2 stream. The cost of compression and injection alone at Sleipner was \$80 million in 1996 (Ref 12).

An extensive program has been put in place to monitor and model the distribution of injected CO_2 in the Utsira Formation.



Figure A4.1. CCS Scheme in Sleipner Platform

A4.2. IN SALAH

The In Salah plant in Algeria, which is no longer operational, annually produced 9 billion cubic meters of gas from natural gas reserves.

The CO₂ content in the natural gas feed to the plant ranged from 1 to 9 mol%. In order to meet exportation specifications the quantity of CO₂ was reduced to a maximum of 0.3 mol% before being sent to market. The plant used an amine-based process to remove the CO₂. The captured CO₂ was then compressed, dehydrated and injected at 140-180 bara in the Krechba field at 1800 m depth. The scheme captured approximately 1.2 mtpa of CO₂. Injection was achieved through the use of long reach horizontal wells and its success gives positive insight into the potential for CO₂storage in similar wells which are widely seen in the US, Northwest Europe and China.

Overall the project has allowed significant experience to be gained on the operational aspects of CO_2 storage. The development facilitates testing of alternative monitoring technologies to improve the understanding of CO_2 behaviour in the subsurface, particularly the development of the CO_2 plume which is strongly controlled by initially unpredicted geological factors. Through monitoring the performance of the injection wells it has become apparent that understanding the rock mechanical behaviour of the CO_2 storage reservoirs is a key factor in early decision making process, ultimately impacting on the injection capacity.



Figure A4.2. CCS Scheme in In Salah Plant



A4.3. BOUNDARY DAM

The Boundary Dam CCS Project involves SO_2 and CO_2 removal from the flue gas from a new 161 MW unit at the Boundary Dam coal fired power station located in Saskatchewan, Canada. The CCS scheme was installed as part of a refurbishment to the power generation plant to capture 1 mtpa of CO_2 . The scheme was implemented in response of environmental regulations and to enable the plant to continue operation using lignite fossil fuels of which there are said to be reserves of up to 250-500 years in Saskatchewan. The CO_2 captured is used to increase oil production in a declining oil industry in the area, i.e. used for Enhanced Oil Recovery (EOR) purposes. The CO_2 is geologically stored in an oil reservoir approximately 1400 m deep and also in a deep saline aquifer approximately 3200 m deep.



Figure A4.3. Boundary Dam Plant CCS Scheme

The cost of the project was estimated to be \$1.24 CAD billion which rose to \$1.38 CAD billion, about half of this cost was for the capture scheme and the remainder was for refurbishments and upgrades of the power plant. The project was supported financially by the Canadian government, who provided \$240 CAD million in funding, (Ref 53)

Part of the project saw the old steam turbine replaced with a new 161 MW dual-mode turbine with enhanced steam and thermal integration. This was fully integrated to provide power and heating for the capture processes resulting in the following balance:

Installed capacity	161 MW
Power generation parasitic load	-11 MW
CO ₂ capture solvent regeneration	-14 MW
CO ₂ capture power demand	-9 MW
CO ₂ compression power demand	-15 MW
Power to grid	112 MW

The new turbine has the capability to handle the variable power demand from the capture system, ensuring that the power plant can always operate at full load. Additional modifications included a boiler upgrade, a new feed water system for the additional use of steam, the cooling system was converted to closed loop and flue gas ducting to the capture plant was installed.

The flue gases are routed to the existing vent stacks when the CO₂ capture plant is not in operation, which is the case when the power plant operates at less than 50% load (i.e. 50% is the minimum turndown capacity of the capture plant). Flue gases are cooled in a waste heat recovery unit, pre-heating the condenser water used for steam generation in the power plant resulting in an improvement to the overall power generation efficiency by 3.5 MW. A direct contact cooler (pre-scrubber) further reduces the temperature of the flue gases and removes particulates and other contaminants.

The CO_2 capture process utilises Shell Cansolv amine-based absorption technology. The process uses a ceramic lined concrete CO_2 absorber tower standing at 55 m tall, while the stripper column has a diameter just over 7 m and stands at over 40 m tall. The captured CO_2 is then compressed by an eight stage integrally geared reciprocating compressor to around 170 bara and dehydrated. The flue gas that has been stripped of CO_2 , SO_2 , particulates and other contaminants is water-washed and released to the atmosphere through a vent stack at the top of the CO_2 absorber.



Figure A4.4. Boundary Dam Plant

Lessons learned in the commissioning, start-up and in the initial operation period are expected to facilitate future retrofits, reducing costs and duration. In this way, it is perceived that design margins applied on the capture plant could be reduced to allow lower CAPEX and engineering costs. Initial operation has indicated that the dual-mode power generation capability (being a major contributor to cost) is not needed. The possibility of standardisation of modules to be installed in other similar plants to reduce CAPEX is under evaluation. It has been suggested that implementation of the lessons learned would allow cost reductions in the order of 20% to 30% for future projects.



A4.4. PETRA NOVA

The Petra Nova CO_2 capture project at Thompsons, Texas (US) which is due to start operations in 2017 will be the world's largest post combustion CO_2 capture project capturing 1.4 mtpa of CO_2 from a 240 MW coal fired power plant. The cost of the project was \$1 billion, with \$167 million provided by the US Department of Energy as part of their Clean Coal Power Initiative (Ref 54).

The captured CO_2 is compressed for use in EOR to boost daily oil production of a nearby field from 500 barrels to 15,000 barrels, potentially recovering 60 million additional barrels of oil in total. The size of the CO_2 capture plant was quadrupled from original plans, as it was identified that the original 60 MW project would produce an insufficient supply of CO_2 to pressurise the oil field and would not produce enough oil to make the project financially attractive.

In the absence of a substantial penalty for CO_2 emissions, the economics of the project are based on the oil field production, with the power plant taking ownership of a 50% equity share in the oil field in which the EOR is employed. This strategy (unlike that used in the Boundary Dam project, whereby the captured CO_2 is sold) will generate revenue from the increase in oil production. This contributes to the business case of the project, making this feasible from the commercial point of view as the sale price of CO_2 (\$15-35 per tonne of CO_2) represents only a small fraction of the potential value of the additional oil that is recovered (\$150 per tonne of CO_2 based \$50 per barrel of oil).





The flue gas (slipstream off the 650 MW coal fired boiler) containing 11.5 mol% of CO_2 will be processed using Mitsubishi Heavy Industries' (MHI) KM-CDR amine-based solvent technology, with the plant having a capacity to capture 4776 tonnes/day of CO_2 . This technology had been previously proven at a 500 tonnes/day of CO_2 demonstration plant at the Southern Company Berry Coal Power Plant. The flue gas is cooled in a flue gas "quencher" tower and then treated in a 97.5 m high absorber. The absorber is fabricated as steel modules with a square cross sectional area, while the stripper is a column.



In contrast to the Boundary Dam project, integration with the power plant is minimised. A new cogeneration plant consisting of a combustion turbine generator and heat recovery steam generator (HRSG) will provide power and steam required for the CO_2 capture and compression system's operation. Any excess electricity from the generator is exported. Balance of plant facilities includes new systems for water cooling (mechanical-draft cooling tower), water treatment, demineralised water, instrument air and electrical transformer. Refurbishment of existing plant is limited to minimum tie-in work and there is no additional parasitic load on the existing plant. The new cogeneration plant is more efficient than the coal plant, provides the flexibility to match the energy requirements of the CO_2 capture process and minimises disruptions to existing operations.

A4.5. OTHER RELEVANT PROJECTS

The Uthmaniyah CO_2 -EOR (Enhanced Oil Recovery) Demonstration Project in Saudi Arabia captures 0.8 mtpa of CO_2 from the Hawiyah NGL Recovery Plant. The CO_2 produced from the NGL recovery plant is compressed and dehydrated as part of the CCS scheme then injected into the Ghawar oil field as part of an EOR scheme. The project started in 2015 and is expected to last for 3 to 5 years. The objectives of the CCS demonstration scheme are to determine the potential for incremental oil recovery and gain experience to identify potential operational concerns and address primary risks and uncertainties in the , including migration of CO_2 within the reservoir. A comprehensive monitoring and surveillance plan has been implemented including the use of new technologies for CO_2 plume tracking and CO_2 saturation modelling.

Part of the 1200 MW expansion project at Sinopec Shengli coal fired power plant, the Sinopec Qilu Petrochemical CCS project will treat a 200 MW slipstream of flue gas. This will be achieved by a process using monoethanolamine (MEA) as an absorbent for post combustion capture of CO_2 . The project, currently anticipated to commence operation by 2020 will produce a 1 mtpa of CO_2 to increase oil recovery by 10-15% through EOR.

The Rotterdam Opslag en Afvang Demonstraieproject (ROAD) currently in define phase and due to start operations in 2019/20 is planned to capture CO_2 from flue gas from an existing power station in Maasvlakte in Rotterdam, Netherlands using an amine-based absorption process. The scheme will capture 1.1 mtpa of CO_2 from a 240 MW equivalent flue gas slip stream from the power plant. The CO_2 will then be compressed, dehydrated and metered onsite, before being piped for storage in an offshore reservoir. The project has received €180 million worth of funding from the European Commission (EC) as part of the European Energy Program for Recovery and €150 million from the Dutch government (Ref 55). The project has experienced slow down due to funding gaps caused by low carbon prices but has seen reductions in the expected operating expenditure (OPEX) due to an expected fall in electricity prices (approximately 50% of OPEX cost is for electricity) with electricity requirements being estimated at about 0.33 MWh per tonne of CO_2 captured.

A5.1. PRE-COMBUSTION

Power generation with pre-combustion CO_2 capture involves reacting fuel with oxygen or air to produce synthesis gas (a mixture of hydrogen and carbon monoxide) either from reforming natural gas or gasification of coal or biomass, making pre-combustion capture suitable for use in Integrated Gasification Combined Cycle (IGCC). It is noted that at present, none of the existing coal-fired IGCC plants includes shift conversion with CO_2 capture.

The carbon monoxide is reacted with steam in a catalytic reactor (shift converter) to give CO_2 and more hydrogen. This is a reversible slightly exothermic process, so conversion to CO_2 and hydrogen is promoted at low temperatures. The hydrogen yield is maximised by shift reactors operating at successively lower temperatures. CO_2 is then separated, usually by a physical or chemical absorption process, resulting in a hydrogen-rich fuel. The hydrogen stream with a purity of typically above 93 mol% is then conditioned and burnt in a CCGT.





The advantage of this technology is that the synthesis gas from the shift converter reactor is rich in CO_2 and at high pressure, making separation of the CO_2 less energy intensive and the separation equipment more compact (compared to a post-combustion scheme). The efficiency loss due to CO_2 capture is in the order of 3-4% as a result of steam demands in the shift reactor.

Pre-combustion capture technologies which have been considered include both physical and chemical absorption.

- Chemical absorption techniques, similar to those considered for post-combustion capture, could be appropriate but these are mainly used for low CO₂ partial pressures such as those found in flue gases.
- Physical absorption is appropriate for pre-combustion capture and becomes more energy efficient than chemical absorption as the partial pressure of CO₂ in the shifter syngas is much higher than in flue gases in post-combustion capture. Physical absorption processes selectively absorb the CO₂ in a solvent at high pressure, which is then let down in pressure to desorb the CO₂. The two major physical solvents available commercially are Selexol (mixture of polyethylene glycol dimethylether) and Rectisol (methanol). The main energy penalty comes from pumping the solvent, steam stripping and refrigeration.

Removal of CO_2 from natural gas to meet downstream gas specifications (e.g. pipeline gas specification or as dictated by downstream cryogenic processing) has in some instances being referred as pre-combustion as the CO_2 contained in the gas ultimately used as fuel is removed prior to combustion. However, this may be considered as a gas pre-treatment process rather than a pre-combustion capture route as described in this section.

Developments in pre-combustion capture include combined reaction / separation techniques to overcome equilibrium limitations in the reactor. Such methods could remove the need for shift reactors and a separate CO_2 separation system. One method which has been investigated is sorption enhanced reforming where a catalyst and solid adsorbent packed bed is used to drive the reaction to completion by continual removal of CO_2 . High temperature tolerant and hydrogen selective inorganic membranes are also being considered for the reformers with the retentate being mainly CO_2 and steam, and the permeate low pressure hydrogen.

Pre-combustion capture is based on well-known and proven technologies used in chemicals and refinery industries to produce hydrogen rich gas. Gas turbines with hydrogen capability have been developed and tested and some models are commercially available. A major barrier to implementation of pre-combustion is the high capital cost. Each new plant requires considerable development compared to a plant based on established technologies, such as the more mature pulverised coal technology for coal power generation plants or gas fired power plants using natural gas fuel. Further development on a commercial scale, competitive costs and proven reliability are required before pre-combustion can become a more widely accepted technology.

A primary consideration for selection of this method of CO₂ capture in LNG plants is that the hydrogen used as fuel for power generation has to be produced by an external plant with specific technology. The availability of fuel from this source becomes critical to LNG production, impacting the availability of the LNG plant.

Project economics for pre-combustion scheme may become feasible if economies of scale suggest a larger production facility with export of hydrogen rich gas for other applications, e.g. ammonia production, Fischer-Tropsch processes or power generation (Ref 19).



A5.2. OXYFUEL COMBUSTION

Oxyfuel combustion of fuels in pure oxygen rather than air results in flue gas consisting of mainly CO_2 and water. Flue gases have a very high CO_2 concentration, practically higher than 75 mol% on a dry basis (Ref 32), compared with 15 mol% from conventional combustion of coal and much lower (in the order of 3 mol%) for gas combustion. The balance of the flue gas is made up mainly of nitrogen, with some oxygen and argon together with SO_x and NO_x impurities.

Low temperature separation technology can be applied for purification of the CO_2 flue gas to remove oxygen, nitrogen and argon. Temperatures must be maintained above approximately -55°C, constrained by the CO_2 triple point and consideration of CO_2 solidification which may limit CO_2 recovery. Hybrid technology, including low temperature distillation and membrane separation, for improved recovery of CO_2 could potentially be applied to purification of oxyfuel flue gas. Feed oxygen purity and air ingress are the main factors influencing flue gas composition.



The main energy penalty arises from the need to produce large quantities of oxygen in an air separation unit. Cryogenic air separation is the most economic option at large scale. It is a very mature technology and only small improvements in performance could be expected, for example through improved compression efficiency.

Developments in oxyfuel power generation technology have been focussed primarily on boiler design for steam turbine power generation, with temperatures maintained at close to conventional levels by recycling of a portion of the CO₂ rich flue gas.

Oxyfuel combustion is at a relatively early stage of development but integrated pilot plants are being built and development programmes are underway. Adaption of current boiler designs to accommodate oxyfuel firing is expected to require relatively small modifications in comparison with oxyfuel gas turbine cycles, which require fundamental design modifications to accommodate the change from a nitrogen rich to an oxygen/CO₂ rich combustion mixture.

Oxyfuel combustion of natural gas in gas turbines is still under development and not yet available on a commercial scale. Technical risks and design uncertainties suggest that oxyfuel will not be implemented in the near future.

On the assumption that new gas turbines can be designed or existing gas turbines retrofitted to handle the oxyfuel combustion mixture, selection of oxyfuel technology in LNG plants would mean that LNG production would become dependent on the ability of the air separation to produce oxygen to fuel the power generation.

A5.3. POST-COMBUSTION

Post-combustion is the most mature technology for capturing CO₂ from power plants, with a large number of research and development projects, pilot plants and recent large-scale plants being built and operated. Efforts have been focused on coal fired power plants, but experience (mainly at pilot level) also exists for gas fired power plants.

Power generation with post combustion capture involves fuel being burned in air in the traditional way, with CO_2 captured from the resultant flue gas mainly using chemical absorption.

The benefit of this CO_2 capture scheme is that it can be retrofitted to existing plants with fewer modifications to existing equipment than e.g. pre-combustion capture or oxyfuel, although the additional plot area and utilities required could make this difficult unless considered in the initial design. As a result of this, new "capture ready" plants are being built with additional plot put aside for CO_2 capture schemes to be adopted in the future.



Post-combustion capture schemes can be considered as standalone processes to some extent, with unavailability of the capture plant having a minimum/negligible impact on operations of the core power generation process. Untreated flue gases would be vented to atmosphere when the capture plant is unavailable.

Capture plants based on solvent absorption have a limited turndown capacity of typically 50%, which is a consideration for power plants operating at partial load, but it is not an issue for baseload plants.

Where steam requirements of the post-combustion capture plants are integrated with the power generation cycle, significant modifications to the core power generation may be required, particularly to ensure high efficiency when the capture plant is not operating. The capture plant can be designed with independent utility systems to decouple operations e.g. a dedicated Combined Heat and Power (CHP) plant.

Flue gases tend to be close to atmospheric pressure with CO_2 concentration of approximately 3% from gas turbines and 10-15 % from coal fired boilers. This combination of low pressure and low CO_2 concentration result in an energy intensive capture scheme, and many developments are focussed on lowering this energy penalty.

Physical absorption processes, which require a pressure driving force for CO_2 removal, do not lend themselves well to this application, primarily because of the large power requirement for flue gas compression.

Chemical absorption is better suited, as mass transfer and absorption is driven by chemical reaction. Amine solvents are attracting the most attention as they have already been used for CO₂ removal from natural gas on a commercial scale and considerable operational experience exists.



Equipment size is a key consideration given the large volumetric flow of flue gas as a result of low flue gas pressure and relatively large mass flow of excess air used in combustion. Although the size of the capture plant required for full scale post-combustion capture in power plants is significantly larger than the existing amine plants in natural gas processing, increasing the equipment size is not considered to be a major problem. For example, columns with a very large diameter are required, particularly for the absorber. Alternative designs for the absorber with a rectangular cross sectional area and constructed in concrete or from steel modules have been considered. Stripper columns with large diameters in the order of 7 metres have been built.

There is a need to identify a solvent that has low vapour pressure to minimise losses, is not easily oxidised, is highly reactive and has low heat of reaction with CO₂. Monoethanolamine (MEA) typically in a 30 wt% aqueous solution has been proposed for this particular application because of its fast reaction rate, giving a larger driving force for absorption and lowering the height required for the absorbers. Solvent formulations based on concentrated MEA (40-50 wt%) with additives including foaming, corrosion and oxidation inhibitors and proprietary solvents have been developed, tested and are commercially available. These proprietary solvents and formulations are reported to have lower regeneration energy requirements and reduced degradation. Other solvents such as chilled ammonia or amino acids also have the potential to reduce energy consumption substantially.

Process streams are being developed to make the absorption process more efficient. In addition to implementing existing technology such as split flow and absorber side stream coolers, novel schemes such as the use of vapour recompression to reduce solvent regeneration reboiler duty or the use of multi pressure stripper configurations can reduce the energy requirements.

APPENDIX 6. CO₂ CAPTURE TECHNOLOGIES

A6.1. CHEMICAL ABSORPTION

Chemical absorption is a standard technology for CO₂ removal in natural gas processing which uses liquid absorption solvents, particularly based on alkanolamines. Amino acids, hot potassium carbonate, chilled ammonia and caustic wash processes are also used. These processes remove acid gases CO₂ and/or H₂S and sulphur compounds from the gas stream by chemical reaction with the solvent. The reaction is reversed by high temperature and/or low pressure in the stripper. Chemical solvent processes are particularly applicable where acid gas partial pressures are low and/or low levels of acid gas are desired in the residue gas.

A6.1.1. Amines

Processes using amine-based chemical solvents are commercially proven systems widely used for bulk removal of CO_2 in natural gas processing including removal of CO_2 from flue gases. The versatility of the amine solvents allows this technology to be applied for gases containing a wide range of CO_2 contents from below 1 mol% to 50 mol% (Ref 34).

Amine systems (see Figure A6.1) involve the use of two main columns: an absorber (contactor) and a solvent regeneration column (stripper). A water wash section is typically included on top of the absorber column to recover some of the vaporised and/or entrained amine from the gas leaving the absorber. The treated gas leaving the absorber is saturated with water. The rich solvent is pumped to the stripper column where absorbed CO_2 is stripped off the solvent, allowing recirculation of the lean solvent. Overhead vapours are partially condensed, with highly pure CO_2 saturated with water leaving as vapour in the overhead condenser separator. Lean solvent produced at the column bottoms is passed to a lean-rich solvent exchanger for heat recovery and further cooled to the operating temperature of the absorber.



Figure A6.1. Typical Amine-Based Absorption Process (Ref 33)

The reaction of CO_2 with amines produces a stable carbamate which further hydrolyses to bicarbonate. These are exothermic reactions that increase the solvent temperature which reduces the absorption performance, so absorber intercooling can be installed.

In natural gas processing applications, the rich amine solution leaving the absorber is flashed to remove light hydrocarbons absorbed, with heavier hydrocarbons removed as a separate liquid phase. The CO_2 rich solvent from the absorber is heated and fed to the stripper column where the CO_2 is removed and the solvent regenerated. The lean solvent is cooled, filtered and pumped back to the top of the absorber.

In gas processing, specific amines can be used to selectively remove H_2S in the presence of CO_2 .

- Monoethanolamine (MEA) removes both H₂S and CO₂ and is used where there are low contactor pressures and/or stringent treated gas specifications. CO₂ concentrations as low as 100 ppmv can be obtained at low to moderate pressures.
- Diethanolamine (DEA) can be also used, with DEA loadings being much higher than those achieved with MEA at high pressures. However, MEA is more efficient at low pressures.
- Diisopropanolamine (DIPA) and Methyldiethanolamine (MDEA) are used to selectively remove H₂S at moderate to high pressure.
- Triethanolamine (TEA) exhibits selectivity for H₂S over CO₂ at low pressures.

Solvent formulations may contain activators such as piperazine to achieve very low CO_2 concentrations in the treated gas. For the treatment of flue gases removal of H_2S is not required, and absorption is carried out at low pressure, so 30 wt% MEA in aqueous solution is typically selected as the most appropriate solvent. Proprietary solvent formulations include more concentrated MEA (40-50 wt%), blends of secondary and tertiary amines (DEA, DIPA, MDEA and TEA) and may include additives including foaming, corrosion and oxidation inhibitors such as piperazine derivatives (Ref 30). This technology is commerically available and proven on full scale coal fired power plants (see Appendix 4). There are a number of references on pilot schemes for CO_2 capture from CHP plants e.g. at the Technology Centre Mongstad (Ref 41) with flue gas composition similar to that expected from gas turbines (3.5 mol% CO_2 and 14 mol% oxygen).

Relatively high energy requirements are associated with solvent regeneration. Specific Reboiler Duty (SDR) is a widely used reference value representing the stripper reboiler duty required to regenerate solvent per tonne of captured CO_2 . This duty is indicative of the amount of steam and overall operating costs. The SDR of a standard post-combustion capture absorption process using 30 wt% aqueous MEA is typically reported in the order of 3.7 GJ per tonne of captured CO_2 . Reduction in the SDR reported as low as 2.3 GJ per tonne of captured CO_2 can be achieved by using more concentrated MEA or using proprietary solvent formulations.

Amines degrade through reaction with oxygen, NO_x and SO_x present in flue gases, to form heat stable salts, introducing the need for amine reclaiming and increasing energy requirements. Depending on the flue gas composition, a facility upstream of

the CO₂ removal process may also be required to remove these components e.g. flue gas desulphurisation and selective catalytic reduction, required for coal fired plants but not expected for gas fired plants.

The presence of oxygen needs consideration, particularly for flue gases from gas turbines using a large volume of excess air. Oxygen can cause degradation of some solvents and corrosion of equipment. Novel solvents including formulations with oxidation inhibitors have been developed and tested (Ref 41) and are commercially available to treat flue gases from gas fired plants with relatively high oxygen content. As an alternative to remove the need for inhibiting agents to prevent corrosion, variations of the amine processes have been developed to remove dissolved oxygen by flashing the rich solvent prior to regeneration (Ref 30, 42).

Degradation products including heat stable salts, suspended solids, acids and iron compounds are periodically removed from the circulating solvent by thermal reclaiming of a slip steam of the lean solvent (1 to 3% of the total amine circulation rate). This is a distillation process where amine is recovered while degradation products are removed as sludge and treated as waste.

Periodic solvent make-up is required due to degradation and carryover loses in treated flue gas leaving the absorber. Proprietary solvents are reported to be more stable and tolerant of oxygen levels in typical flue gases. The result is make-up rates in the order of 0.2 kg of solvent per tonne of captured CO₂, which is potentially up to 10 times lower than make-up requirements using MEA. The relatively higher costs of proprietary solvents compared to generic MEA need to be taken into consideration.

Upstream of the CO_2 capture process, there will be a need to cool and boost the pressure of the flue gas stream. Flue gases are delivered at atmospheric pressure at the gas turbine vent stacks, so pressure boosting using an industrial fan allows is used to overcome the pressure drop in the flue gas cooling and absorption column. As far as practical, waste heat can be recovered from the flue gas for use in CO_2 capture process to increase overall energy efficiency and reduce CO_2 emissions associated with solvent regeneration. A direct contact cooler column using a recirculation of condensed water is used to cool flue gases to the optimal temperature for stable operation of the CO_2 absorption process.

The CO_2 product is delivered at low pressure and water saturated and therefore requires compression and dehydration. This downstream equipment is considered to be outside of the core CO_2 capture process.

A6.1.2. Ammonia

In chilled ammonia processes, the flue gas is precooled to a temperature of about 0 to 10° C, such that refrigeration is likely to be needed. The cooled gas is then introduced to the absorber column where the CO₂ in the flue gas is reacts with an ammonium carbonate aqueous solution (~28 wt%) to produce ammonium bicarbonate. The low operating temperature promotes the forward reaction producing ammonium bicarbonate as it causes the product to precipitate out of solution due to its limited water solubility. The process can be operated with no solids production but this reduces the CO₂ loading.



The absorption process is carried out in a spray column to allow free flow of precipitated ammonium bicarbonate. Process development has incorporated the use of proprietary packed columns.

The solvent leaves the absorber in slurry form which is pumped and concentrated and sent to the regeneration unit where it is heated to about $80-100^{\circ}$ C to release CO₂ which pressurises the system. Regeneration is carried out at high pressure of 20-40 bara, producing a high purity CO₂.

Overall, the process flowsheet is perceived to be more complex than the aminebased absorption processes with more equipment involved (see Figure A6.2). The technology has been extensively tested at pilot scale (Ref 41).



Figure A6.2. Chilled Ammonia Process

The heat of absorption of carbon dioxide by ammonia is significantly lower than for amines, so solvent regeneration heat requirements are lower in comparison to regeneration of amine-based solvents, with reported SDR in the order of 2 GJ per tonne of captured CO_2 Furthermore, the regeneration temperature is low (Ref 29). Furthermore, since the regeneration is performed at high pressure, the downstream compression requirement could potentially be reduced for the CO_2 stream downstream of the CO_2 capture plant.

It is reported that the solvent is more stable than amines, with impurities such as SO_x and NO_x not affecting the performance, with potential for simultaneous removal of these impurities.

Main concerns on the process performance are related to the loss of solvent as ammonia in the treated gas stream. The use water wash and evaluation of additives to reduce evaporative ammonia losses has been considered. Also, the refrigeration load required to cool process streams to an optimally low operating temperature has a significant energy penalty.



A6.1.3. Hot Potassium Carbonates

There are a number of processes available that use aqueous potassium carbonate solution to absorb CO_2 such as the Benfield, Hi-Pure and Catacarb processes (Ref 33) used for the purification of natural gas and synthesis gas. Like all carbonatebased capture systems, these processes are based on the reversible reaction of dissolved CO_2 and carbonate to form bicarbonate. The reaction is benefited from operation at high temperatures, although this reduces the CO_2 solubility in the solution which limits the purity of gas that can be achieved and requires a pressurised feed gas.

The process flow-schemes are similar to amine-based processes with an absorber column and a stripper column for regeneration (see Figure A6.3). The absorber operates at around 100°C which may provide benefits for treating hot flue gas from turbines. CO_2 is absorbed at high pressure (~20 bara) and desorbed at or near atmospheric pressure at roughly the same temperature and aided by steam stripping. Unlike the amine based processes, the lean solvent is not cooled as both columns operate at a similar temperature.



Figure A6.3. Hot Potassium Carbonate Process (Ref 33)

Carbonate salts have lower environmental risks compared to amines and additives; they are inexpensive and widely available, non-volatile and resistant to degradation. Additives or catalysts may be used to improve absorption and desorption rates, reduce corrosion and improve resistance to oxidation.

It is reported that typical feed conditions for the hot potassium carbonate process are between 10 and 125 bara with between 5 and 35 mol% CO_2 content (Ref 34). This suggests a low potential for the application of this process in post-combustion CO_2 capture due to the low CO_2 content and low flue gas pressure.

Research on carbonates for CO_2 capture from power plants has focused on dry processes using solid sodium carbonate or carbonates impregnated in solid supports. Technology development for post-combustion CO_2 capture application can be considered to be at a very early stage of research and development.



A6.1.4. Caustic Wash

Caustic solutions can be used to remove CO_2 from natural gas streams through counter-current contacting of the gas stream with a caustic solution in a contactor column.

A non-regenerable product, sodium carbonate, is produced by the caustic - CO_2 reaction which must be removed from the system and disposed of. The non-regenerable nature of the caustic wash process when used in CO_2 capture schemes results in low potential for this application, as it would result in very high caustic consumption due to the large quantities of CO_2 to be captured and disposal issues for the spent solution which is considered as hazardous waste.

A6.1.5. Amino Acids

In response to concerns regarding potential amine losses in treated gas being vented, susceptibility to degradation and corrosion problems, amino acids such as glycine and alanine have been evaluated. These are non-volatile compounds and have performance comparable to amines. Amino acids offer better resistance to degradation in the presence of oxygen although their performance is reduced when oxygen is present (Ref 31), but developed variations are reported to lead to performance improvements (Ref 32).

A6.2. PHYSICAL ABSORPTION

A6.1.6. Physical Solvents

In the physical absorption process the gas is contacted with a solvent which physically absorbs the CO_2 using partial pressure as the driving force. These processes are adequate when the partial pressure of CO_2 is greater than about 3 bara, which equates to a CO_2 concentration of more than about 5 mol% for gas at high pressure of typically 70 bara in gas processing applications (for some applications operating pressure could be in the order of 35-40 barg). Physical solvent processes ideally below ambient temperature to increase the CO_2 solubility.

The solvent is regenerated by:

- Multi-stage flashing to low pressures
- Regeneration at low temperatures with an inert stripping gas
- Heating and stripping of solution with steam or solvent vapours

Some of the physical absorption solvents processes are refrigerated methanol (Rectisol, Ifpex-2), polyethylene glycol (Selexol), N-methyl-2-pyrrolidone (Purisol), anhydrous propylene carbonate, N-formyl morpholine/N-acetylmorpholine (Morphysorb).

When the process conditions act in favour of physical absorption, these processes are economically attractive because of the relatively low energy requirements for solvent regeneration. Physical solvents are primarily used for bulk CO_2 removal and for selective removal of H_2S or sulphur compounds (COS, CS_2 and mercaptans). The CO_2 loading capacity of physical solvents is low which may lead to recirculation rates up to three times that required by amine processes, which increases relative equipment sizes.





Figure A6.4. Typical Schematic for Physical Absorption Process (Ref 33)

Physical absorption does not show a significant potential for post-combustion CO_2 when compared to chemical (amine-based) absorption due to the low partial pressures of CO_2 in flue gas (even with compression at high pressure) and the low CO_2 loading capacity of physical absorbents.

A6.2.1. Ionic Liquids

An alternative for of physical solvents can be achieved through the use of ionic liquids. Ionic liquids have been widely used in catalysts and synthesis and can be applied for CO_2 absorption. They offer unique properties such as low vapour pressure, thermal stability, high polarity and non-toxicity (Ref 26).

It is reported that the structure of an ionic liquid can be modified to create a taskspecific ionic liquid by altering the amino-function group which reacts with CO_2 . Task-specific ionic liquids have around three times the CO_2 absorption capacity compared to a generic ionic liquid (Ref 27). Supporting ionic liquids on the surface of silica gel is reported to provide CO_2 capture capacity improvements of around 6 times the capacity of an ionic liquid alone (Ref 28) through the provision of a high surface area and allowance for physical absorption on the surface.

Lab scale investigations into the use of CO_2 capture have been conducted (Ref 44) showing some potential. However, the technology is considered to be at an early stage of development. Performance issues related to their low CO_2 loading capacity, high viscosity (about 5 times the viscosity of MEA) and relatively high costs need consideration (Ref 29, 44).

A6.3. HYBRID ABSORPTION

Hybrid solvents can offer the benefits of both chemical solvents (low outlet acid gas concentrations) and physical solvents (lower energy requirements). Sulfinol are hybrid solvent formulations used in CO₂ removal duty. Sulfinol is a mixture of sulfolane (a physical solvent), water and either DIPA or MDEA and Piperazine.

Hybrid solvents are generally more expensive than conventional amine solvents. The capital and operating costs (CAPEX and OPEX) of the hybrid solvent processes tend to be higher than those of amine solvent processes.

A6.4. ADSORPTION

Physical adsorbent systems use a solid adsorbent media with high surface area to selectively remove CO_2 from the feed gas. Physical adsorbents have been used for CO_2 removal due to their availability, cost, thermal stability and their low sensitivity to moisture (Ref 26). There are various types of physical adsorbents for use in CO_2 capture such as activated carbon, zeolites, mesoporous silica and metal organic frameworks, each with varying CO_2 capture capability, and operating temperatures and pressures.

Chemical adsorption involves a chemical reaction that takes place on the surface of an adsorbent which can then be reversed to release the adsorbed component and regenerate the adsorbent. Chemical adsorbents which may be used include; amine-based, metal oxides (e.g. calcium and magnesium oxides), alkali metals, hydrotalcites and double salts.

It is reported that amine-based adsorbents have an advantage over aqueous amine solvents due to their low heat of regeneration associated to the low heat capacity of the solid supports. Amine-based adsorbents have low CO₂ capacity and expensive compared to other alternatives. Two types of amine-based adsorbents exist: amine-impregnated and amine-grafted adsorbents, offering varying levels of CO₂ loading, rates of adsorption and stability in cyclic operation (Ref 26).

A key consideration in the selection of an adsorption process is the finite capacity of the adsorbent. Over time, adsorbents (physical or chemical) become saturated with CO_2 and need regenerating. The way adsorbents are regenerated differs depending on the individual application. However, regeneration routes are similar as they all involve a change in operating conditions to prompt the desorption of CO_2 . Most applications use multiple adsorption beds operated in a cyclic batch-wise manner. At least two beds are required for continuous operation, with one operating in adsorption mode and the other in regeneration mode. The modes will then switch when the capacity of adsorbent in the bed in adsorption service is reached.

The main methods for adsorbent regeneration are:

- Temperature Swing Adsorption (TSA)
- Pressure Swing Adsorption (PSA)

Temperature Swing Adsorption uses temperature as the driving force for regeneration of the adsorbent. During adsorption, the bed will operate at a low temperature. During regeneration, the bed is purged with a regeneration gas at high temperature. Heating and cooling of the media and equipment is often a slow process which prolongs the operational cycle time. TSA relies upon a hot regeneration gas to raise the bed temperature. This is suitable for applications such as pre-treatment of natural gas in LNG peak shaving facilities where the



CO₂ rich regeneration gas can be used as a source of fuel or put back into the main gas transmission system.

An alternative form of TSA adsorption process has been developed for CO_2 capture from flue gas using a potassium-based solid adsorbent as shown in Figure A6.5 (Ref 40). Flue gas is treated in a fluidised bed where CO_2 is captured by a chemical reaction converting potassium carbonate to bicarbonate. The CO_2 rich adsorbent is separated from the treated flue gas in a cyclone and is regenerated in a side regeneration bed. Hot gas consisting of CO_2 and steam is used for regeneration of the adsorbent before it is recycled back to the adsorption column. The process captures in excess of 80% of the CO_2 in the flue gas, producing CO_2 with 95 mol% purity. This technology has been demonstrated at pilot scale.



Figure A6.5. Schematic of Potassium Carbonate Adsorption Process (Ref 40)

Pressure Swing Adsorption (including Vacuum Swing Adsorption, VSA) use pressure as the driving force for adsorption and regeneration. During adsorption, the bed will operate at high pressure. During regeneration, the pressure is reduced which causes desorption of the CO₂ from the adsorbent. There is no specific requirement for regeneration gas to be used in PSA. The cycling of operating pressure in PSA processes is faster (in the order of minutes) than in TSA, which makes it more suited to bulk separation.

In general, PSA technologies require relatively low energy and low CAPEX. The feed to the process is typically in the order of 7 bara to produce a treated gas stream and a CO_2 rich stream at just over atmospheric pressure. The technology has been applied on small scale LNG pre-treatment facilities. PSA using Molecular Gate technology has been demonstrated for pre-treatment of gas from up to 40 mol% to 50 ppm CO_2 (Ref 36).

There are adsorbents (molecular sieve 13X, APG III and 4A, activated carbon) and processes commercially available for the specific purpose of CO_2 removal in gas processing and air separation (Ref 38, 39).



Molecular sieves have the ability to remove CO_2 but are typically used for removal from low CO_2 concentration of about 2 mol% in the feed gas down to a concentration of around 50 ppm. Molecular sieves offer obvious advantages due to their ability to remove CO_2 to a very low level. However, they typically have low CO_2 capacity and require large regeneration flowrates for removal of large amounts of CO_2 which results in the size of the system increasing. It is reported that the size of molecular sieve systems becomes unreasonably large when used for a 1 mtpa feed gas flow containing less than 0.2 mol. % CO_2 (Ref 6). Molecular sieve adsorbents are proven in operation for to remove CO_2 from pipeline gas which is subsequently liquefied in LNG peak shaving facilities (Ref 37).

Opportunities to combine molecular sieve technology with amine absorption have been considered and are reported to have the ability to reduce feed gas CO_2 concentration from around 10 mol% to 50 ppm (Ref 34).

A6.5. MEMBRANES

Semi-permeable membrane processes provide separation based on varying rates of permeation of different gas components through polymeric membrane films. The feed gas at moderate to high pressure is passed thorough the membrane system to produce a CO_2 rich permeate stream and a CO_2 lean treated gas stream. Depending on the selectivity of membranes and the required CO_2 recovery, a two-stage process and recycle is usually required. As the permeate gas diffuses across the first membrane a considerable pressure drop occurs, therefore permeate gas needs to be compressed and cooled before being fed to a second membrane stage (see Figure A6.6).





Maximisation of the pressure differential across the membrane is used to improve the separation rate, with permeate produced at low pressure and potentially requiring recompression, particularly if a portion of this is to be recirculated. A significant recirculation rate may be required to increase the purity and the recovery of the products. The main energy input to the process is in the form of power to compress feed gas or recompress the product.

High recovery of CO_2 requires large membrane areas. This can be provided by the use of multiple membrane modules, but this does not offer substantial economy of scale benefits and the anticipated plot size required is likely to be substantial.

For feed gases with low CO₂ content and/or when high recoveries are required, the recycle rate increases which increase both capital and operating costs due to increase in equipment size (a larger number of membrane modules to provide the required membrane surface area) and compression power requirements, which in turn increases emissions.



Pre-treatment of the gas including removal of solids, water and heavy hydrocarbons and temperature control may be required in order to protect the expensive membrane elements. A relatively large proportion of the operating costs of a membrane unit are in the replacement of membrane units on an ongoing basis.

Membrane processes offer advantages such as not introducing water to the gas stream, has a reduced number of equipment items when compared to solvent systems, and eliminates the need for regeneration, handling and disposal of solvents. Membranes allow faster startup and have lower maintenance (no moving parts, excluding compression) relative to amine systems.

Membranes may be better suited than other CO₂ removal technologies to plants that are situated in remote locations (e.g. offshore) due to complexities in the delivery of solvents and the possibility that the plant may be unmanned.

Membranes are best suited to bulk removal where CO_2 content in the feed gas is relatively high (indicatively 10 mol% and higher) and may only achieve low outlet concentrations if used in conjunction with another process. The combinations of membrane and amine technologies are reported to potentially improve removal efficiencies, increase flexibility and robustness to changes in CO_2 content when compared to the individual technologies (Ref 34). However, this combination still remains suited to feed gas with a high CO_2 content. The combination also increases complexity in both design and operation due to the combination of technologies.

The use of membranes has been evaluated as an option for CO_2 capture. However, the technology for this particular application is at a very early stage of development. The process shows limited potential by the technical limitations noted above, particularly for post-combustion capture where the flue gas would require compression and the CO_2 content is very low. Furthermore, the flue gas flow rates involved in full scale capture plants would require an extremely large membrane plant (footprint).

A6.6. CRYOGENIC SEPARATION

Cryogenic separation consists of a low temperature distillation where gas streams are partially condensed by cooling provided by refrigeration, and the components separated by fractionation. Cryogenic separation is a commercial process commonly used to liquefy and purify CO_2 from sources with a high CO_2 content.

The performance of cryogenic processes improves at elevated feed gas pressures, as condensation (liquefaction) is facilitated.

In terms of capital costs, cryogenic processes benefit from economies of scale. Advantages in terms of operating costs include the production of a liquid CO₂ ready for pumping and transportation by pipeline. This reduces power and cost requirements for CO₂ compression.

The major disadvantages of this process are the significant energy required to provide the refrigeration duty and the necessary removal of components (particularly water and hydrocarbons) that could freeze at the low operating temperatures and potentially cause blockage of process equipment.

Design and operating problems associated with cryogenic distillation are usually related to CO_2 solidification in the cryogenic equipment. This can be solved in natural gas processing by adding ethane and heavier hydrocarbons to the column to increase CO_2 solubility in a process called Ryan Holmes fractionation. This technology is currently in use in several

commercial operations, including for the separation of CO_2 from gas mixtures in EOR operations. A technology called Controlled Freeze Zone has been developed and tested at full scale for removal of CO_2 from natural gas streams containing as little as 8 mol% CO_2 . In this process, CO_2 is allowed to freeze in a specially designed section of a fractionation column. Solids are then be melted and fractionated to strip light hydrocarbons so as to produce liquid CO_2 product at elevated pressure (Ref 35).

In terms of CO₂ capture, cryogenic separation may show some potential for high pressure gases with high CO₂ concentration in pre-combustion processes. However, for post-combustion CO₂ capture flue gases are available at near atmospheric pressure and CO₂ is highly diluted, which makes cryogenic processes less economical than others in separating CO₂.

A6.7. EXHAUST GAS RECYCLE

Exhaust gas recycle (EGR), sometimes referred to as flue gas recycle (FGR), has potential to be an effective way to reduce energy requirements for the capture of CO_2 in postcombustion capture schemes. Fundamentally, use of an EGR scheme allows increasing the CO_2 concentration in the flue gas, which improves the CO_2 removal performance, particularly of a chemical absorption process.

In an EGR scheme, a portion of the flue gas produced from the combustion of fuel in a gas turbine to generate power is cooled and recycled to the inlet of the gas turbine compressor (See Figure A6.7). The recycled flue gas replaces a portion of the air which is normally fed to the turbine compressor. By replacing the air with the recycled flue gas an accumulation of CO_2 occurs within the system which results in a higher concentration of CO_2 in the flue gas fed to the CO_2 capture plant. The solvent flow rate and the specific stripper reboiler duty are both reduced, leading to a potential reduction in costs.





In some applications, EGR has been used for the purpose of controlling NO_x emissions. Cold exhaust recirculated gases reduce the flame temperature and a reduction of the oxygen content which slows the NO_x forming reaction. Use of EGR has been studied for use in precombustion capture to reduce the formation of environmentally damaging NO_x in the flue gas by diluting the hydrogen fuel for the turbine with nitrogen (Ref 47).

It is noted that if the recycled flue gas is at a temperature higher than the turbine compressor design, it will impact the power output and efficiency of the gas turbine. The recycled exhaust gases are cooled as much as practicable, which impact both capital and operating costs of the overall scheme.

The EGR ratio is ultimately limited by the requirement to supply sufficient oxygen for combustion in the gas turbine as oxygen concentration is reduced when EGR ratio is increased. Recycle rates of 35% of the flue gas have been demonstrated in experimental gas turbine systems, increasing the CO_2 concentration in the flue gases from 4 mol% to 7 mol%.

A number of trade-offs need to be considered for increased EGR ratios, which include the impact on the gas turbine performance. As a result, the overall efficiency improvement and the associated operating cost savings are likely to be in the order of less than 1% at best (Ref 48). The greater improvement in performance is particularly realised when the gas turbine is run at low load. Larger benefits are obtained from reductions in the capital costs, potentially in the order of 7% (Ref 48). It is reported that recycle rates of 50% may reduce the capital cost of the CO₂ capture plant above this (Ref 46). Therefore, EGR is a strategy that can be considered to potentially reduce the size of the CO₂ capture plant and reduce the capital investment.

This technology is considered to be provide incremental benefits, contributing to the feasibility of CO_2 capture technologies by potentially reducing capital costs, Technical feasibility and risks need to be considered against the potential capital cost gains.


		_	Sweet Feed Gas End Flash Gas					
Composit	ion	Raw Feed Gas	Eucl Coo	Flue Gas (Note 2)	Eucl Coo	Flue Gas (Note 2)	
			Fuel Gas	Aeroderivative	Industrial	ruel Gas	Aeroderivative	Industrial
Nitrogen	mol%	0.89%	0.91%	74.30%	74.61%	17.74%	74.23%	74.55%
Oxygen	mol%	0.00%	0.00%	13.64%	14.55%	0.00%	13.58%	14.49%
Argon	mol%	0.00%	0.00%	0.88%	0.89%	0.00%	0.88%	0.89%
Carbon Dioxide	mol%	2.00%	0.00% (Note 1)	3.25%	2.83%	0.00%	3.21%	2.79%
Methane	mol%	89.00%	90.81%	0.00%	0.00%	82.25%	0.00%	0.00%
Ethane	mol%	7.00%	7.14%	0.00%	0.00%	0.01%	0.00%	0.00%
Propane	mol%	1.00%	1.02%	0.00%	0.00%	0.00%	0.00%	0.00%
Butane	mol%	0.10%	0.10%	0.00%	0.00%	0.00%	0.00%	0.00%
Pentane	mol%	0.01%	0.01%	0.00%	0.00%	0.00%	0.00%	0.00%
Water	mol%	0.00%	0.00%	7.92%	7.12%	0.00%	8.10%	7.28%
Mol Weight	kg/kgmol	18.02	17.49	28.39	28.44	18.17	28.36	28.42
LHV	kJ/kg		48,880			36,348		
HHV	kJ/kg		54,156			40,372		

Notes:

- 1. CO₂ content reduced to 50 ppmv in AGRU
- 2. Based on combustion calculations considering air flows for PGT25+ (representative of aeroderivative gas turbines) and Frame 5D (MS5002D) as (representative of industrial gas turbines) based on average ambient conditions, see below.

Ambient Conditions	Temp ⁰C	RH, %
Maximum	43	90.0
Average	21	74.7
Minimum	-15	40.0

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Economic Evaluation of CO₂ Capture in LNG Plants

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1.0 ABBREVIATIONS

AGRU	Acid Gas Recovery Unit
APR	Annual Percentage Rate
bara	Bar (atmospheric)
C3MR	Propane Precooled Mixed Refrigerant Liquefaction Process
CAPEX	Capital Expenditure
CCS	CO ₂ Capture and Storage
CHP	Combined Heat and Power
СО	Carbon Monoxide
CO ₂	Carbon Dioxide
EFG	End Flash Gas
EOR	Enhanced Oil Recovery
FOAK	First Of A Kind
FLNG	Floating LNG
GHG	Greenhouse Gas
GJ	Giga Joule
H ₂ S	Hydrogen sulphide
HHV	Higher (gross) Heating Value
HRSG	Heat Recovery Steam Generation
IEA	International Energy Agency
ISO	International Organization for Standardization
kg	Kilogram
kW	Kilo Watt
kWh	Kilo Watt Hour
LNG	Liquefied Natural Gas
m	Meter (distance)
MEA	Mono-Ethanol Amine
MJ	Mega Joule
MMSCFD	Million Standard Cubic Feet per Day
mol	Molar
MR	Mixed Refrigerant
mtpa	Millions of tonnes per annum
MW	Mega Watt
MWe	Mega Watt Equivalent
MWh	Mega Watt Hour
NOAK	n th Of A Kind
OPEX	Operating Expenditure
PCC	Post-Combustion Capture
ppm	Parts per million
ppmv	Parts per million (volume basis)
SO ₂	Sulfur Dioxide
UK	United Kingdom
US	United States (of America)
WHRU	Waste Heat Recovery Unit
wt	Weight (mass)



2.0 EXECUTIVE SUMMARY

This report summarises the economic assessment of a capture, compression and purification of CO₂ streams from two main sources of a typical baseload LNG facility:

- 1. CO₂ that is separated in the Acid Gas Removal Unit (AGRU) as part of the pretreatment of the feed gas to the liquefaction plant.
- 2. CO₂ that is produced from combustion of fuel gas to produce power required by the liquefaction plant (gas turbines in mechanical drive and power generation service).

The base case LNG plant under consideration is representative of baseload LNG liquefaction train designs and capacities, i.e. 4.6 mtpa of LNG (C3MR process using 2 x Frame 7 refrigerant compressor drives).

A summary of the key results of the economic evaluation for the base case is given below:

- The estimated Total Plant Cost (TPC) is estimated as €597.5 million and the total capital requirement (TCR) is approximately €705 million.
- Total OPEX requirement (discounted NPV) for the project over the life of the plant (25 years) is estimated as €567 million.
- The specific cost of CO₂ captured is €47.3/t CO₂ of which 56% is associated with the CAPEX requirements and the remainder associated with the OPEX (over 25 years). The specific cost of CO₂ avoided is €55.2/t CO₂
- The minimum cost of CO₂ emissions (CO₂ tax) would have to be at least €129/t CO₂ to justify the installation of CCS (based on CO₂ emission costs against CO₂ capture costs only).

Sensitivity analyses were performed on a number of variables used in the economic analysis. The key outcomes of the sensitivity analyses performed are:

- Sensitivity analyses performed show a range in capture costs of about €13/t CO₂ to €57/t CO₂, avoidance costs in the range €14/t CO₂ to €78/t CO₂. The lowest specific cost is associated with capture of AGRU CO₂ emissions only (flue gas emissions vented) with a high CO₂ content (14 mol%) in the reservoir/feed gas.
- NOAK plant designs increase attractiveness of CCS due to the ability to reduce CAPEX costs compared to FOAK design. An expected CAPEX reduction of up to 30% is reported to be achievable, resulting in a reduction of capture cost to about €37/t CO₂, avoidance cost about €43/t CO₂.
- It is assumed in the base case that there will be no running costs associated with the
 production of steam since the excess heating is expected to be available in LNG
 facilities through waste heat recovery. This ability to provide heat duty to the CCS plant
 allows for a capture cost savings of about €9/t CO₂ plus a reduction in emissions
 associated to burning fuel to produce steam.



- Large scale LNG trains (such as those found in Qatar, with capacity of 7.6 mtpa LNG) may provide greater benefits for CO₂ capture as a result of economies of scale. The total capture cost for plants this size is reduced to about €42/t CO₂, with avoidance cost reduced to about €48/t CO₂.
- The CO₂ emission cost (i.e. CO₂ tax) required to justify implementation of post-combustion CCS schemes in LNG plants purely on economics, i.e. reduction of costs paid for emitted CO₂, is in the order of €100/t CO₂. Current world emission policies set CO₂ tax at a relatively low value (if any), with the majority of emissions currently priced at less than about €10/t CO₂. This indicates that implementation of post-combustion CCS would only occur for either significant CO₂ tax increases or by drivers other than plant economics. However, when the CCS scheme only considers sequestration of the CO₂ that separated from the feed gas in the AGRU, the minimum CO₂ tax required is in the order of €30/t CO₂. This level of CO₂ tax is within current environmental policies in some regions (e.g. Norway and Finland), which indicates the potential of this route for the implementation of CCS.



3.0 BACKGROUND

Natural gas demand is forecasted to grow continuously for the next 10 years, playing a vital role in the global energy mix in 2030. In the particular case of Liquefied Natural Gas (LNG), projections indicate a continued considerable growth. The majority of near-term growth in liquefaction capacity is likely to happen in North America and Australia, although a number of other projects have the potential to add significant liquefaction capacity in the long term as well.

As a key contributor to the global energy supply, the LNG supply chain is expected to be subject to global requirements on reduction to greenhouse gas (GHG) emissions. While the majority of emissions in the LNG supply chain are at the end user (i.e. power plant), the LNG liquefaction plants produce a significant proportion of the total chain CO_2 emissions. The majority of the CO_2 emissions from LNG liquefaction plants arise from combustion of fuel for cycle compressors drive and power generation purposes. It is also necessary to remove and dispose of CO_2 from the natural gas feed, to avoid solidification in the liquefaction process.

The CO₂ emissions depend on the liquefaction plant configuration (e.g. feed gas pre-treatment, liquefaction technology, heat and power integration) and also local ambient conditions. The LNG industry has already improved the overall thermal efficiency of the LNG supply chain to reduce the associated CO₂ emissions. LNG plants may provide an early opportunity for CO₂ capture and storage (CCS) subject to the availability of high-concentration CO₂ streams and vicinity to storage reservoirs.

Based on this, the International Energy Agency (IEA) GHG programme considered necessary to evaluate the techno-economics of different CO₂ capture technologies in LNG plants in detail.

4.0 SCOPE

IEAGHG commissioned Costain to develop a techno-economic evaluation and cost assessment of different CO₂ capture technologies for LNG plants including for small scale and floating (FLNG) plants.

The scope of work comprises three main tasks:

- 1) Technical evaluation of different CO₂ capture technologies for LNG plants
- 2) Cost assessment of different CO₂ capture technologies for LNG plants
- 3) Feasibility study of different CO₂ capture technologies at small scale and FLNG Plants

This report summarises the evaluations developed for Task 2.

Cost estimates have been developed for capture, compression and purification for subsequent transport and storage of CO_2 separated from gas turbine flue gas via post-combustion using a chemical absorption process. The base case also considers compression and purification for transport and storage of CO_2 captured in the feed gas pre-treatment facilities (Acid Gas Removal Unit).

The LNG plant upon which the base case is based is representative of baseload LNG train designs.

Estimates of capital and operating costs (CAPEX and OPEX) are presented in this report for the base case and a number of sensitivity cases.



5.0 BASE CASE

5.1 Basis of Design

Basis of Design are included in Appendix 1.

5.1.1 Liquefaction Plant Size

The base case LNG plant under consideration is representative of baseload LNG liquefaction train designs and capacities, i.e. 4.6 mtpa of LNG (C3MR process using 2 x Frame 7 refrigerant compressor drives).

5.1.2 Power Generation

For the base case, it is considered that the power generation output equates to about 25% of the cycle power or 9 MW/mtpa of LNG, as typical/representative of baseload LNG plants:

- Little to moderate feed gas pre-treatment low to moderate acid gas content (CO₂, H₂S and sulphur compounds) in feed gas, in the order of a few percent.
- Lean feed gas requiring minimum to moderate processing (and fractionation) of hydrocarbon liquids (NGL and condensate).
- No feed gas compression required as part of the liquefaction train.
- No nitrogen rejection required.
- Power plant does not support other operations (offshore, other plants in the site)

In addition to this, it is assumed that an industrial gas turbine (Frame 6 or similar) in simple cycle is installed as part of the CO_2 capture plant scope to supply with power to the CO_2 capture, compression and conditioning plant.

5.1.3 Process Heating

In typical baseload liquefaction plants, gas turbines (in mechanical drive or power generation service) are not fitted with heat recovery steam generators (HRSG) as part of integrated cogeneration or combined cycle power generation schemes. The available heat from waste heat recovery is well in excess of the heating requirements of liquefaction trains. Therefore, waste heat recovery units are not installed in all of the gas turbines.

The base case considers that the entire process heating demand from the liquefaction train (including pre-treatment) and processes associated with the capture and purification of CO_2 (thermal regeneration of chemical absorption solvent and CO_2 dehydration media) can be supplied by waste heat recovery from gas turbines. Therefore, no additional fuel gas (e.g. in a dedicated boiler or fired heater) is required to provide heating.

Installation of waste heat recovery units in turbine exhausts will impact the efficiency of the gas turbines due to increased backpressure. As gas turbines operate at full load, this represents a reduction in power output and therefore liquefaction capacity. However, it is assumed that the gas turbines are mainly de-rated by temperature, and no detailed de-rating has been applied for exhaust pressure losses (as in any case this is expected to be relatively small about 1% loss in shaft power, see Section 5.3.1)



in consideration of the level of evaluation/estimate accuracy. The assumed de-rating of gas turbine performance is assumed to include any impact due to additional backpressure in turbine exhausts.

5.1.4 Liquefaction Plant Feed Gas

A feed gas composition representative of LNG plants is included in Appendix 1. CO_2 content for the base case is 2 mol% and negligible content of H₂S and sulphur compounds.

5.1.5 CO₂ Emissions

The base case for this assessment considers the capture, compression and purification of CO_2 streams from two main sources:

1. CO₂ that is separated in the Acid Gas Removal Unit (AGRU) as part of the pre-treatment of the feed gas to the liquefaction plant.

The CO₂ produced by the feed gas pre-treatment stage is separated as to meet the LNG liquefaction requirements (CO₂ removed to avoid freezing leading to blockages in the liquefaction equipment). CO₂ is separated from the feed gas together sulphur compounds (mainly H₂S). In absence of the later (as per feed gas composition considered for the base case), the stream has a CO₂ high purity, saturated with water and containing a relatively low content of hydrocarbons (absorbed in the solvent and released during regeneration). The processing of this stream is limited to compression and purification prior to transportation to storage.

2. CO_2 that is produced from combustion of fuel gas to produce power required by the liquefaction plant.

 $\rm CO_2$ emissions from gas turbines in mechanical drive and power generation service. The flow and composition of the flue gases are estimated based on the following:

- Fuel gas used in gas turbines is assumed to be taken from feed gas following acid gas removal treatment. Fuel gas lower heating value is 48.88 MJ/kg.
- Output power and efficiency (de-rated) of gas turbines in mechanical drive and power generation associated to the liquefaction process.
- Gas turbine exhaust flows (for the specific models under consideration).

Post-combustion CO_2 capture with chemical absorption is used to treat the flue gas from the exhaust of liquefaction compressor gas turbine drives and from the gas turbines used in power generation service.

The base case considers that the exhaust gases produced by the dedicated gas turbine supplying power to the CO_2 capture process are not treated, and emissions released to atmosphere. The impact of including this flue gas stream in the feed to the capture plant is evaluated in a sensitivity case.

The combined flue gas flow rate from all gas turbines (with the exception indicated above) is 2,630 t/h (for 4.6 mtpa LNG plant). Composition is 3.2 mol% CO₂, 13.8 mol% Oxygen, 7.8 mol% water (the balance is mostly nitrogen). It is



assumed that the level of nitrogen and sulphur oxides, particulates and other impurities is low so no further gas treatment is required prior to passing the flue gas to the CO₂ conditioning (cooling) and capture process. The flue gas stream is at near atmospheric pressure and the combined temperature assumed 200°C (following waste heat recovery).

5.1.6 CO₂ Capture Process

Chemical absorption is the technology selected for capture of CO₂ from flue gases. It is assumed that a proprietary solvent formulation is used, allowing improved performance (reduced energy requirements for solvent regeneration, reduced solvent degradation and losses, compared to generic MEA solvent

Proprietary solvents have been tested (to treat flue gases with a similar composition from CHP plants) and are commercially available to cope with the relatively high oxygen content in terms of solvent degradation.

CO₂ capture process performance:

- CO₂ removal: 90% of CO₂ in flue gas
- Specific Reboiler Duty: 3 GJ/tCO₂ captured (energy required for solvent regeneration)
- Solvent Make-up: 0.2 kg/tCO₂ captured

5.1.7 CO₂ Compression and Purification

The captured CO_2 is combined with the CO_2 from pre-treatment (AGRU) for combined compression and purification (dehydration) prior to transportation to storage.

- The CO₂ stream from the AGRU is water saturated. A nominal 1 mol% hydrocarbon content is included.
- The CO₂ stream from the capture plant is water saturated.

Both streams are at near atmospheric pressure and temperature near ambient (discharged from air coolers).



5.2 CO₂ Capture Scope

Figure 1 shows a Block Flow Diagram representing the scope of facilities included in the economic evaluation of the base case.



Figure 1. CO₂ Capture Scope

For the purpose of this study, only the capture, compression and purification facilities (and associated power generation plant) are included in the capital cost estimate. Cost estimates do not include the capital cost associated to the transport and storage infrastructure (project costs excluded). A nominal specific cost to cover operating costs of transport and storage per tonne of captured CO_2 are included (see Appendix 2).

The scope of the project includes all interfaces with LNG facilities and utilities, including ducting to transport flue gases from the sources to the capture plant and all required piping tie-ins and electrical connections.

In estimating project costs, it is considered that the CCS scope is implemented as a standalone project (e.g. installed as a plant retrofit), and therefore it does not benefit of cost savings of combined projects (i.e. if the CCS scope was part of the liquefaction plant construction project).

It is considered that the LNG plant is capture-ready to some extent with no significant modifications to the LNG plant being required to accommodate the CCS scheme, i.e. brownfield modifications limited to the installation of tie-ins with minimum modifications and plot plan and utilities (excluding power) are available. This is an assumption taken for simplification of this evaluation. It is recognised that a CCS readiness is a relatively recent feature and that all the existing LNG plants have not been constructed with capture-ready design allowances. The ultimate extent of brownfield modifications will depend on the particular plant circumstances (e.g. feasibility and extent of modifications to fit tie-ins and piping/ducting, availability of spare capacity in utility systems, availability of plot area for the capture plant, etc.).



5.3 **Process Description**

5.3.1 General

CO₂ is recovered in a post-combustion capture scheme from the flue gas delivered at the exhaust of gas turbines used in mechanical drive service for the liquefaction cycle refrigerant compressors and from the gas turbines used for power generation.

The CO_2 capture plant operates as a standalone unit such that the LNG liquefaction process is independent of the capture plant. It is assumed that unavailability of the capture plant will not impact on the production of LNG. The flue gas can either be routed to the individual stack on each gas turbine (bypassing the capture plant) or to the collection ducting to the capture plant.

When flue gas is routed to the capture plant there may be a slight reduction in the performance of the gas turbine due to increased backpressure on the gas turbine discharge due to additional pressure losses in the system transporting flue gases to the capture plant. For reference, the resultant reduction in performance may be expected to be comparable to the impact of installing a waste heat recovery unit. For example, performance maps for an LM6000PD gas turbine show a reduction in shaft power of about 1% for an exhaust pressure loss of 20 to 30 mbar.

5.3.2 Flue Gas Compression and Cooling

Flue gas supply to the CO₂ capture facility is available at atmospheric pressure and relatively high temperature (200°C) with respect to the operating temperature required by the capture process (40°C). Upstream of the CO₂ capture process, the flue gas is boosted in pressure and then cooled to a temperature optimum for the capture (absorption) process.

A booster fan increases the pressure of the flue gas to ensure there is sufficient driving force to overcome pressure drops across the capture plant.

Downstream of the booster fan, the flue gas is fed to a direct contact cooler which uses cooled recirculating water to cool the flue gas stream to the optimum temperature for operation of the CO_2 absorption process and to remove impurities. Any accumulation of water due to the cooling of the flue gas is removed from the recirculating system.

5.3.3 CO₂ Capture

Downstream of the direct contact cooler, the cooled flue gas is fed to an absorber column where 90% of the CO_2 in the flue gas is removed to produce a treated CO_2 lean flue gas stream which is released from the top of the absorber column to atmosphere. A CO_2 rich absorbent stream leaves the bottom of the absorber column which is pre-heated and fed to a stripping (regeneration) column.

The solvent is heated in the regeneration column to promotes the desorption of CO_2 from the solvent. A high purity CO_2 stream leaves to top of the regeneration column and is routed, with the CO_2 stream from pre-treatment, to compression and purification. Regenerated CO_2 lean solvent is recirculated from the bottom of the regeneration column back to the absorption column via a heat exchanger, where it is used to pre-heat the CO_2 rich solvent prior to regeneration.



5.3.4 CO₂ Compression and Purification

The CO_2 streams from pre-treatment (AGRU) and from the post-combustion capture plant are compressed to 110 bara by a multiple stage compressor. Inter-stage coolers and knock-out drums remove condensed water from the CO_2 stream. The CO_2 stream is further dehydrated in a solid bed adsorption process (molecular sieve) and any traces of oxygen removed in a solid bed reactor to meet CO_2 pipeline specifications for transport to storage.

5.4 Process Performance

Process Flow Diagrams, Heat and Mass Balance and Equipment List for the base case are provided in Appendices 3, 4 and 5 respectively.

5.4.1 CO₂ Captured, Avoided and Emitted

The CO₂ emissions for the LNG plant without CCS are compared against the CO₂ emissions associated with the base case in Figure 2. The CO₂ emissions for the LNG plant are calculated as the sum of CO₂ vented in the AGRU plus the CO₂ produced by combustion of fuel gas in gas turbines used for mechanical drive and power generation.

The basis for plant availability is 95% for LNG plant and 85% for CO₂ capture plant (including compression and purification).

- **CO₂ Emissions (without CCS):** total annual CO₂ emissions are estimated as 1.35 mtpa, calculated as the sum of:
 - 0.27 mtpa CO₂ emitted from AGRU
 - 1.09 mtpa CO₂ emitted from gas turbine exhausts (all gas turbines of LNG process in mechanical drive and power generation service)
- **CO₂ Captured:** 90% of CO₂ in flue gases from combustion process is captured in the post-combustion plant, the total CO₂ captured is 1.06 mtpa, calculated as the sum of:
 - 0.23 mtpa CO₂ captured from AGRU
 - o 0.83 mtpa CO₂ captured from gas turbine exhausts
- CO₂ Emissions (with CCS): total annual CO₂ emissions are estimated as 0.45 mtpa, calculated as the sum of:
 - o 0.30 mtpa CO₂ emitted from gas turbine exhausts (and not captured)
 - 0.15 mtpa CO₂ emissions from power generation associated to capture plant (absorption process, CO₂ compression and purification)
- **CO**₂ **Avoided:** Avoided CO₂ emissions is 0.9 mtpa, calculated as the difference between the total CO₂ emissions of LNG plant without and with CO₂ capture:
 - o 1.35 mtpa CO₂ emissions of LNG plant without CO₂ capture
 - \circ 0.45 mtpa CO₂ emissions of LNG plant with CO₂ capture



Figure 2. CO₂ Emissions With and Without CCS



Table 1 shows the flows of CO₂ sent to storage.

Table 1. Captured CO₂ (4.6 mtpa LNG plant)

	CO ₂ Captured				
Source	t/h	t/day	mtpa (Note 3)	proportion	
AGRU (Note 1)	32.0	768	0.23	21.4%	
Post-combustion Capture (Note 2)	117.4	2817	0.83	78.6%	
Total	149.4	3584.9	1.06	100.0%	

Notes:

- 1. Practically all the CO2 in feed gas is removed in the AGRU
- 2. CO₂ flows consider capture of 90% of CO₂ in flue gases
- 3. Annual flows calculated based on 95% LNG plant availability and 85% $\rm CO_2$ capture plant availability

5.4.1 Power Demand

The estimated power demand for the base case is 34.8 MW, based on estimated electrical loads:

- CO₂ post-combustion capture 16.5 MW*
- CO₂ compression and purification 18.3 MW

*Includes 11 MW for flue gas booster fan

5.4.2 Heating Demand

Estimated flowrate of LP steam for solvent regeneration and reclamation is 160 t/h which equates to a regeneration reboiler duty of about 350 GJ/h.

The base case assumes that the solvent regeneration duty can be met by waste heat recovery (available in excess in an LNG plant), therefore production of steam has practically no associated cost.



6.0 ECONOMIC EVALUATION

6.1 Capital Cost (CAPEX)

A capital cost estimate has been produced for the base case and sensitivity cases using reference project data for representative CCS schemes of similar scope and size. This includes available information from technology licensors and published data.

The cost estimate is performed to an accuracy of $\pm 35\%$. Inputs and assumptions used in the economic evaluation are provided in Appendix 1.

A breakdown of the CAPEX for the capture, compression and purification facilities is shown in Table 2 and Table 3.

6.1.1 Total Plant Cost

The Total Plant Cost (TPC) is the installed cost of the plant including:

- Engineering costs: Front End Engineering and Design, Detailed Engineering, Procurement, Project Management
- Materials: Equipment cost and bulk materials
- Construction costs
- Contractor's fee
- Project contingency

6.1.2 Total Capital Requirement

The Total Capital Requirement (TCR) represents the full project cost including TPC plus the cost of capital spares and working capital, start-up costs, owner's costs, and financial costs during construction.

- **Capital Spares.** The cost associated with the provision of spare parts deemed necessary in order to meet plant availability specification. It is assumed that spare parts have no value at the end of the plant life due to obsolescence
- Working Capital. Cost associated with inventories of chemicals (materials held in storage). For the capture plant, this consists of initial inventory of solvent, chemicals and adsorbent media.
- Start-up costs. Consist of costs associated with the following:
 - Required plant modifications to bring the unit up to full capacity
 - Fuel requirements to cover the operation of the plant at reduced efficiency during start-up operations.
 - \circ $\,$ Operating and maintenance labour costs, including training.
 - Cost of required chemicals and waste disposal during start-up
- **Owners Costs.** Covering the costs of feasibility studies, surveys, land purchase, construction or improvement to roads and railways, water supply etc. beyond the site boundary, owner's engineering staff costs, permitting and legal fees, arranging financing and other miscellaneous costs.
- Interest during Construction. Expenditure is assumed to take place at the end of each year, and interest during construction payable in a year is calculated based on money owed at the end of the previous year.



6.1.3 State of Development

The cost estimates are based on current knowledge of the technology. Postcombustion capture technology based on chemical absorption has been extensively tested, considered in a number of FEED studies and commercially available. Despite this, full scale post-combustion capture plants (with capture capacity in the order of 1 mtpa of CO₂) built to date can still be considered first-of-a-kind plants (FOAK). Although the expected cost reduction for nth-of-a-kind (NOAK) capture plants has been suggested and estimated, it is considered that further experience (i.e. further implementation) is necessary to benefit from technology maturity. Therefore, current cost estimates are to some extent closer to a FOAK level. The capital cost generated in this study is derived from available cost data, and it is recognised that potential costs reductions are likely once the maturity of the technology and experience in its implementation increase. A sensitivity analysis is included to evaluate the impact based on perceived potential for cost reduction for NOAK capture plants.

	Project Scope / System					
Cost	CO₂ Capture Plant	Power Generation	CO₂ Compr. and Purification	CO ₂ Onshore Pipeline	Total	
FEED Manhours	€4.4	€0.5	€1.0	€0.1	€6.0	
EPC Manhours	€28.7	€3.6	€6.9	€0.9	€40.1	
Technology License	€0.2	-	-	-	€0.2	
Major Equipment	€118.1	€14.9	€24.2	€0.4	€157.6	
Bulk Materials	€97.4	€7.5	€22.7	€0.2	€127.8	
Construction	€98.3	€9.9	€18.3	€0.5	€127.0	
Site services	€45.5	€4.6	€8.5	€0.2	€58.8	
Total	€392.6	€41.0	€81.6	€2.3	€517.5	

Table 2. Plant Materials and Labour Cost Breakdown for Base case (million EUR)

Notes:

- 1. Scope of CO₂ capture plant includes tie-ins, flue gas ducting, direct contact cooler, absorber/stripper system, water treatment and all utility connections
- 2. Scope of Power Generation includes gas turbine generator and all electrical equipment (to support CO₂ capture plant and compression operations)
- 3. CO₂ compression and purification includes compressor system (vessels, coolers, machinery), dehydration, contaminants removal (oxygen) and all utility connections
- 4. CO₂ onshore pipeline scope includes CO₂ pipeline upstream of connection to offshore CO₂ pipeline, including pig trap



Table 3. CAPEX Breakdown for Base	case (million EUR)
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А	HOME OFFICE ENGINEERING & PROCUREMENT SERVICES		
A.1	FEED Manhours	€	6.0
A.2	EPC Manhours	€	40.1
A.3	Allowance for Licensor's Process Design Package	€	0.2
тот	TOTAL - HOME OFFICE SERVICES		46.3

В	MATERIALS	
B.1	Major Equipment	€ 157.6
B.2	Bulk Materials	€ 127.8
ТОТ/	AL - MATERIALS	€ 285.4

С	CONSTRUCTION	
C.1	Construction	€ 127.0
C.2	Site services	€ 58.8
тот	AL - CONSTRUCTION	€ 185.8

TOTAL PLANT MATERIALS AND LABOUR (See Table 2)	€ 517.5

D	OTHER PROJECT COSTS		
D.1	Contingency	€	51.8
D.2	Contractors Fee	€	28.2
тоти	AL - OTHER COSTS	€	80.0

TOTAL PLANT COST (TPC)€ 597.5

Е	OTHER CAPITAL COSTS			
E.1	Interest during construction	€	43.9	
E.2	Capital spares	€	3.0	
E.3	Working capital - solvent	€	3.5	
E.4	Start-up costs - modifications	€	11.9	
E.5	Start-up costs - fuel gas	€	0.4	
E.6	Start-up costs - operating and maintenance labour	€	1.8	
E.7	Start-up costs - maintenance materials	€	0.7	
E.8	Start-up costs - waste disposal	€	0.1	
E.9	Owner's costs	€	41.8	
тот	TOTAL CAPITAL REQUIREMENT (TCR) € 704.6			



6.2 Operating Cost (OPEX)

A breakdown of operating costs for the capture, compression and purification facilities is shown in Table 4. Inputs and assumptions used in the calculation of the operating costs can be found in Appendix 2.

6.2.1 Fixed Operating Costs

Fixed operating costs do not change with an increase or decrease in productivity and must be paid regardless of the plant's performance.

Fixed operating costs are comprised of the following key components:

- Operating and maintenance costs including: cost of materials required for maintenance, the associated maintenance labour cost, the cost of operating labour including supervision, administrative labour and support labour.
- Cost of insurance
- Local taxes and fees.

6.2.2 Variable Operating Costs

Variable operating costs are a function of productivity and are comprised of the following key components:

- Power generation (fuel gas)
- Steam generation
- Solvent cost
- Chemicals and waste disposal cost
- CO₂ transport and storage cost
- CO₂ emission costs.

Insurance	€ 34.3
Taxes & Fees	€ 34.3
Operation and Maintenance	€ 196.0
Power	€ 144.9
Steam (Note 3)	€ 0.0
Solvent	€ 10.9
Chemicals and Waste Disposal	€ 7.3
CO ₂ Transport and Storage	€ 139.3
CO ₂ Emissions	€ 0.0
TOTAL OPERATING COSTS (Note 1, 2)	€ 567.0

Table 4. Breakdown of OPEX for Base Case (million EUR)

Notes:

- 1. Operating costs shown are lifetime costs (25 years).
- 2. Operating costs presented have been discounted to year 1 of operation, based on the discount rate specified in Appendix 1.
- 3. It is assumed that all steam is produced by waste heat recovery (i.e. no additional fuel gas required to produce steam). See sensitivity case in Section 7.7.



6.3 CCS Specific Costs

6.3.1 Cost of CO₂ Captured and Avoided

Total lifetime costs of the CO_2 capture plant (including compression and purification) are:

Total	€1,271.6 million
OPEX	€ 567.0 million
CAPEX	€ 704.6 million

The total specific costs are calculated as follows:

 $Total Specific Cost of CO_2 Captured = \frac{Total Capture Plant Lifetime Costs}{CO_2 captured over Capture Plant Lifetime}$

 $Total Specific Cost of CO_2 Captured = \frac{\pounds 1,271.6 \ million}{1.06 \ \frac{million \ tCO_2}{y \ ear} x \ 25 \ y \ ears} = \pounds 47.3/tCO_2$

 $Total Specific Cost of CO_2 Avoided = \frac{Total Capture Plant Lifetime Costs}{CO_2 avoided over Capture Plant Lifetime}$

 $Total Specific Cost of CO_2 Avoided = \frac{\notin 1,271.6 \text{ million}}{0.9 \frac{\text{million } tCO_2}{\text{year}} x \text{ 25 years}} = \# 55.2/tCO_2$

The specific cost breakdown is shown in Table 5.

Table 5. S	pecific Cost	of CO ₂ Captured	and Avoided fo	r Base case	(€/t Q	CO2)
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Item	Cost of CO ₂ Captured	Cost of CO ₂ Avoided
TOTAL	€ 47.3	€ 55.2
CAPEX (Note 1)	€ 26.7	€ 31.2
OPEX	€ 20.6	€ 24.0
Insurance	€ 1.3	€ 1.5
Taxes & Fees	€ 1.3	€ 1.5
Operation and Maintenance	€ 7.4	€ 8.7
Power	€ 5.5	€ 6.4
Steam	-	-
Solvent	€ 0.4	€ 0.4
Chemicals and Waste Disposal	€ 0.2	€ 0.3
CO ₂ Transport and Storage	€ 4.5	€ 5.2
CO ₂ Emissions Cost	-	-

Notes:

1. CAPEX is based on the TCR.



The cost of CO₂ capture for the base case is \in 47.3 per tonne of CO₂ captured. Over half of this cost (56%) is associated with the capital expenditure and the remainder associated with the operating costs of which the main contributors are operation and maintenance (\in 7.4/t CO₂), power generation (\in 5.5/t CO₂) and CO₂ transport and storage (\in 4.5/t CO₂).

The cost of CO₂ avoided is €55.2 per tonne of CO₂ avoided.

6.3.2 Levelised Cost of LNG

Levelised cost of LNG is defined as the average LNG price that would be necessary over the life of the plant to give a zero net present value. Levelised LNG costs are given in Table 6. These costs only include costs associated to liquefaction. Other costs (including transport costs, which account for a significant proportion of the delivered LNG price) are excluded. Cost penalties associated to emissions (e.g. CO₂ tax) are also excluded in this table. The estimated costs indicate that inclusion of the CCS scheme costs would increase the LNG levelised costs by about 20%.

Cost	Lifetime Costs			
Item	Only Liquefaction	Including CO ₂ capture	Notes	
CAPEX	€4,751m	€5,455m	Based on nominal CAPEX €1033/tpa (USD \$1200/tpa) of LNG capacity, representative of recent liquefaction plants (first train developments). This includes the cost of reception facilities, pre-treatment, liquefaction, utilities, LNG storage and loading Facilities. Subsequent additional liquefaction trains are likely to be less expensive due to the ability to utilise the infrastructure already in place at the LNG plant such as: utilities, storage and loading facilities, etc.	
OPEX	€1,521m	€2.065m	Based on annual OPEX equivalent to approximately 3% of CAPEX (typical) over 25 years, discounted to year 1 of operation	
Total	€6,272m €54.5/tLNG (€1.18/MMBtu)	€7,520m €65.4/tLNG (€1.41/MMBtu)	Only liquefaction CAPEX and OPEX (excluding storage, transport, sales and commercial costs, emissions tax, etc.)	

Table 6	Levelised	LNG Costs
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A comparison of the levelised cost of LNG with and without CCS for CO₂ emission cost (i.e. CO₂ tax) within the range of $\in 0$ to $\notin 200/t$ CO₂ is shown in Figure 3. Emission costs have been discounted to year 1 of operation, based on the discount rate specified in Appendix 1. This analysis provides a method of gauging how economically attractive implementing CCS on an LNG facility is.

The results of the levelised cost analysis indicate that implementation of a CCS scheme on an LNG liquefaction plant would become financially attractive for CO_2 emission costs (i.e. CO_2 tax) above $\in 129/t$ CO₂.



Figure 3. Levelised LNG Cost against CO₂ Emission Cost

Notes:

 Levelised LNG cost shown is includes the cost of LNG production (see Table 6) plus a nominal allowance of €5.2/MMBtu (USD \$6/MMBtu) added to cover other cost components of the LNG supply chain (exploration and production, shipping/transport and regasification). Typically, the price of LNG falls within the range of about €5.2 to €7.7/MMBtu (USD \$6 to \$9/MMBtu) which accounts for factors such as LNG transportation, gasification, market conditions and profit etc.



7.0 SENSITIVITY ANALYSES

7.1 Introduction

In order to assess the impact a change in any of the key inputs and assumptions will have on the cost of CO_2 capture, a number of sensitivity analyses have been performed. The sensitivity analyses will use the same assumptions and design basis used in the base case estimate and will differ only in the parameter which is being investigated.

The parameters that are investigated in the sensitivity analyses are:

- Increased CO₂ content in feed gas
- Fuel Gas Price
- Discount Rate
- CO₂ Transport and Storage Cost
- Plant Life
- Process Heating Costs
- LNG Plant Size
- Capture of CCS Associated Emissions
- nth of a Kind Plant Costs
- Capture of AGRU CO₂ Emissions Only

The results of the sensitivity analyses are detailed in the sections below. A summary of cost of CO₂ capture for all sensitivity cases is shown in Section 8.0.

7.2 Increased CO₂ Content in Feed Gas

The base case considers feed gas with a 2 mol% CO_2 content. In consideration of the increased emissions, particularly from CO_2 vented from the AGRU, a sensitivity is considered for 14 mol% CO_2 in feed gas, representative of the LNG plant currently operating with high CO_2 in feed gas.

Error! Reference source not found. shows the breakdown of CO_2 emissions for this s ensitivity case. This shows an increase of more than double the amount of CO_2 captured and avoided with respect to the base case.







Capital and operating costs have been adjusted to account for the increased CO₂ flow, impacting particularly on the compression and purification duty.

Total lifetime costs of the CO₂ capture plant (including compression and purification) are:

Total	€1	,546.3 million
OPEX	€	732.7 million
CAPEX	€	813.6 million

Although the total costs are greater than costs of the base case, the significant increase in the captured and avoided CO_2 emissions (see **Error! Reference source not found.**) results in I ower specific costs with respect to the base case. This reduction in specific costs is mainly due to the relatively ease to capture CO_2 emissions from AGRU due to the reduced processing (only compression and purification is required). The specific cost of CO_2 capture significantly reduces by about 46% to \in 25.6/t CO_2 , and the avoided cost reduces to \notin 27.3/t CO_2 , see **Error! R eference source not found.**

Table 7. Specific Cost of CO₂ Captured and Avoided (€/t CO₂) – Increased CO₂ Content in Feed Gas

Sensitivity Case	14 mol% CO₂ in feed gas		
Cost	Captured	Avoided	
Total	€ 25.6	€ 27.3	
CAPEX	€ 13.5	€ 14.4	
OPEX	€ 12.1	€ 13.0	

For this sensitivity case, the minimum CO_2 emission cost (CO_2 tax) at which CCS becomes economically attractive is significantly reduced to about $\in 64/t CO_2$.

7.3 Fuel Gas Price

To calculate the costs of fuel gas used in gas turbines, a fixed price of natural gas of $\in 6/GJ$ (LHV) has been used for the base case. This is representative of expected current and future energy prices.

Sensitivities are performed on the natural gas price to evaluate the impact on cost of CO_2 capture and provide comparison to the base case. The following natural case price sensitivity cases are investigated:

- Low Fuel Gas Price: €3/GJ (LHV)
- High Fuel Gas Price: €12/GJ (LHV)

The CO₂ emissions for the LNG plant with and without CCS remain as per the base case, as shown in Figure 2. Total lifetime costs of the CO₂ capture plant for fuel gas price \in 3/GJ are:

 CAPEX € 704.4 million

 OPEX € 471.1 million

 Total €1,175.5 million



Total lifetime costs of the CO₂ capture plant for fuel gas price €12/GJ are:

Total	€1	,393.4 million
OPEX	€	688.4 million
CAPEX	€	705.0 million

For a high fuel gas price of $\leq 12/GJ$ (i.e. double the price used in the base case), the cost rises by 12% to $\leq 52.8/t$ CO₂ captured, $\leq 61.7/t$ CO₂ avoided. For a low fuel gas price of $\leq 3/GJ$ (i.e. half the price used in the base case), the cost reduces by 6% to $\leq 44.5/t$ CO₂ captured, $\leq 52/t$ CO₂ avoided.

Sensitivity Case	Natural Gas Price €3/GJ		ity Case Natural Gas Price €3/GJ Natural Gas Price €12		Price €12/GJ
Cost	Captured	Avoided	Captured	Avoided	
Total	€ 44.5	€ 52.0	€ 52.8	€ 61.7	
CAPEX	€ 26.7	€ 31.2	€ 26.7	€ 31.1	
OPEX	€ 17.8	€ 20.8	€ 26.1	€ 30.5	

Table 8. Specific Cost of CO₂ Captured and Avoided (€/t CO₂) – Sensitivity to Natural Gas Price

Based upon data provided by the US Energy Information Administration (Ref 1) natural gas price is forecasted to be within the range of about $\leq 4.3/GJ$ to $\leq 5.3/GJ$. Total lifetime costs of the CO₂ capture plant for variable forecast gas prices are:

Total	€1	.218.1 million
OPEX	€	513.7 million
CAPEX	€	704.5 million

The specific costs are \leq 46.2/t CO₂ captured, \leq 44.1/t CO₂ avoided, a difference of only 2% with respect to the base case estimates, which confirms the gas prices used for the base case being representative.

The main contributor to the differential in costs of CO₂ capture is the cost of power generation which is directly proportional to the change in natural gas price.

For this sensitivity case, the minimum CO_2 emission cost (CO_2 tax) at which CCS becomes economically attractive is about $\in 122/t$ CO₂ for a gas price of $\in 3/GJ$, $\in 145/t$ CO₂ for a gas price of $\in 12/GJ$ and $\in 126/t$ CO₂ considering variable forecast gas prices.

7.4 Discount Rate

Sensitivities are performed on the discount rate to evaluate the impact on cost of CO₂ capture and provide comparison to the base case. The following discount rate sensitivity cases are investigated:

- Low Discount Rate: 5%
- High Discount Rate: 10%

It is noted that interest rate has been assumed to be equal to the discount rate, so the values above also apply for interest rate.

The CO₂ emissions for the LNG plant with and without CCS remain as per the base case, as shown in Figure 2.



Capital costs are affected by changes in the assumed values by changing the payable interest during construction. Operating costs are affected by changes in discount rate to a far greater extent when expenditure over the lifetime is converted to values at year 1 of operation. Total lifetime costs of the CO₂ capture plant calculated with a discount/interest rate of 5% are:

Total	€1	,518.2 million
OPEX	€	830.3 million
CAPEX	€	687.9 million

Total lifetime costs of the CO₂ capture plant calculated with a discount/interest rate of 10% are:

Total	€1	,136.8 million
OPEX	€	420.9 million
CAPEX	€	715.9 million

For a high discount/interest rate (10%) the cost of CO_2 capture reduces by about 9% to \in 43/t CO_2 captured, \in 50.3/t CO_2 avoided. For a low discount/interest rate (5%), the cost of CO_2 capture increases by about 22% to \notin 57.5/t CO_2 captured, \notin 67.2/t CO_2 avoided.

Table 9. Specific Cost of CO ₂ Captured and Avoided (€/t CO ₂) – Sensitivity to Discount/Interest
Rate

Sensitivity Case	Discount Rate 5%		Discount	Rate 10%
Cost	Captured	Avoided	Captured	Avoided
Total	€ 57.5	€ 67.2	€ 43.0	€ 50.3
CAPEX	€ 26.0	€ 30.4	€ 27.1	€ 31.7
OPEX	€ 31.4	€ 36.7	€ 15.9	€ 18.6

For this sensitivity case, the minimum CO_2 emission cost (CO_2 tax) at which CCS becomes economically attractive is about $\in 122/t$ CO₂ for a 5% discount rate and $\in 139/t$ CO₂ for a 10% discount rate.

7.5 CO₂ Transport and Storage Cost

The costs associated with the transport and storage of the captured CO_2 are highly site specific and dependent upon various factors including; quantity of CO_2 to be transported and stored, relative location of CCS plant to storage site, the pipeline pathway and the type of storage etc.

Sensitivities are performed on the CO_2 transport and storage cost to evaluate the impact on cost of CO_2 capture and provide comparison to the base case.

A wide range of transport and typical storage costs have been quoted, ranging from \in 5/tCO₂ (Ref 9), \in 11/tCO₂ (\$10/tCO₂, Ref 11) to \in 90-115/tCO₂ (£80-100/tCO₂, Ref 12). The base case assumes a notional \in 10/tCO₂, and the following CO₂ transport and storage cost sensitivity cases have been investigated:

- Nil CO₂ Transport and Storage Cost: €0/t CO₂
- Moderate CO₂ Transport and Storage Cost: €20/t CO₂

The $\in 0/t$ CO₂ cost case represents enhanced oil or gas production with modest revenues that offset the costs of CO₂ transport and storage.

The CO₂ emissions for the LNG plant with and without CCS remain as per the base case, as shown in Figure 2.



Total lifetime costs of the CO₂ capture plant excluding transport and storage costs are:

 CAPEX € 704.6 million

 OPEX € 425.1 million

 Total €1,129.7 million

Total lifetime costs of the CO₂ capture plant assuming transport and storage costs €20/t CO₂ are:

Total	€1	,135.4 million
OPEX	€	420.9 million
CAPEX	€	703.6 million

For a moderate CO₂ transport and storage cost of $\leq 20/t$ CO₂ (i.e. double the price used in the base case), the cost rises by about 9% to $\leq 51.7/t$ CO₂ captured, $\leq 60.5/t$ CO₂ avoided. When CO₂ transport and storage costs are excluded from the estimate, the cost reduces by about 9% to $\leq 42.8/t$ CO₂ captured, $\leq 50/t$ CO₂ avoided.

Table 10. Specific Cost of CO₂ Captured and Avoided (€/t CO₂) – Sensitivity to Transport & Storage Cost

Sensitivity Case	Transport & Storage Cost €0/t CO₂		Transport & Storage Cost €20/t CO	
Cost	Captured	Avoided	Captured	Avoided
Total	€ 42.8	€ 50.0	€ 51.7	€ 60.5
CAPEX	€ 26.7	€ 31.2	€ 26.7	€ 31.2
OPEX	€ 16.1	€ 18.8	€ 25.1	€ 29.3

Based on the assumptions and methodology used in this report the only cost component affected by this sensitivity analysis is the cost of CO_2 transport and storage, all other costs remain equal to the base case.

For this sensitivity case, the minimum CO_2 emission cost (CO_2 tax) at which CCS becomes economically attractive is about $\in 117/t$ CO₂ if CO₂ transport and storage costs are excluded and $\in 142/t$ CO₂ if a CO₂ transport and storage cost of $\notin 20/t$ CO₂ is considered.

7.6 Plant Life

A sensitivity analysis was performed on the life of the plant to evaluate the impact on cost of CO_2 capture and provide comparison to the base case plant life (25 years). An economic plant life of 40 years is considered for this sensitivity case.

The CO₂ emissions for the LNG plant with and without CCS remain as per the base case, as shown in Figure 2.

CAPEX required for the CCS scheme is assumed to be unaffected by the increase in plant life, i.e. any additional CAPEX required to extend the plant life (e.g. during project execution or upgrades/modifications at a later stage) are excluded.

As per CAPEX, any additional OPEX associated to the increased maintenance costs due to equipment with a degraded performance or in general costs associated to the extension of the plant life are excluded. Total lifetime operating costs are increased with respect to the base case by about 20%.

Total lifetime costs of the CO₂ capture plant for 40 years life are:

 CAPEX € 704.6 million

 OPEX € 651.1 million

 Total €1,355.7 million

The specific costs reduce by about 32% to €32.1/t CO₂ captured, €30.7/t CO₂ avoided.

Sensitivity Case	Plant Life 40 Years			
Cost	Captured Avoided			
Total	€ 32.1	€ 37.5		
CAPEX	€ 16.7	€ 19.5		
OPEX	€ 15.4	€ 18.0		

Table 11. Specific Cost of CO₂ Captured and Avoided (€/t CO₂) – Sensitivity to Plant Life

Both the specific capture costs of the CCS scheme and the specific production costs of the LNG facility are reduced by increasing the plant life (for a fixed CAPEX). Consequently, the minimum CO_2 emission cost (CO_2 tax) at which CCS becomes economically attractive is very similar to the base case at $\in 126/t CO_2$.

7.7 Process Heating Costs

It is assumed in the base case that there will be no running costs associated with the production of steam since the excess heating is expected to be available in LNG facilities through waste heat recovery. This benefit is particular of LNG plants where combined power generation cycles are not implemented. This is likely to be the case, and even if, for a retrofit project, the installed capacity of the waste heat recovery units (WHRU) is not sufficient, additional waste heat recovery capacity can be installed at a relatively small fraction of the overall project expenditure.

A sensitivity analysis is performed to consider costs of steam generation to evaluate the impact on cost of CO_2 capture and evaluate the potential contribution of steam costs (considering steam is produced in a standalone boiler system, either part of the liquefaction plant or provided as offsite utility). This is relevant for costs benchmarking against CCS schemes in other industries.

Steam costs are derived from fuel gas costs and a boiler efficiency of 80%. The scenario assumes that emissions associated to steam generation are not captured, in the consideration that this simplifies the flue gas gathering system, being limited to treating flue gas from gas turbines. Also, this is applicable for the case in which steam is imported from off-site utility generation plants.

For this case, additional emissions associated to the operation of the CO_2 Capture plant are 0.15 mtpa for power generation and 0.17 mtpa for steam generation. The case assumes that these emissions of these are captured. See **Error! Reference source not found.** for b reakdown of CO_2 emissions for this sensitivity case.



Figure 5. CO₂ Emissions for Steam Production Sensitivity Case

Total lifetime costs of the CO₂ capture plant for this sensitivity case are:

CAPEX € 705.1 million <u>OPEX</u> € 710.5 million Total €1,415.6 million

In comparison to the base case, the specific cost of CO_2 capture rises by about 13% to \in 53.6/t CO_2 . The specific cost of CO_2 avoided increases in a greater proportion by 41% to \in 77.6/t CO_2 as it is assumed that the additional emissions produced by the steam boilers are emitted.

Sensitivity Case	Steam Cost Included		
Cost	Captured	Avoided	
Total	€ 53.6	€ 77.6	
CAPEX	€ 26.7	€ 38.7	
OPEX	€ 26.9	€ 39.0	

Table 12. Specific Cost of CO₂ Captured and Avoided (€/t CO₂) – Consideration of Steam Cost

For this sensitivity case, the minimum CO_2 emission cost (CO_2 tax) at which CCS becomes economically attractive is significantly high compared to the base case at about $\in 182/t$ CO₂, mainly due to the increase in operating costs associated to the steam production.

7.8 LNG Plant Size

The chosen liquefaction capacity of the base case (4.6 mtpa) is representative of baseload liquefaction plants, with capacity in the range 3-5.5 for liquefaction trains built in the last 15 years.

A sensitivity case is performed on the size of the LNG plant upon which the CCS scheme is installed to evaluate the impact on cost of CO_2 capture and provide comparison to the base case. A plant representative of the largest LNG train in operation to date (~7.8 mtpa LNG) is considered for this sensitivity case.



Total lifetime costs of the CO₂ capture plant for this sensitivity case are:

Total	€1	,833.2 million
OPEX	€	830.3 million
CAPEX	€́	1,002.9 million

Emissions from fuel gas combustion are produced in proportion to the LNG plant size, with a proportional power requirement for capture, compression and purification. Emissions for the 7.8 mtpa LNG plant with and without CCS are shown in Figure 6.





The total specific cost for the 7.8 mtpa LNG plant size is \in 41.6/t CO₂ captured and \in 48.4/t CO₂ avoided, a reduction of about 12% in comparison to the base case. This in specific costs (mainly CAPEX) is due to economies of scale.

Sensitivity Case	7.8 mtpa LNG Plant		
Cost	Captured Avoided		
Total	€ 41.6	€ 48.4	
CAPEX	€ 22.7	€ 26.5	
OPEX	€ 18.8	€ 21.9	

Table 13. Specifi	c Cost of CO ₂ Capture	d and Avoided (€/t	t CO ₂) – 7.8 mtpa	LNG plant

For the 7.8 mtpa LNG plant, the minimum CO_2 emission cost (CO_2 tax) at which CCS becomes economically attractive is about \in 113/t CO_2 .

7.9 Capture of CCS Associated Emissions

The base case assumed that the CO_2 emissions associated with the operation of the CO_2 capture plant, CO_2 compression and purification are released to atmosphere. A sensitivity case was considered to evaluate the impact of processing the flue gas produced by the gas turbines supplying power to operate the CO_2 capture plant, CO_2 compression and purification



processes. This increases the size of the capture plant and power requirement by about 15% (from 34.8 MW to 40 MW), which further increase emissions.

Figure 7 shows the effect of capturing the CCS associated emissions on the CO_2 captured, avoided and emitted for the base case and the sensitivity case. The total CO_2 generated for the sensitivity case is greater than the base case (see Figure 2) due to the additional fuel gas combustion to produce the power required for the operation of the bigger capture, compression and purification plant. Despite this, the net CO_2 being emitted is reduced (from 0.45 mtpa CO_2 to 0.34 mtpa CO_2).





Due to the increased size of the capture, compression and purification plant (including requirements such as additional ducting, etc.) there is an increase of about 10% in both CAPEX and OPEX:

Total	€1	,382.8 million
OPEX	€	609.3 million
CAPEX	(€	773.5 million

The increased costs offset the benefits gained via the ability to further reduce the amount of CO_2 emitted. Consequently, the specific CO_2 capture and avoidance costs are only reduced by about 2%, to $\leq 46.5/t CO_2$ captured, $\leq 54.5/t CO_2$ avoided.

Table 14. Specific Cost of CO₂ Captured and Avoided (€/t CO₂) – Capture of CCS Emissions

Sensitivity Case	CCS Emissions Captured			
Cost	Captured	Avoided		
Total	€ 46.5	€ 54.5		
CAPEX	€ 26.0	€ 30.5		
OPEX	€ 20.5	€ 24.0		



the minimum CO₂ emission cost (CO₂ tax) at which CCS becomes economically attractive is about $\leq 125/t$ CO₂.

While costs savings may be achievable for this option (as a result of economies of scale), the increase complexity in design and operation of plant with capture from flue gas from CCS scheme may limit the attractiveness of this integrated scheme being considered.

7.10 nth of a Kind Plant

Typically, when a FOAK plant begins operation and gains experience in how the plant performs, designs for future plants can be streamlined and cost savings may be achieved. For example, cost savings can be realised by removing equipment items and design margins (equipment oversizing) included to mitigate the risk of breaching performance specifications due to uncertainties in process performance. Reductions in plant material and labour cost can therefore be achieved for subsequent plant designs.

As noted in Section 6.1.3, further experience is deemed necessary for CCS projects to benefit from technology maturity and associated NOAK cost reductions. As the cost estimates have been derived from current knowledge of the technology and relatively limited project execution experience, the base case can be considered to be a FOAK plant.

There appears to be a consensus derived from experience in CCS project execution (e.g. Ref. 2) and research work in CCS cost reduction, suggesting that indicative CAPEX reductions of up to 30% are achievable in future plants.

Total lifetime costs of the CO₂ capture plant for this sensitivity case are:

Total	€	966.8 million
OPEX	€	472.2 million
CAPEX	€	494.6 million

The CO₂ emissions for the LNG plant with and without CCS remain as per the base case, as shown in Figure 2.

Consideration of a 30% reduction on estimated TPC leads to a reduction in total specific CO₂ capture and avoidance costs of \in 36.6/t CO₂ and \in 42.8 /t CO₂ respectively.

Sensitivity Case	NOAK capture plant			
Cost	Captured	Avoided		
Total	€ 36.6	€ 42.8		
CAPEX	€ 18.7	€ 21.9		
OPEX	€ 17.9	€ 20.9		

Table 15. Specific Cost of CO₂ Captured and Avoided (€/t CO₂) – NOAK plant

In line with the 30% reduction in TPC, the minimum CO_2 emission cost (CO_2 tax) at which CCS becomes economically attractive is about $\in 100/t$ CO_2 .

7.11 Capture of AGRU CO₂ Emissions Only

Significant costs are associated to the post-combustion capture of CO_2 from combustion processes (CO_2 removal from flue gas from gas turbine exhausts). The potential for implementation of post-combustion capture to approach decarbonisation of LNG production appears to be limited under the current scenario. As noted in previous sections, a relatively high CO_2 tax in above $\in 100$ per tonne of CO_2 emitted would be necessary to make post-combustion capture to LNG plants.



Increased attractiveness and potential to implement CCS schemes in LNG plants exist for sequestration of CO_2 contained in the reservoir gas and removed from the liquefaction plant feed gas in the acid gas removal pre-treatment process. The CCS scheme is essentially limited to compression and dehydration of the highly pure CO_2 stream produced in the AGRU. This is particularly attractive for plants with a moderate to high CO_2 content (nominally above 5 mol%), resulting in large volumes of CO_2 being otherwise vented and incurring in significant emission penalties. This has been the case of CCS schemes implemented in the Snøhvit and Gorgon developments.

Sensitivity cases have been performed to evaluate only the cost associated to sequestration of CO_2 removed from the feed gas, considering a CO_2 content of 2 mol%, 6 mol% and 14 mol%. These cases exclude the requirement for a CO_2 capture plant and its associated CAPEX and OPEX cost, as the separation of CO_2 is performed by the AGRU. The process consists of compression and purification. The compression duty is shown in Table 16.

Feed Gas CO ₂ Content Case	Power Requirement	Notes		
2 mol%	3.9 MW	Assumed that power requirement can be supplied by the		
6 mol%	11.8 MW	spare capacity of the liquefaction power generation plant		
14 mol%	27.4 MW	Additional power generation capacity required (new gas turbine)		

Table 16. Estima	ted power requirement fo	r compression and de	ehydration of AGRU CO ₂ stream
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Essentially all the CO_2 contained in the feed gas is removed in the AGRU, and the entire CO_2 stream produced from the AGRU is compressed and transported for storage (when the CO_2 compression and purification plant is operating, assumed availability 85%). The emissions considered are those associated to the power generation for CO_2 compression and dehydration/purification (burning of fuel gas).

CO₂ emissions with and without CCS are shown in Figure 8.



Figure 8. CO₂ Emissions With and Without CCS – CO₂ from AGRU only



The project execution period is taken to be two years. Total lifetime costs of the CO₂ capture plant for the compression and purification plant for CO2 from AGRU are as shown below.

For 2 mol% CO₂ in feed gas (0.23 mtpa CO₂):

Total	€	93.5 million
OPEX	€	63.1 million
CAPEX	€	30.4 million

For 6 mol% CO₂ in feed gas (0.68 mtpa CO₂):

Total	€	220.8 million
OPEX	€	157.8 million
CAPEX €		63.0 million

For 14 mol% CO₂ in feed gas (1.58 mtpa CO₂):

CAPEX € 144.0 million <u>OPEX</u> € 352.4 million Total € 496.4 million

Specific costs for capture of AGRU CO₂ emissions only are €16.5/t CO₂, €13/t CO₂ and €12.5/tCO₂, whilst specific CO₂ avoided costs are €17.9/t CO₂, €14.1/t CO₂ and €13.6/t for 2 mol%, 6 mol% and 14 mol% feed gas CO₂ content respectively.

Table 17. Specific Cost of CO₂ Captured and Avoided (€/t CO₂) – CO₂ from AGRU only

Sensitivity Case	2 mol% CO ₂		6 mol% CO ₂		14 mol% CO ₂	
Cost	Captured	Avoided	Captured	Avoided	Captured	Avoided
Total	€ 16.5	€ 17.9	€ 13.0	€ 14.1	€ 12.5	€ 13.6
CAPEX	€ 5.4	€ 5.8	€ 3.7	€4.0	€ 3.6	€ 3.9
OPEX	€ 11.2	€ 12.1	€ 9.3	€ 10.1	€ 8.9	€ 9.6

Breakeven analysis shows that capture of AGRU CO₂ emissions only, the emission costs (CO₂ tax) at which CO₂ capture becomes economically attractive are as low as \in 29/t CO₂ for 2 mol% CO₂ and \in 33/t CO₂ for 6 mol% and 14 mol% CO₂. This level of CO₂ tax is within current environmental policies in some regions (e.g. Norway and Finland). This confirms that for high reservoir gas CO₂ content capture of AGRU CO₂ emissions only is likely to be more attractive than capture of both AGRU and flue gas CO₂ emissions while the cost of flue gas CO₂ capture remains high.

8.0 SUMMARY

8.1 Base Case

- It is estimated that the total plant, materials and labour cost for the capture, compression and purification plant is €517.5 million, Total Plant Cost (TCP) including contingency and contractor's fee is estimated as €597.5 million and the Total Capital Requirement (TCR) is approximately €705 million.
- Total OPEX requirement (discounted NPV) for the project over the life of the plant is estimated as €567 million.
- The cost of CO₂ captured for the base case is €47.3/t CO₂. The main cost component for the base case is CAPEX which accounts for 56% of the total cost. OPEX accounts for the remaining cost; the key components are operation and maintenance (16%), power generation (12%) and CO₂ transport and storage (9%).
- The cost of CO₂ avoided for the base case is €55.2/t CO₂.
- Breakeven analysis indicates that the minimum cost of CO₂ emissions (CO₂ tax) would have to be at least €129/t CO₂ to justify the installation of CCS (based on CO₂ emission costs against CO₂ capture costs only).

8.2 Sensitivity Analyses

- Due to the relatively ease to capture CO₂ separated in AGRU limited to compression and purification (dehydration mainly), an increase in the CO₂ content in feed gas will allow overall reductions in the specific costs. For 14 mol% CO₂ content in feed gas (representative of LNG feed gas with high CO₂ content), specific costs are significantly reduced (by about half compared to the base case) to €25.6/t CO₂ captured, €27.3/t CO₂ avoided.
- Natural gas price impacts specific cost due to the requirement for fuel gas to produce power which constitutes a large portion of the total cost. Considering a range of natural gas prices between €3/GJ and €12/GJ (LHV) result in capture costs of about €44.5/t CO₂ and €52.8/t CO₂ respectively, avoidance costs €52/t CO₂ and €61.7/t CO₂ respectively. It should be noted that steam production is assumed to be provided by waste heat recovery. While there may be costs associated with adapting the existing facilities for the CCS scheme, the cost of steam production is assumed to be zero.
- At discount and interest rates of 5 and 10% the capture costs are about €57.5/t CO₂ and €43/t CO₂ respectively, avoidance costs €67.2/t CO₂ and €50.3/t CO₂ respectively.
- Adjusting the cost for CO₂ transport and storage affects the total specific project cost, with higher costs for transport and storage resulting in higher total costs. At CO₂ transport and storage costs of €0/t CO₂ and €20/t CO₂ the capture costs range between €42.8/t CO₂ and €51.7/t CO₂ respectively, avoidance costs range between €50/t CO₂ and €60.5/t CO₂ respectively. All cases do not include for the CAPEX requirement for the infrastructure associated to CO₂ transport and storage.
- Increasing plant life to 40 years reduces capture costs by about a third with respect to the base case to €32.1/t CO₂ with avoided cost reduced to €37.5/t CO₂. This assumes that there are no additional significant costs associated with the design, construction or operation of the plant as a result of the extended plant life.



- It is assumed in the base case that there will be no running costs associated with the production of steam since the excess heating is expected to be available in LNG facilities through waste heat recovery. Generation of steam required for the capture, compression and purification plant by a standalone boiler (either part of the liquefaction plant or provided as offsite utility) introduces additional cost and emissions to the project. Capture cost is increased with respect to the base case to about €53.6/t CO₂, avoidance cost is increased to over €77.6/t CO₂.
- Large scale LNG trains (such as those found in Qatar with capacity of 7.6 mtpa LNG) may provide greater benefits for CO₂ capture because of economies of scale. The total capture cost for plants this size is reduced with respect to the base case to €41.6/t CO₂ and avoidance cost reduced to about €48.4/t CO₂.
- Capturing the emissions associated with burning fuel to provide power to the capture, compression and purification processes results in a slight reduction in specific capture costs to €46.5/t CO₂, avoidance cost €54.5/t CO₂. The selection of this option may be dictated by increased complexity in the design and operation of the plant, but the benefits appear small.
- NOAK plant designs increase attractiveness of CCS due to the ability to reduce CAPEX costs compared to FOAK design. An expected CAPEX reduction of up to 30% (Ref 2) is reported to be achievable, resulting in a reduction of capture cost to €36.6/t CO₂, avoidance cost about €42.8/t CO₂.
- CAPEX, OPEX and specific cost of capture are all significantly reduced when the CCS scheme only considers sequestration of the CO₂ that separated from the feed gas in the AGRU. For moderate to high reservoir gas CO₂ content (i.e. higher than typical 2 mol% CO₂), the specific capture cost is estimated as €16.5/t CO₂, avoidance cost €17.9/t CO₂. This shows the potential for implementation of CCS schemes limited to sequestration of CO₂ in the feed gas over post-combustion schemes to approach decarbonisation of LNG production.
- The minimum CO₂ emission cost (i.e. CO₂ tax) required to justify implementation of post-combustion CCS schemes in LNG plants purely on economics, i.e. reduction of costs paid for emitted CO₂, is in the order of €100/t CO₂. Current world emission policies set CO₂ tax at a relatively low value (if any), with the majority of emissions currently priced at less than about €10/t CO₂. This indicates that implementation of post-combustion CCS would only occur for either significant CO₂ tax increases or by drivers other than plant economics.

It is noted that when the CCS scheme only considers sequestration of the CO₂ that separated from the feed gas in the AGRU, the minimum CO₂ tax required is in the order of \leq 30/t CO₂. This level of CO₂ tax is within current environmental policies in some regions (e.g. Norway and Finland), which indicates the potential for the implementation of CCS as this is necessary by the project economics (as in the case of the Snøhvit and Sleipner developments).




Figure 9 - Summary of Specific Capture Cost for Base Case and Sensitivity Analyses









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APPENDIX 1 – BASIS OF DESIGN

This report is in line with the IEAGHG's criteria for technical and economic assessment of liquefied natural gas (LNG) plants. An overview of the technical criteria used is specified in this section.

A1.1. LNG Plant

A1.1.1. Location

The LNG liquefaction plant is located in the Gulf Coast region, USA. The cost implication has been considered in labour productivity and labour cost. Due to currency volatility at the time of writing, variations in exchange rate would make the main difference in costs (depending on reported currency costs, origin of equipment).

A1.1.2. Capacity

Base case:

- Plant Capacity: 4.6 mtpa of LNG
- Liquefaction Technology: Propane pre-cooled mixed refrigerant (C3MR)
- Specific Power: 0.3 kWh/kg of LNG
- Refrigerant Compressor Drive: 2 x Frame 7 (split MR configuration)
- De-rated Mechanical Drive Gas Turbine Efficiency: 32.6%
- Refrigeration Cycle Power: 166 MW
- Power Generation: 41.4 MW (Industrial Gas Turbines in Simple Cycle)
- De-rated Power Generation Gas Turbine Efficiency: 32.8%

Sensitivity case

- Plant Capacity: 7.8 mtpa of LNG
- Liquefaction Technology: AP-X
- Refrigerant Compressor Drive: 3 x Frame 9 (with export of surplus power)
- De-rated Mechanical Drive Gas Turbine Efficiency: 34.1%
- Specific Power: 0.3 kWh/kg of LNG
- Refrigeration Cycle Power: 276 MW (81 MW C3, 109 MW MR, 86 MW N₂)
- Refrigeration Cycle Power: 95 MW (produced by generators in Frame 9 compressor string, balance of cycle compression power)

A1.1.3. Availability

Availability of the LNG liquefaction plant is 95%.

The CO_2 capture plant is designed to avoid impact on availability of the LNG liquefaction plant, with flue gas bypassing the capture plant and diverted to the existing vent stacks if the CO_2 capture plant is not available or to assist with start-up / shut-down operations.

The CO₂ capture plant is designed to have availability of at least 85% and appropriate equipment sparing is included to achieve this.



A1.1.4. Life

The plant is designed for an operating life of 25 years.

A1.1.5. Construction Considerations

It is assumed that there are no restrictions on plant area and no special construction requirements or constraints on delivery of equipment. The site is assumed to be level with minimum site preparation required.

The provision of rail lines, roads, fresh water supply, high voltage electricity transmission lines, high pressure CO_2 and natural gas pipelines up to be the battery limits of the plant is assumed to be available and therefore falls outside the scope of design.

Seismic risk is assumed to be negligible.

A1.2. Local Ambient Conditions

The plant performance is calculated based on the annual average ambient conditions.

Gas turbine performance (power and efficiency de-rating and combustion calculations) is estimated at average conditions. For the economic assessment, the plant is assumed to operate at its design performance with no performance degradation over time. Design margins are included to counter the effects of degradation and variation in ambient conditions.

Condition	Minimum	Average	Maximum
Air Temperature (dry bulb)	-15 ⁰C	21ºC	43ºC
Relative humidity	40%	74.7%	95%
Atmospheric pressure	-	1.013 bara	-

Table A1.1 - Annual Average Ambient Conditions

A1.3. Natural Gas - Composition and Conditions

Natural gas fed to the LNG production facility has a temperature approaching ambient and pressure of 70 bar.

An analysis of the natural gas is specified in Table A1.2.

Component	Units	Value
Methane (CH ₄)	mol%	89.0
Ethane (C ₂ H ₆)	mol%	7.0
Propane (C ₃ H ₈)	mol%	1.0
Butane (C ₄ H ₁₀)	mol%	0.1
Pentane (C ₅ H ₁₂)	mol%	0.01
Carbon Dioxide (CO ₂)	mol%	2.0
Nitrogen (N ₂)	mol%	0.89
Lower Heating Value (LHV)	kJ/kg	46,502
Higher Heating Value (HHV)	kJ/kg	51,473



A1.4. Flue Gas - Composition and Conditions

A1.4.1. Flue Gas Flow

The CO_2 capture plant size for the base case, based on flue gas flows, is shown in Table A1.3 below.

Parameter	Basecase
CO ₂ Capture Plant Capacity (Note 1)	2800 t/day of CO ₂
Flue Gas Flow	2,630 t/h

Table	A1.3 -	Flue	Gas	Flow	Rate
Table	A1.3 -	Flue	Gas	Flow	Rat

Note 1. Based on separation of 90% of CO_2 in flue gases

A1.4.2. Flue Gas Flow

The flue gas is fed to the CO₂ capture plant at atmospheric pressure.

A1.4.3. Flue Gas Temperature

Exhaust gas temperature from gas turbines is about 500°C. Waste heat recovery is assumed to be present (supplying heat to users in the natural gas pre-treatment units), reducing the flue gas temperature to 200°C.

The flue gas will be cooled upstream of the CO₂ absorber in the Direct Contact Cooler to achieve the optimal temperature at inlet to the absorber.

A1.4.4. Flue Gas Composition

The flue gas composition is typical of gas turbine exhaust gases. The plant design is based on the values given under the Design Basis column in Table A1.4 below, with consideration of the potential range of compositions noted.

The composition given is that of the hot flue gas upstream of cooling, and the water content of the stream entering the absorber (depending on required temperature and pressure) will be lower.



Component	Units	Design Basis	Range	
Nitrogen (N ₂)	mol%	74.3	70.8 – 75.5	
Argon (Ar)	mol%	0.9	(balance)	
Oxygen (O ₂)	mol%	13.8	13.1 - 14.5	Note 1
Carbon Dioxide (CO ₂)	mol%	3.2	2.5 – 3.3	
Water (H ₂ O)	mol%	7.8	7.1 – 13.2	
Nitric Oxide (NO)	ppmv	-	20 – 220	
Nitrogen Dioxide (NO ₂)	ppmv	-	2 – 20	
Carbon Monoxide (CO)	ppmv	-	5 – 330	
Sulphur Dioxide (SO ₂)	ppmv	-	Trace – 100	Note 2
Sulphur Trioxide (SO ₃)	ppmv	-	Trace – 4	
Unburned Hydrocarbons	ppmv	-	5 – 300	
Particulate Matter Smoke	ppmv	-	Trace - 25	

Table A1.4 - Flue Gas Composition

Notes:

- 1. Potential range of compositions depending on type of gas turbine, fuel gas and ambient conditions for processes and locations under consideration
- 2. Typical gas turbine exhaust emissions from combustion of conventional gas fuels (GE Power Systems, Specification GER-4211)

A1.5. CO₂ From AGRU - Composition and Conditions

The properties of the CO_2 separated in AGRU which is combined with the CO_2 from postcombustion capture and fed to compression and purification are shown in Table A1.5 below.

Table A1.5 - CO₂ From P	re-treatment Properties
-------------------------	-------------------------

Parameter (CO ₂ From AGRU)	Basecase		
Flow Rate	32 t/h		

A1.5.1. CO₂ From Pre-treatment Pressure

The CO_2 from pre-treatment is fed to compression and purification at around atmospheric pressure (1.1 bara)

A1.5.2. CO₂ From Pre-treatment Temperature

The CO_2 from pre-treatment is available at a temperature of 30°C. The temperature is dictated by the process used in the Acid Gas Removal Unit (AGRU).

A1.5.3. CO₂ From Pre-treatment Composition

The composition of the CO₂ stream from pre-treatment is shown in Table A1.6.

Component	Units	Design Basis
Carbon Dioxide (CO ₂)	mol%	95.1
Water (H ₂ O)	mol%	3.9
Hydrocarbons	mol%	1.0

Table A1.6 ·	 Flue Gas 	Composition
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A1.6. Required Unit Performance

Recovery of CO_2 within the CO_2 capture plant is 90%.

CO₂ captured from the both the pre-treatment stage and the CO₂ capture plant is compressed to 110 bar, cooled to a maximum temperature of 30 °C and purified to the specifications shown in Table A1.7.

The specifications depend on technical factors related to CO₂ transport and storage including corrosion, hydrate formation and health and safety, and regulatory requirements.

Component(s)	Maximum Pipeline Specification
Water (H ₂ O)	50 ppmv
Nitrogen (N ₂) / Argon (Ar)	4 mol%
Oxygen (O ₂)	100 ppmv
Carbon Monoxide (CO)	0.2 mol%
Methane (CH ₄) and other hydrocarbons	4 mol%
Hydrogen Sulfide (H ₂ S)	20 ppmv
Oxides of Sulfur (SOx)	100 ppmv
Oxides of Nitrogen (NOx)	100 ppmv
Total non-condensable	4 mol%

Table A1.7	- CO ₂	Maximum	Impurit	v for	Pipelines
				J	

A1.7. Utilities

A1.7.1. Electricity

It is assumed that dedicated power generation equipment will be installed as part of the capture plant scope in order to meet the power requirements of the CO_2 capture, compression and purification processes.

A1.7.2. Steam

Heating requirements are assumed to be fully provided by means of waste heat recovery from gas turbine exhausts. LP steam (\sim 3 bara) is required for regeneration of the solvent used in the CO₂ capture plant.

A1.7.3. Fuel Gas

Fuel gas is taken from the liquefaction plant feed gas downstream of the AGRU and dehydration. It is assumed that the fuel gas required for the CO₂ capture, compression and purification does not have an impact in the product deliverability of the LNG plant (i.e. sufficient fuel gas is available from the reservoir such the production profile over the lifecycle of the plant is not affected, allowing the plant to operate at baseload).

The fuel gas composition is based on typical natural gas feed gas to LNG plants following pre-treatment in the acid gas removal unit.



Component	Units	Feed Gas
Nitrogen	mol%	0.91%
Carbon Dioxide	ppmv	50
Methane	mol%	90.81%
Ethane	mol%	7.14%
Propane	mol%	1.02%
Butane	mol%	0.10%
Pentane	mol%	0.01%
Molecular Weight	kg/kgmol	17.49
Lower Heating Value (LHV)	kJ/kg	48,880
Higher Heating Value (HHV)	kJ/kg	54,156

A1.7.4. Cooling

Process cooling duties will be provided against air in fin-fan coolers. Design ambient air temperature (dry bulb) for air cooling is 25°C.

A1.7.5. Water

Raw water for process purposes is assumed to be available. The water is treated as required for process use in a dedicated water treatment plant. A water treatment system is included as part of the scope of the CO_2 capture plant.

A1.7.6. Nitrogen and Instrument Air

Nitrogen and instrument air for the CO_2 capture plant is assumed to be within the capacity of the system supplying nitrogen and instrument air to the liquefaction plant.

A1.7.7. Chemicals and Consumables

Buffer storage is included within the design to assist with start-up, shut-down or abnormal processing operations.

A1.8. By-products and Waste

Waste in the form of sludge will be produced from thermal treatment (thermal reclamation) of solvent by the process. The waste will be taken offsite for further processing and/or disposal as required.



APPENDIX 2 – INPUTS AND ASSUMPTIONS

The purpose of this section is to define the inputs and assumptions used when generating the economic evaluation. The inputs and assumptions can be found in Table A2.1 and are pertinent for the base case.

Various sensitivity analyses are presented in Section 7.0 to assess the sensitivity of the economic analysis on various key inputs and assumptions.

Project Parame	eters	
Overall Project Duration	4 years	A project duration has been assumed for the purpose of this study which is representative of the time required for the engineering (FEED and Detailed Design), procurement, construction and commissioning of a facility of this magnitude. The duration of construction and commissioning is assumed to be 2 years.
First Gas	2024	Assumed first gas date. The economic analysis assumes that cost inputs are representative of first gas year prices.
Plant Life	25 Years	An operating life which is considered standard for economic evaluations of this nature has been selected. It should be noted that in reality the plant life may be in excess of this value.

Table A2.1 - Inputs and Assumptions used for Cost Estimate

Economic Para	meters	
Currency	€	 The costs presented are in Euros (€). Where necessary, the foreign exchange rates below have been applied for converting US Dollars (\$) and British Pounds (£) into Euros: 1 British Pound = 1.15 Euros 1 US Dollar = 0.86 Euros
Discount Rate	8%	Standard discount rate used in constant money values. Where applicable, a discount rate of 2% is used for any costs incurred after the plant closure date.
Loan	Various	It is assumed that a loan is required to cover the Total Capital Requirement (TCR). It is assumed that the loan repayments are spread evenly across the plant operating life based on an interest rate equal to the discount rate. It is also assumed that the loan will be provided in full from the bank (i.e. no external funding such as government grants are available).



Economic Para	meters	
Interest During Construction	8%	Interest rate during construction is assumed to be equal to the discount rate. When calculating the interest accumulated during construction it is assumed that expenditure takes place at the end of each year and interest during construction payable in a year is calculated based on money owed at the end of the previous year. For example, the money owed at the end of the first year of construction is used to calculate the interest payed in the second year of construction. It is assumed that the total capital requirement (excluding interest accumulated during construction) will be required in the following proportions: 1 st Year: 3% of TCR (excluding interest)
		2 nd Year: 13% of TCR (excluding interest)
		3 rd Year: 48% of TCR (excluding interest)
		4 th Year: 36% of TCR (excluding interest)
CO ₂ transport and storage cost	€10/t CO₂	The cost of CO_2 transport and storage is included in the cost estimate as a variable operating cost. It is assumed that the infrastructure required for the transportation and storage of CO_2 falls outside of the scope of the cost estimate and as a result, no CAPEX costs are incurred.
CO ₂ Emission Cost	€0/t CO2	There are no costs associated with the emission of CO ₂ . It is noted that this assumption is dependent upon the location of the plant and in some regions this assumption may not be representative.
Price Escalation	3% (APR)	A typical / indicative annual percentage rate of escalation is used. Price escalation is applied to all operating costs except from the costs associated with power and steam generation (see 'Electricity Price' and 'Steam unit cost' remarks).
LNG Plant CAPEX (without capture)	€1,200/tpa LNG	Representative value for capital cost of the LNG facility without CO ₂ capture is used for the purpose of this study. It should be noted that this value is likely to be project / plant specific as the capital requirement for an LNG facility is dependent on a wide range of variables including cost of raw materials, cost of labour and foreign exchange rates. It should also be noted that this cost does not include for the capital costs associated with the rest of the LNG supply chain: exploration and production, shipping and regasification.



Fixed Operatin	g Cost Paramet	iers
Annual Maintenance cost	2% of TPC/year	Annual maintenance costs are estimated as a function of Total Plant Cost (TPC) and account for the cost of materials used in maintenance. Maintenance costs are assumed to remain constant throughout the plant life and include cost of maintenance labour (see below).
Annual Maintenance Labour cost	40% Annual Maintenance Cost	Annual maintenance labour costs are estimated as a function of annual maintenance costs.
Operating labour cost	€60,000 / person-year	Typical labour cost representative of the 'fully burdened' cost of labour, including social security payments. Operating labour is included in the cost estimate as a fixed operating cost.
Number of operators	5 / shift	It is assumed that personnel requirements for the capture, compression and purification facilities are; a plant manager (may be shared with existing plant), control room operators (one per shift), maintenance personnel, a laboratory technician, an electrical technician, a mechanical technician and an instrument technician.
Number of operating shifts	5	A 5-shift working pattern is assumed for the operation of the capture, compression and purification plant.
Operating labour supervision	20% Operating Labour Cost	Supervision of operating labour is estimated as a function of operating labour.
Administrative and Support labour	30% of Operating Labour 12% of maintenance cost	Administrative and support labour requirements are estimated as a function of operating labour requirements and maintenance cost. Administrative and support labour is classed as a fixed operating cost.
Insurance cost, local taxes and fees	1% of TPC	Typical cost of insurance, local property taxes and miscellaneous regulatory and overhead fees estimated as a function of total plant cost.
LNG OPEX (without capture)	3% of CAPEX per annum	Representative value for annual operating cost of the LNG facility without CO_2 capture is used for the purpose of this study (Ref 3,4). It should be noted that this value is likely to be project / plant specific as the capital requirement for an LNG facility is dependent on a wide range of variables. It should also be noted that this cost does not include for the operating costs associated with the rest of the LNG supply chain: exploration and production, shipping and regasification.

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Variable Operating Cost Parameters							
Natural Gas Price	€6/GJ (LHV)	Price of natural gas set in line with IEAGHG's standard natural gas price to facilitate comparisons with other studies. The natural gas price is independent of year of operation (fixed for the duration of the plant operating life).					
Electricity Price	€65.85/MWh	Cost of electricity is a function of natural gas price and is based on power supply via Frame 6 gas turbine in simple cycle operation with an efficiency of 32.8%. The study assumes a fixed natural gas price throughout the operating life therefore electricity price is not affected by cost escalation.					
Steam unit cost	€16.39/t	Cost of steam is a function of natural gas price, enthalpy of steam (0.61 MWh/t) and estimated efficiency of steam generation (boiler) (80%). The study assumes a fixed natural gas price throughout the operating life therefore steam unit cost is not affected by cost escalation.					
Solvent unit cost	€5/kg	Representative solvent cost					
Raw process water	€0.2/m³	Representative process water cost					

Start-up Cost P	arameters	
Maintenance and operating and support labour costs	3 months	Salary requirements for labour costs required over and above routine maintenance, operating and support inefficient operation that occurs during the start-up period, including training.
Maintenance materials	1 month	To cover material costs required over and above routine maintenance associated with initial start up.
Chemicals, consumables and waste disposal costs	1 month	Costs associated with the chemicals, consumables and waste disposal incurred during start-up.
Fuel cost	25 % of 1 month fuel requirement @ full load	Percentage of the cost of the full capacity fuel cost for one month, to cover inefficient operation that occurs during the start- up period
Modifications	2 % of TPC	Cost to cover modifications to equipment that may be required to bring the unit up to full capacity.



Working Capital Cost Parameters						
Fuel Stocks	0	Fuel is taken from natural gas source (i.e. the process stream) therefore it is assumed that there are no fuel reserves.				
Chemicals & Consumables	700 m ³ solvent	Costs associated with the chemicals & consumables assumed equal to the solvent and that the costs for other consumables and disposals will be minor in comparison.				

Other Cost Para	ameters	
Engineering Contractors Fees	Various	Engineering contractor fees calculated as the summation of 10% of home office service cost plus 5% on materials cost plus 5% on construction cost. In practice, fees would be subject to the preferred contracting strategy, these factors are suitable for the purpose of this report only.
Project contingency	10% Material & Labour Cost	Typical project contingency estimated as a function of plant materials and labour costs.
Owners Costs and Fees	7% of TPC	Owners costs and fees estimated as a function of the Total Plant Cost (TPC)
Spare parts (Capital spares)	0.5% of TPC	Spare parts (capital spares) are estimated as a function of the Total Plant Cost (TPC).
Decommissioning	0	The salvage value of equipment and materials is assumed to be equal to the costs of dismantling and site restoration, resulting in a zero net cost of decommissioning. To the knowledge of the authors, there are no significant net costs associated with de- commissioning activities.



APPENDIX 3 – PROCESS FLOW DIAGRAMS

C-101-A/E	B E-101	E-201	E-204-	-A/B E-20	5-A/B/C/D/E/F	E-206	K-101	K-201	K-202	P-101-A	↓/B/C/D	P-201-A/B
FLUE GAS	DIRECT CON	ITACT LEAN SOLV	ENT LEAN/RICH	SOLVENT CO	O ₂ STRIPPER	CO ₂ STRIPPER	DIRECT CONTACT	CO ₂ ABSORBER	CO ₂ STRIPPEI	R DIRECT (CONTACT	RICH SOLVENT
BOOSTER FA	NS COOLER W	ATER COOLER	R HEAT EXCH	IANGERS F	REBOILERS	CONDENSER	COOLER			COOLER	(PUMPS	PUMPS
	COOLE	R										
	/_ /_	/-			/_							
	P-202-A/B/C	P-203-A/B	P-206	P-208-A/B	P-209-A/B	S-201	T-201	V-201		W-201	W-501	
	WATER WASH	ABSORBER FEED	ION EXCHANGE	LEAN SOLVENT	CO ₂ STRIPPER	LEAN SOLVE	NT LEAN SOLVEN	T REFLUX SEPA	RATOR ION	EXCHANGE	THERMA	L
	PUMPS	PUMPS	FEED PUMPS	PUMPS	REFLUX PUMPS	5 FILTER	TANK		P	ACKAGE	RECLAIME	R
											PACKAGF	-









APPENDIX 4 – HEAT & MASS BALANCE

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	Client	: IEA GHG R&D Programme	•	
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<u>NOTES</u>

line may contain trace quantities of N_2 , O_2 , H_2O and H_2 .

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	Strea	am No.	1	2	3	4	5	6	7	8							
Notes (See Sheet 1)			FLUE GAS	BOOSTER FAN	FLUE GAS TO	TREATED FILLE	CO ₂ FBOM	CO ₂ FBOM	CO ₂ TO	PROCES	SS						
			FROM GAS	DISCHARGE	ABSORBER	GAS TO VENT	CAPTURE	AGRU	STORAGE	WATER T-201	то						
Tomporaturo	U	nits	200	200	40	21	21	20	20	1-201							
Pressure	h	-C	200	1.01	40	1.01	1 20	1 10	110	5.00			-	-	<u>+</u>		
Total Molar Flowrate	kan	nol/hr	92600	92600	91828	89058	2771	761	3399	133							
Total Mass Flowrate	k	g/hr	2630000	2630088	2616188	2496947	119240	32529	149365	2404						-	
Molecular Weight	kg/ł	kgmol	28.4	28.4	28.49	28.04	43.03	42.73	43.95	18.02					1	-	
Composition	m	ol %	-	-	-	-	-	-	(Note 1)	-							
CO ₂	m	ol %	3.2	3.2	3.2	0.3	96.3	95.1	99.8	0.0							
H ₂ O	m	ol %	7.8	7.8	7.0	7.1	3.8	3.9	0.0	100.0							
02	m	01 %	13.8	13.8	13.9	14.4	0.0	0.0	0.0	0.0					<u> </u>	┥──	
N	m	01 %	74.3	74.3	74.9	0.9 77 3	0.0	0.0	0.0	0.0			-	-			
CH4	m	ol %	0.0	0.0	0.0	0.0	0.0	1.0	0.2	0.0					+		
TOTAL	m	ol %	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0					1		
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APPENDIX 5 – EQUIPMENT LIST

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Nomenc	clature				Abbrevia	ations			
General	Format: A-XYY-Z				ATM	Atmospheric			
А	Equipment Prefix Code				CS	Carbon Steel			
х	Number denoting the proces	ss unit			FV	Full Vacuum			
YY	Serial number of the equipm	ient item			IEA	International Energy A	gency		
Z	Identical equipment (Duty/St	andby, Spare etc.)			KO	Knock Out	0,		
					LNG	Liquefied Natural Gas			
Equipme	ent Prefix Codes				R&D	Reasearch and Devel	opment		
С	Compressors, Blowers, Fans	s, Vacuum Pumps			SS	Stainless Steel			
E	Unfired Heat Transfer Equip	ment: Heat Exchangers, Condens	ers, Coolers, Re	boilers and Electric Heaters					
К	Trayed and Packed Column	s / Towers							
Р	Pumps (Centrifugal, Positive	e Displacement)							
R	Reactor								
S	Gravity and Mechanical Sep	arators - e.g. Filter							
Т	Atmospheric Storage Tanks	and Interceptors							
V	Vessels								
W	Packaged Units								
Process	Unit Codes								
1YY	Flue Gas Conditioning								
2YY	CO ₂ Capture								
3YY	Compression								
4YY	Dehydration & Purification								
5YY	Solvent Reclamation								

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NO.	ITEM TAG	NAME	NO. OF UNITS	CAPACITY	RATED POWER	DESIGN PRESSURE	DESIGN TEMP	DIFF. PRESSURE	DEPTH	SIZE (mm) WIDTH / DIA.	HEIGHT/LE NGTH	RATED DUTY	ТҮРЕ	MATERIAL	REMARKS	•	
					(KVV)	(BANG)	(0)	(DAN)	(m)	(m)	(m)	(KVV)					
1	K-101	Direct Contact Cooler	1	-	-	0.1	-15/230	-	18.0	17.0	25.0	-	-	SS internals,	Internals - Structured Packing Type		
2	P-101- A/B/C/D	Direct Contact Cooler Pumps	4	1600 m ³ /h	350	9.0	-15/230	4.0	-	-	-	-	Centrifugal	CS CS	4 x 33.3% Capacity Capacity is for individual pump Motor Driven		
3	C-101-A/B	Flue Gas Booster Fans	2	-	11000	0.8	-15/230	0.1	-	-	-	-	Centrifugal	CS	2 x 100% Capacity Normal Capacity = 2,630,000 kg/h		
4	E-101	Direct Contact Cooler Water Cooler	1	-	-	9.0	-15/230	-	-	-	-	146000	Fin Fan Air Cooler	CS	-		
CO ₂ A	BSORPTION	i													·		
5	K-201	CO ₂ Absorber	1	-	-	0.1	-15/85	-	15.0	29.0	70.0	-	-	SS	Internals - Structured Packing Type		
6	P-201-A/B/C	Rich Solvent Pumps	3	850 m ³ /h	500.00	18.0	-15/85	13.6	-	-	-	-	Centrifugal	SS	3 x 50% Capacity Motor Driven		
7	P-202-A/B	Water Wash Pumps	2	2800 m ³ /h	400.00	10.0	-15/85	3.4	-	-	-	-	Centrifugal	SS	2 x 100% Capacity Motor Driven		
8	E-202	Water Wash Cooler	1	-	-	10.0	-15/85	-	-	-	-	68250	Fin Fan Air Cooler	SS	-		
9	P-203-A/B	Absorber Feed Pumps	2	1700 m ³ /h	500.00	12.0	-15/85	6.6	-	-	-	-	Centrifugal	SS	2 x 100% Capacity Motor Driven		
10	E-201	Lean Solvent Cooler	1	-	-	12.0	-15/85	-	-	-	-	18500	Fin Fan Air Cooler	SS	-		
SOLV	ENT HANDL	ING	-	-	-	-				r					1		
11	T-201	Lean Solvent Tank	1	1900 m ³	-	0.2	-15/85	-	-	11.0	22.0	-	Vertical Cylindrical	SS	-		
12	T-202	Fresh Solvent Tank	1	800 m ³	-	0.2	-15/85	-	-	10.0	11.0	-	Vertical Cylindrical	SS	-		
13	P-204	Fresh Solvent Transfer Pump	1	30 m³/h	5.50	2.7	-15/85	1.9	-	-	-	-	Centrifugal	SS	Motor Driven		
14	P-205	Solvent Container Pump	1	20 m³/h	2.20	1.6	-15/85	0.9	-	-	-	-	Centrifugal	SS	Motor Driven		
15	W-201	Ion Exchange Package	1	5 m ³ /h	-	-	-	-	5.0	5.0	5.0	-	-	-	-		
16	P-206	Ion Exchange Feed Pump	1/1	70 m³/h	30.00	8.0	-15/85	7.1	-	-	-	-	Centritugal	SS	Motor Driven		
17	E-203	Ion Exchange Feed Cooler	1	-	-	12.0	-15/85	-	-	-	-	200	Fin Fan Air Cooler	SS	-		
18	S-201	Lean Solvent Filter	1	70 m³/h	-	12.0	-15/85	-	-	-	-	-	Cartridge	SS			
19	T-203	Solvent Drain Tank	1	60 m ³	-	1.0	-15/160	-	-	3.0	9.0	-	Horizontal Cyclindrical	SS			
20	P-207	Solvent Drain Pump	1	25 m ³ /h	7.50	7.0	-15/160	4.7	-	-	-	-	Centrifugal	CS	Motor Driven		
21	S-202	Solvent Drain Filter	1	20 m ³ /h	-	7.0	-15/160	-	-	-	-	-	Cartridge	SS	-		

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										SIZE (mm)							
NO.	ITEM TAG	NAME	NO. OF UNITS	CAPACITY	POWER (kW)	PRESSURE (BARG)	TEMP (°C)	DIFF. PRESSURE (BAR)	DEPTH (m)	WIDTH / DIA.	HEIGHT/LE NGTH	DUTY (kW)	TYPE	AREA (m²)	MATERIAL	REMARKS	
SOL	ENT REGEN	EBATION								(m)	(m)						
22	K-202	CO ₂ Stripper	1	-	-	3.5	-15/160	-	7.4	7.5	35.0	-	-	-	SS	Internals - Structured Packing Type	
23	V-201	Reflux Separator	1	80 m ³	-	3.5	-15/105	-	-	3.5	8.5	-	Vertical	-	SS	Internals - Half open pipe inlet, mesh, mist	eliminator
24	P-208-A/B	Lean Solvent Pumps	2	1800 m ³ /h	200.00	6.2	-15/160	3.5	-	-	-	-	Centrifugal	-	SS	2 x 100% Capacity Motor Driven	
25	P-209-A/B	CO ₂ Stripper Reflux Pumps	2	50 m ³ /h	200.00	12.0	-15/160	6.0	-	-	-	-	Centrifugal	-	SS	2 x 100% Capacity Motor Driven	
26	E-204-A/B	Lean/Rich Solvent Heat Exchangers	2	-	-	18.0	-15/140	-	-	-	-	140500	Welded Plate	-	SS	2 x 50% Capacity Design conditions are identical for hot/cold exchanger.	sides of heat
27	E-205- A/B/C/D/E/F	CO ₂ Stripper Reboilers	6	-	-	5.0	-15/150	-	-	-	-	133000	Welded Plate	-	SS	-	
28	E-206	CO ₂ Stripper Condenser	1	-	-	12.0	-15/160	-	-	-	-	39000	Fin Fan Air Cooler	-	SS	-	
SOL	ENT THERM	AL RECLAIMATION	1			1		r	1		1		1	r	1		
29	W-501	Thermal Reclaimer Package	-	5 m ³ /h	-	-	-	-	-	-	-	2600	-	-	SS	Series of semi-batch distillation columns (< inluding reboiler and condenser/reflux pump shown.	1m dia, ~10m height) os. Total rebolier duty
30	T-501	Thermal Reclaimer Feed Tank	1	60 m ³	-	0.3	-15/85	-	-	3.4	6.7	-	Vertical Cylindrical	-	SS	-	
31	P-501	Thermal Reclaimer Feed Pump	1	5 m ³ /h	2.20	8.00	-15/160	3.1	-	-	-	-	Centrifugal	-	SS	Motor Driven	
32	E-501	Thermal Reclaimer Pre-heater	1	-	-	8.0	-15/160	-	-	-	-	350	Welded Plate	-	SS	-	
33	T-502	Reclaimer Waste Tank	1	150 m ³	-	0.2	-15/205	-	-	6.0	5.4	-	Vertical Cylindrical	-	CS Lined	-	
	VHODUCT CC	JMPRESSION		150000	1					1		1				1	
34	W-301	CO ₂ Compressor Package	1	kg/h	15000	-	-	110.0	-	-	-	-	-	-	SS	Multistage Integrally Geared	
35	E-301	1st Stage Suction KO Drum 1st Stage Discharge Cooler	1	- 56 m°	-	10.0	-15/170	-	-	- 3.80	4.90	- 2000	Fin Fan Air	- 620	SS	-	
37	V-302	2nd Stage Suction KO Drum	1	25 m ³		10.0	-15/80	_	_	3 20	4 30	_	Vertical	_	55	Internals - Demister	
38	E-302	2nd Stage Discharge Cooler	1	-	-	10.0	-15/170	-	-	-	-	2900	Fin Fan Air	900	SS	-	
39	V-303	3rd Stage Suction KO Drum	1	22 m ³	-	10.0	-15/80	-	-	2 70	3 80	-	Vertical	-	SS	Internals - Demister	
40	E-303	3rd Stage Discharge Cooler	1	-	-	10.0	-15/170	-	-	-	-	3000	Fin Fan Air Cooler	940	SS	-	
41	V-304	4th Stage Suction KO Drum	1	13 m ³	-	10.0	-15/80	-	-	2.20	3.30	-	Vertical	-	SS	Internals - Demister	
42	V-305	5th Stage Suction KO Drum	1	8 m ³	-	25.0	-15/80	-	-	1.90	2.90	-	Vertical	-	SS	Internals - Demister	
43	E-304	5th Stage Discharge Cooler	1	-	-	25.0	-15/170	-	-	-	-	4300	Fin Fan Air Cooler	1350	SS	-	
44	E-305	6th Stage Discharge Cooler	1	-	-	115.0	-15/170	-	-	-	-	9000	Fin Fan Air Cooler	2960	SS	-	

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NO.	ITEM TAG	NAME	NO. OF UNITS	CAPACITY	POWER (kW)	PRESSURE (BARG)	TEMP (°C)	PRESSURE (BAR)	DEPTH (m)	WIDTH / DIA. (m)	HEIGHT/LE NGTH (m)	DUTY (kW)	ТҮРЕ	AREA (m²)	MATERIAL	REMARKS	
	RODUCT PL	JRIFICATION						1	1			1		1			
45	R-401	Oxygen Removal Unit	1	12 m ³	-	47.0	-15/310	-	-	2.1	3.4	-	Axial	-	SS	Catalyst Bed	
46	E-401	Oxygen Outlet Cooler	1	-	-	47.0	-15/160	-	-	-	-	2900	Cooler	900	SS	-	
47	S-401	Dehydration Filter Coalescer	1	1750 m ³ /h	-	47.0	-15/160	-	-	-	-	-	Coalescer	-	SS	-	
48	K-401-A/B	Dehydration Adsorbers	2	-	-	47.0	-15/310	-	-	2.6	8.6	-	-	-	SS	2 x 100% Capacity. Adsorption/Regeneration Internals - Molecular sieve adsorbent, ceran grid support.	on cyclic duty. nic ball supports,
49	V-401	Regeneration Gas Separator	1	1 m ³	-	47.0	-15/300	-	-	0.7	2.6	-	Vertical	-	SS	Internals - Inlet Device, Mist Eliminator	
50	C-401	Regeneration Gas Compressor	1	-	150	41.9	-15/70	5.00	-	-	-	-	- Electric	-	-	Normal Capacity = 23,733 kg/h	
52	E-402 E-403	Regeneration Gas Cooler	1	-	-	47.0	-15/310	-	-	-	-	2000	Fin Fan Air	-	SS	-	
53	S-402-A/B	Dehydration Outlet Filters	2	1800 m ³ /h	-	47.0	-15/160	-	-	-	-	-	Basket	-	CS		
54	S-403-A/B	Regeneration Gas Filters	2	700 m ³ /h	-	47.0	-15/310	-	-	-	-	-	Basket	-	SS		
MISC	ELLANEOUS	5	г.,					1	1			1		1			
55	V-601	Water Condensate Drum	1	12 m ³	-	5.0	-15/160	-	-	2.0	5.0	-	Horizontal	-	CS	Internais - Inlet Device, Weir Plate	
57	E-601-A/B	Water Condensate Coolers	2	- 230 m ⁻ /n	-	8.0	-15/160	- 2.0	-	-	-	- 21840	Fin Fan Air	-	SS	-	
58	V-602	Process Water Beturn Tank	1	60 m ³	-	10.0	-15/170	_	_	4 00	5.00	-	Horizontal	-	SS	_	
59	P-602-A/B	Process Water Return Pumps	2	1 m ³ /h	1.50	5.0	-	3.1	-	-	-	-	Centrifugal	-	SS	2 x 100% Capacity Motor Driven	
60	V-603	Instrument Air Buffer Vessel	1	90 m ³	-	10.0	-15/85	-	-	3.4	10.1	-	Vertical	-	SS	-	
61	W-601	Instrument Air Compressor Package	1	1550 m ³ /h	-	10.0	-15/85	-	-	-	-	-	-	-	CS	-	
62	W-602-A/B	Instrument Air Drier Package	2	1180 m ³ /h	-	10.0	-15/85	-	-	-	-	-	-	-	CS	-	
63	T-601	Towns Water Tank	1	25 m ³	-	ATM	-15/85	-	-	3.0	3.6	-	Vertical Cylindrical	-	CS Lined	-	
64	P-603-A/B	Towns Water Pumps	2	10 m ³ /h	7.50	6.10	-	6.0	-	-	-	-	Centrifugal	-	SS	2 x 100% Capacity Motor Driven	
65	T-602	Waste Water Tank	1	1580 m ³	-	ATM	-15/85	-	-	12.0	14.0	-	Vertical Cylindrical	-	CS	-	
66	P-604	Waste Water Pump	1	8 m ³ /h	2.20	2.90	-	2.0	-	-	-	-	Centrifugal	-	CS	Motor Driven	
67	W-603	Water Treatment Package	1	60 m ³ /h	-	-	-	-	-	-	-	-	-	-	CS	-	
68	E-602	Vent Vapouriser	1	-	-	FV/27.0	-15/270	-	-	-	-	2300	Snell & Tube	67	CS	-	
69	V-604	Vent KO Drum	1	8 m ³	-	10.0	-60/250	-	-	1.5	4.5	-	Horizontal	-	SS	Internals - Weir Plate	



A comparison of the results of this economic evaluation against costs reported in the literature is provided in this appendix. CCS projects outside of the LNG industry have been included to gain a wider appreciation of how the costs estimated in this report for CCS in LNG compare with the wider CCS industry.

It should be noted that while the information within this section provides a comparison of costs, the scope, assumptions and methodology used for the other projects and/or studies is likely to differ from those used in this report. Therefore, any comparisons are made at the order of magnitude level.

The TPC estimated in this report for the base case is \in 517.5 million excluding contingency and contractor's fee (i.e. total installed cost, material and labour only). This is broadly comparable with and lies within the range of costs of other CCS projects found in the literature (e.g. Ref 7 and 8). The CAPEX for a post combustion CCS plant based on a NGCC power plant with comparable flue gas CO₂ emission rates as this project is in the order of \in 500 million for the onshore capture plant scope (e.g. Ref 7).

The reported cost of CO_2 avoided for a post combustion CCS plant based on a NGCC power plant (Ref 7) is \in 73/t CO₂ for an injection period of 15 years. Scaling the cost of CO₂ avoided using an injection period of 25 years (as in this study) would give a cost of CO₂ avoided of about \in 44/t CO₂ which is similar to the base case avoidance cost estimated in this study.

The cost of post combustion CO₂ capture using a chemical solvent (MEA) on a 600 MW coal fired power plant on a 2005 basis is reported to be about \notin 40/t CO₂ avoided (Ref 6). In contrast, a study performed on a 2007 basis estimated that the cost of CO₂ avoided is about \notin 60.5/tCO₂ (%65/tCO₂) for post combustion capture using a chemical solvent (MEA) and exhaust gas recycle on a NGCC (Ref 10).

Another study reports that post combustion capture using chemical solvent (MEA) results in costs of CO₂ avoided of \in 32/t CO₂ and \in 41/t CO₂ for pulverised coal (PC) and NGCC plants respectively. It should be noted that these figures do not include cost of CO₂ transportation and injection which is reported to add about \in 10/t CO₂ to the avoided CO₂ cost (as per the basis used in this study).

The cost of CO_2 avoided estimated within this report for the base case is about \in 50/t CO₂. Note that this cost is calculated based on the TCR for the project).

In summary, the above review of available literature shows that this estimate is broadly comparable with and lies within the range of costs of other CCS projects found in the literature.

It is reported that for retrofit of CCS capturing only reservoir gas CO_2 from the existing pretreatment facilities on and LNG plant including compression and injection of captured CO_2 into a depleted gas field, the estimated cost of CO_2 capture is about $\in 12/t$ CO₂ (Ref 13). This is in line with the estimates for the sensitivity case performed on sequestration of AGRU CO₂ only.

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IEA Greenhouse Gas R&D Programme

CO₂ Capture for Small Scale and Floating LNG Plants

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1.0 ABBREVIATIONS

AP-C3MR [™]	Air Products Propane Precooled Mixed Refrigerant Liquefaction Process
B&V	Black & Veatch
CAPEX	Capital Expenditure
CCS	CO ₂ Capture and Storage
CO ₂	Carbon Dioxide
DMR	Dual Mixed Refrigerant
FEED	Front End Engineering and Design
FID	Financial Investment Decision
FLNG	Floating Liquefied Natural Gas
FPSO	Floating Production Storage and Offloading
GHG	Greenhouse Gas
IEA	International Energy Agency
kg	Kilogram
KSMR	Kogas Single Mixed Refrigerant
kWh	Kilo Watt Hour
LiMuM	Linde Multistage Mixed refrigerant
LNG	Liquefied Natural Gas
mol	Molar
MR	Mixed Refrigerant
mtpa	Millions of tonnes per annum
MW	Mega Watt
NDX	Nitrogen Dual Expander
OCX	Open Cycle Expander
OSMR	Optimized Single Mixed Refrigerant
PCMR	Pre-Cooling Mixed Refrigerant
PFLNG	Petronas Floating LNG
SCMR	Single Cycle Mixed Refrigerant
SMR	Single Mixed Refrigerant
SSLNG	Small Scale Liquefied Natural Gas
t	Tonne



2.0 EXECUTIVE SUMMARY

IEAGHG commissioned Costain to develop a techno-economic evaluation and cost assessment of different CO_2 capture technologies for LNG plants. This report presents the findings of a review performed on Small Scale LNG (SSLNG) and Floating LNG (FLNG) plants with a view to assess the potential for implementation of CO_2 Capture and Storage (CCS) schemes.

CCS implementation on SSLNG

- As per baseload LNG plants, post-combustion capture would be the most viable CCS route for SSLNG plants. However, post-combustion capture using solvents requires the availability of large plots of land and provision of power and heating.
- CCS schemes on SSLNG plants would not benefit from economies of scale, and the specific capture cost (€/t CO₂) of SSLNG plant will be higher than the specific capture cost of a baseload LNG plant.
- Simplicity, availability, reliability and quick project implementation are key selling points for SSLNG plants, therefore it is likely that the introduction of a CCS scheme which adversely affects these attributes would be unfavourable.
- SSLNG plants are not likely to be a main priority for implementation of CCS schemes as means to achieve global decarbonisation targets. The key reason is that the global capacity of SSLNG plants is low and as such, associated net emissions from the key emission sources are small and distributed across a large number of plants. Global CO₂ emissions from SSLNG are an order of magnitude smaller in comparison with emissions from baseload LNG plants and three orders of magnitude smaller than global CO₂ emissions from power plants. It is likely that other CO₂ abatement strategies will be the focus of attention for SSLNG plants before the decision is made to implement CCS schemes.

CCS implementation on FLNG

- The FLNG industry is still at an early stage of development, where technical and commercial feasibility is still to be proven at production scale, and only gas fields representing a low risk (e.g. fields in shallow water requiring little pre-treatment) will be considered.
- CCS in either form would introduce significant risks. CO₂ capture from combustion processes is technically unfeasible, with space/weight and interface requirements being the main constraints. Sequestration of CO₂ separated in feed gas pre-treatment appears to be the only feasible CCS scheme, but this introduces considerable design, operating, and economic challenges. Transfer of CO₂ to storage will require extensive transport infrastructure (interfaces with subsea systems, pipelines etc.) which will add complexity, cost and environmental risks.
- As per SSLNG plants, FLNG plants are not likely to be a main priority for implementation of CCS schemes as means to achieve global decarbonisation targets due to the small contribution to global emissions. It is likely that other CO₂ abatement strategies (e.g. improving the process efficiency and reducing the specific emissions of gas turbines) will be the focus of attention.

3.0 BACKGROUND

Natural gas demand is forecasted to grow continuously for the next 10 years, playing a vital role in the global energy mix in 2030. In the particular case of Liquefied Natural Gas (LNG), projections indicate a continued considerable growth. The majority of near-term growth in liquefaction capacity is likely to happen in North America and Australia, although a number of other projects have the potential to add significant liquefaction capacity in the long term as well.

As a key contributor to the global energy supply, the LNG supply chain is expected to be subject to global requirements on reduction to greenhouse gas (GHG) emissions. While the majority of emissions in the LNG supply chain are at the end user (i.e. power plant), the LNG liquefaction plants produce a significant proportion of the total chain CO_2 emissions. The majority of the CO_2 emissions from LNG liquefaction plants arise from combustion of fuel for cycle compressors drive and power generation purposes. It is also necessary to remove and dispose of CO_2 from the natural gas feed, to avoid solidification in the liquefaction process.

The CO_2 emissions depend on the liquefaction plant configuration (e.g. feed gas pretreatment, liquefaction technology, heat and power integration) and also local ambient conditions. The LNG industry has already improved the overall thermal efficiency of the LNG supply chain to reduce the associated CO_2 emissions. LNG plants may provide an early opportunity for CO_2 capture and storage (CCS) subject to the availability of highconcentration CO_2 streams and vicinity to storage reservoirs.

Based on this, the International Energy Agency (IEA) GHG programme considered necessary to evaluate the techno-economics of different CO₂ capture technologies in LNG plants in detail.

4.0 SCOPE

IEAGHG commissioned Costain to develop a techno-economic evaluation and cost assessment of different CO₂ capture technologies for LNG plants including for Small Scale and floating (FLNG) plants.

The scope of work comprises three main tasks:

- 1) Technical evaluation of different CO₂ capture technologies for LNG plants (Ref.1)
- 2) Cost assessment of different CO₂ capture technologies for LNG plants (Ref. 2)
- 3) Feasibility study of different CO₂ capture technologies at small scale and FLNG Plants

This report summarises the technical investigations and evaluations developed for Task 3.

A review undertaken of CO_2 capture technologies at small scale and FLNG plants is summarised within this report. Suitable CO_2 capture technologies for Small Scale and FLNG plants are highlighted as well as any key differences that exist in comparison with baseload LNG plants.



5.0 OVERVIEW OF SMALL SCALE AND FLOATING LNG

5.1 Introduction

Significant global growth within the LNG industry has been experienced within recent years and projections indicate that this is likely to remain true in the coming years with global LNG liquefaction capacity forecasted to continue to rise. There has been a strong focus on building LNG plants with high liquefaction capacities which benefit from economies of scale to process the gas from the increased number of natural gas reserves being brought on stream. However, project developers may look for other ways of getting natural gas to the market due to increased pressure on gas price due to low oil prices and competitiveness in supply.

The present report evaluates the potential for CO_2 capture and storage (CCS) for liquefaction plants with LNG capacities in the indicative range 0.3 to 1.5 mtpa i.e. below the capacity of baseload liquefaction plants. These includes plant referred to as Small Scale LNG (SSLNG) and Floating LNG (FLNG) plants. It is recognised that there are particular design, operational and economic considerations applicable to (onshore) small scale plants and (offshore) floating LNG plants. However, these plants are considered in conjunction for the purposes of this evaluation based on the range of liquefaction capacities. Where relevance is significant, particular considerations of each type of plant (i.e. SSLNG and FLNG) with respect to the potential for CCS implementation are identified.

5.2 Small Scale LNG Plants

SSLNG plants have a capacity in the order of 1 mtpa and below. The global installed production capacity of SSLNG is of the order of 20 mtpa distributed across more than 100 plants (Ref. 3). It is expected that SSLNG capacity will increase towards 30 mtpa by 2020. This constitutes only a fraction of the total global capacity from baseload LNG plants, which currently stands at over 300 mtpa distributed in about 100 liquefaction trains in over 40 export terminals.



Table 1. 0.44 mtpa SSLNG plant (Ref. 3)



Most of the SSLNG growth is in China due to the need for clean fuels for transport and the availability of gas. SSLNG plants can also be found in Japan, Spain, Portugal, Turkey and Norway.

Small scale LNG tends to fit markets such as transport fuels, peak shaving facilities and demand (domestic and industrial) from remote regions. Technology selection is dictated by the sector demands such as intermittent service, minimum capital cost (at the expense of thermal efficiency), variable loads, etc.

The drivers for SSLNG plants differ to some extent to those for baseload LNG plants, particularly associated to the availability to smaller gas reserves in remote locations, with monetisation via supply to nearby markets having a suitable demand. Replacement of diesel with LNG for transport in response to environmental concerns creates a market for SSLNG. Use of LNG (supplied by SSLNG) for transport is particularly motivated by large differences between natural gas and oil prices.

Main factors dictating differences between baseload and SSLNG plants (or factors dictating the applicability/justification of SSLNG plants) are outlined below:

- Natural Gas Reserves The size and/or the production profile of natural gas source may be such that the construction of a high capacity LNG plant would not be justifiable and small scale LNG would offer a more economically attractive solution over the lifetime of the plant.
- Natural Gas Demand and Location Where demand for natural gas is low and geographically stranded with no connection to the natural gas supply via pipeline. Baseload LNG liquefaction plants are typically located in coastal regions. Transportation of LNG provides justification for small scale LNG liquefaction plants local to relatively lower demand geographically isolated markets.
- Gas Storage Small scale LNG liquefaction plants can be positioned across the natural gas supply network as satellite stations which can be used for gas storage (as LNG) or peak shaving service at times when gas demand is low, releasing the stored gas back into the distribution network when the demand is high.
- Availability of Financing Small scale LNG liquefaction plants are significantly cheaper than baseload LNG liquefaction plants which facilitates access to the investment required to cover the project costs.
- Readiness Baseload LNG liquefaction plants are often significantly more complex in both design, project execution and operation than SSLNG plants. SSLNG plants typically have shorter start-up time from the final investment decision (FID) than baseload LNG plants, which allows sale of LNG and associated profits to be made in a shorter period.
- Space Availability of space may provide basis for small scale LNG plant which are much smaller than baseload LNG plants.

The challenge of SSLNG is the relatively expensive supply chain due to the absence of economies of scale. Despite of this, SSLNG has seen growth as LNG can be produced at remote locations and distributed to end-users.

5.3 Floating LNG Plants

Floating LNG (FLNG) refers to LNG liquefaction plants (also called LNG FPSO) designed for operation either offshore or close to the shore on a floating platform (barge or vessel). The unit is provided with gas reception facilities, gas pre-treatment, liquefaction, products storage (usually in the vessel hull), offloading systems (transfer to LNG carriers), power generation and utilities.

FLNG plants aim to market gas from small stranded offshore fields which cannot be economically developed by building a pipeline to an onshore liquefaction plant. The projects also aim to minimise the permitting and regulatory approval requirements and processes associated with onshore projects. Cost reduction is also targeted by considering offsite modular construction in cost competitive shipyards as opposed to stick built construction in isolated regions.

Capacities of FLNG plants are of the order of around 1 mtpa per train as dictated by gas reserves, space and weight constraints and overall economics. Train capacities of current projects range from 0.5 mtpa to a maximum of 3.6 mtpa (Ref 6). Capacities in the low range are increasingly being evaluated as this may facilitate project economics, secure off-takers and reach Financial Investment Decision (FID) faster.

A considerable number of FLNG projects have been announced, particularly in the last decade, with projects progressing into the concept and FEED stage. However, only a limited number of projects have reached the execution phase due to significant challenges faced by the industry. These include technical complexities (e.g. equipment and process marinisation, technology maturity, space and weight constrains), project definition considerations (consideration of gas fields with ideal conditions for initial deployment of FLNG, with manageable risks and uncertainties) and commercial challenges (cost escalation, overall project economics and access to financing).

Four FLNG projects in Australia, Malaysia, and Cameroon have been sanctioned and moved to the construction phase (see Table 2). Total liquefaction capacity from these projects is 8.7 mtpa.

Country	Project	Start	Capacity	Liquefaction	Refrigerant		
Country	(Note 1)	Year	(mtpa)	Technology	Compressor Driver		
Australia	Prelude FLNG	2018	3.6	Shell DMR	Steam		
Cameroon	Cameroon FLNG	2017	2.4	B&V PRICO (SMR)	PGT25+G4		
Malaysia	PFLNG Satu	2017	1.2	Air Products AP-N	PGT25+G4		
Malaysia	PFLNG 2	2020	1.5	(dual N ₂ expander)	LM6000-PF+		

Table 2. Sanctioned FLNG Projects

Notes:

1. Cameroon FLNG is based on a Moss type LNG carrier (conversion) fitted with liquefaction equipment. Other projects use a purpose-built vessel.

24 FLNG proposals have been announced, mainly in the United States, Canada, and Australia. Others are proposed to be located in Congo, Djibouti, Equatorial Guinea, Indonesia, Iran, Mauritania and Senegal, Mozambique, Papua New Guinea, and Russia. Total capacity from these proposals is 157 mtpa.


Table 3. 3.6 mtpa Prelude FLNG Plant



Similar to SSLNG plants, the main driver for FLNG is the monetisation of stranded (offshore) gas fields with gas reserves below those at which a baseload plant is economical, or such that investment in pipelines (to a new or existing onshore liquefaction plant) cannot be justified.

- Natural Gas Reserves FLNG plants allow access to smaller, remote and stranded offshore gas fields that would otherwise be too risky or economically unattractive to develop. For example, the pipeline transporting the natural gas from source to an onshore facility may be too complicated or too long leading to unacceptably high cost or the size of the natural gas source may be too small to justify the cost of an onshore facility.
- Location Onshore liquefaction plants are fixed to a permanent location whereas the philosophy of FLNG plants is for these to be retractable and have potential for redeployment in a different location by modifying original design at a fraction of the original cost. Factors affecting the location of onshore liquefaction plants such as the requirement of coastal location close to the sea front and water depth are not relevant for FLNG. There is also potential for cost optimisation in areas with high construction costs (such as Australia) with plant modules and hull fabricated in the most cost competitive yards.
- Early Production System Where FLNG plants are already available, these can be used as an early production system during the development of a new natural gas source which speeds up time to production. This enables a cash flow to be established earlier in the project lifecycle to ease financial pressures.
- Environment FLNG allows for reduced impact on the environment by removing the need for long pipelines on the ocean floor and removing the need for compression of gas to the onshore facility and the associated emissions etc.
- Safety FLNG plants located far offshore have less public exposure than onshore liquefaction plants. Operators, on the other side, would face increased challenges in this regard as per offshore production.



5.4 Liquefaction Technologies

In baseload LNG plants the propane pre-cooled mixed refrigerant (AP-C3MR[™]) process is most widely used (Ref. 1) on the basis of high efficiency, technology mature and use of proven and conventional equipment available. Ultimately, the capacity of the gas turbines driving the refrigerant compressors and their configuration dictate the optimum liquefaction train capacity.

The selection of liquefaction technology for SSLNG and particularly for FLNG plants may be driven by additional considerations. For example, size and weight constraints, potential loss of performance due to motion impacting liquids handling equipment and risks associated with the storage and use of flammable refrigerants may be more attractive than efficiency for FLNG. Consequently, other liquefaction technologies become more prevalent for SSLNG and FLNG liquefaction plants.

Two main liquefaction process types are currently used for SSLNG, i.e. expansion cycle processes and single mixed refrigerant processes. These can include (propane) pre-cooling, but this is not a common feature, with only a few plants currently installed with pre-cooling. Table 4 shows the main SSLNG liquefaction technologies.

Expansion cycle and SMR processes are expected to lead to very similar capital costs, with main differences in operating costs and plant operability. For operation load near design capacity, SMR is expected to have lower operating costs due to higher efficiency. Nitrogen expander cycles are preferred for plants with low running hours and wide part-load profiles (operating costs associated to low efficiency are less relevant) and for plants in remote areas, removing dependency of supply of refrigerants for make-up (Ref. 3) if on-site refrigerant production is not possible.

Technology Type	Process	Licensor		
	PRICO	Black & Veatch		
	AP-SMR	Air Products		
	LiMuM	Linde		
0.40	SCMR	Kryopak		
SIVIR	PCMR (Note 1)	Kryopak		
	Single MR	Chart		
	KSMR	KOGAS		
	OSMR (Note 1)	LNG Limited		
	Single N ₂ Expander	Various (Open)		
Expansion cycle	Dual N ₂ Expander	Various (Open)		
	NDX-1	Mustang		
	OCX	Mustang		
	OCX-R (Note 1)	Mustang		
	C3-N ₂ Expander (Note 1)	Air Products		
	Niche LNG (Note 1)	CB&I Lummus		

Table 4. Main SSLNG	Liquefaction	Technologies	(Ref. 3)
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Notes:

1. Process technology includes pre-cooling step

Liquefaction technologies on offer for FLNG fall under four main categories: single mixed refrigerant, dual mixed refrigerant, nitrogen expander technology and cascade.

- **Single Mixed Refrigerant (SMR)** Use of a SMR for pre-cooling, liquefaction and subcooling, with operational experience on SSLNG plants. It allows for reduction in equipment count for FLNG at the cost of a reduction in power efficiency in comparison with a pre-cooled MR process. Train capacity is up to 1 mtpa.
- **Dual MR (DMR)** It is a development of the AP-C3MR[™] process but uses two separate mixed refrigerant cycles which are cascaded to achieve a better match to the natural gas cooling curve. It is a more complex process than SMR but it gives a high efficiency, and potentially a more compact plant. DMR technology has the highest capacity per train which is typically from 1.5 mtpa and potentially up to 5 mtpa LNG.
- Expander Technology Train capacity is limited to about 1.5 mtpa, but it is often attractive due to the reduction in flammable inventory since non-flammable nitrogen is the sole refrigerant used (with flare size being reduced) and there is no need to import refrigerant for make-up. Also, it is not sensitive to motion as there is no liquid evaporating refrigerant. The efficiency is lower than SMR or DMR, but it can be increased (together with capacity) by introducing pre-cooling.
- Cascade Three discrete stages of refrigeration using propane, ethane and methane each generally with multiple expansion and compression stages. The additional stages may offer lower power requirements than other technologies at the expense of capital cost due to higher equipment count (although optimised Cascade and DMR processes have similar efficiencies). The technology could be viable for large train capacities in the order of 5 mtpa.

DMR and nitrogen expander cycle processes have generally been evaluated and selected (e.g. see Table 2). DMR has been selected due to its high efficiency and large production capacity in a single train (as dictated by reserves and project economics), whilst nitrogen expander cycle processes have been selected based on safety considerations (elimination of flammable refrigerants) and insensitivity to FLNG vessel motion.

5.5 Refrigeration Cycle Compressor Drives

As per baseload LNG plants, gas turbines are usually installed as refrigerant compressor drives. Heavy frame industrial gas turbines are conventional and proven in mechanical drive service. These will continue to be used specially in cases where efficiency is not a differentiator or for small liquefaction train capacities (<0.5 mtpa) where a more efficient aeroderivative gas turbine equivalent is not available.

For the particular case of FLNG plants where power requirements are lower and/or weight limitations apply, aeroderivative gas turbines are typically selected for the following reasons (see also Reference 1):

- High power to weight ratio
- Proven availability and reliability
- Modular engine sections that can be serviced/maintained easily
- Higher thermal efficiency in comparison with industrial units



Aeroderivative gas turbines offer 10-15% more efficiency compared to industrial type gas turbines This leads to a reduction in the specific emissions for SSLNG and FLNG plants opting for this turbine type.

A selection of compressor drives and their associated available power and efficiency is shown in Table 5.

Refrigerant Compressor Driver	Power @ 30°C (MW)	Efficiency @ 30°C (%)	LNG Plant Capacity (mtpa) (Note 1)
LM2500+	26.1	38.5	0.6
LM2500+(G4)	28.6	38.9	0.7
LM6000PC	33.7	38.1	0.8
LM6000PG	40.2	39.3	0.9
RR Trent	43.6	40.0	1.0

Table 5. Example Aeroderivative Compressor Drives (Ref. 5)

Notes:

1. Nominal liquefaction train capacity per turbine, i.e. single turbine driving refrigerant compressor(s), equivalent for power output indicated, specific liquefaction power 0.35 kWh/kg LNG and 93% plant availability.

Steam turbines have been considered for refrigerant compressor drive in the Prelude FLNG project. Boilers generate high pressure steam used to drive the refrigerant compressors and turbines that generate electricity, as well as low pressure steam for process heat. While steam turbines may not be as energy efficient as gas turbines, factors promoting the selection of steam driven turbines may include (Ref 7):

- High reliability which is proven in marine environment (steam boilers have been used on LNG ships for many decades)
- Ease of operation and maintenance
- Reduced complexity due to reduction in rotating equipment
- Allows for the use of low pressure fuel gas, which reduces energy requirements for compression of end flash gas for fuel gas
- Flexibility in fuel gas composition (advantageous due to potential redeployment of FLNG)
- Allows movement away from the use of very large variable speed electric motor drivers and/or fired equipment in the liquefaction module on FLNG.

Design efficiency of gas turbines will be maximised for operation at baseload. It is expected that FLNG plants will operate at baseload, with liquefaction train capacity dictated by the output power of gas turbine drives. SSLNG plants may operate at part-load depending on demand or in general due to unavailability of one of the elements the supply chain (demand included). SSLNG plants in peak-shaving service may operate at baseload intermittently. Part-load operation will lead to gas turbines operating at reduced efficiency, and therefore increasing fuel gas usage and associated specific operating costs and emissions.

5.6 Liquefaction Efficiency

Fuel gas usage and associated CO₂ emissions are a function of liquefaction efficiency (i.e. specific power). Table 6 compares the representative power consumption of liquefaction technologies relative to cascade cycle which has a typical specific power consumption of approximately 0.33 kWh/kg LNG (Ref 4). These are representative figures and depending on the particular operating conditions, the specific efficiency figures may vary.

For FLNG plants, the DMR technology is also applicable which has a typical specific power similar to the AP-C3MR[™] process (Ref 11). Table 6 generally shows that cascade and mixed refrigerant processes have similar efficiencies when compared to nitrogen cycles.





5.7 CO₂ Emission Sources

For SSLNG plants emission sources are likely to be the same as baseload liquefaction plants, albeit in lower quantities consistent with smaller plant capacities. In general, there are two main sources of CO₂ emissions.

A detailed discussion and identification of all potential sources of CO₂ emission sources is given in Reference 1.

5.7.1 CO₂ in reservoir gas

 CO_2 removed in the pre-treatment stage and vented to atmosphere. CO_2 emissions are in function of CO_2 content in feed gas (on the basis that practically all of the CO_2 is removed) and LNG plant capacity (i.e. feed gas flow) and independent of liquefaction technology and efficiency.

 CO_2 removal from feed gas represents a significant processing burden and it is an energy intensive process. For FLNG the design and operation of absorption and stripper columns represents a challenge due to loss of performance associated to vessel motion. Generally, it is expected that preference will be given to develop fields with a low CO_2 content with project economics acting in favour of the reduced gas processing duty. Indicatively it is expected that this would be the case for reservoir gas containing a maximum CO_2 in the order of 2 mol%.



This is particularly the case of FLNG given the technical challenges and tight economics. FLNG plants may be limited to processing low CO_2 content feed gas in order to reduce the size and weight of the pre-treatment facilities due to low availability of deck space and stringent weight limitations imposed on the FLNG vessel. Exceptions exist depending on the particular project drivers e.g. Prelude FLNG (processing gas with relatively high CO_2 9 mol%) and Browse LNG (cancelled, proposed to process up to 12 mol% CO_2 gas).

5.7.2 CO₂ in flue gases

 CO_2 produced as a result of fuel gas burning for power and heat generation. CO_2 emissions are proportional to the total fuel gas being combusted to provide the power and heating required by the liquefaction plant. Fuel gas usage is a direct function of the total power demand and thermal efficiency (a combination of liquefaction efficiency and gas turbine efficiency). Reported thermal efficiencies for baseload liquefaction process are in the range 90-93%, with the balance 7-10% representing the fuel consumption (Ref. 1).

SSLNG plants may opt to run their main rotating equipment on electric power. Due to the lower power requirements of SSLNG compared to baseload, the electric power may be supplied by the public grid as opposed to a dedicated electrical power plant. Equivalent emissions associated with the production of the required electricity will be accounted for the liquefaction process.

Liquefaction specific power requirements for SSLNG and FLNG are typically higher than for baseload LNG plants, therefore the fuel gas usage and specific emissions (tCO₂/tLNG) will increase. The lower efficiency of SSLNG and FLNG processes compared to baseload processes may result in fuel gas usage and associated specific emissions in the order of up to 50% higher, particularly for the case of nitrogen expander processes.

Relatively high specific emissions of 0.64 tCO₂/tLNG have been reported for Prelude FLNG (Ref 8, 10), in comparison to specific emissions of baseload LNG plants ranging between approximately 0.3-0.6 tCO₂/tLNG (Ref 1). This is due to a combination of low liquefaction efficiency mainly attributed to the use of steam boilers/turbines and processing of reservoir gas with a high CO₂ content, see Table 7. The major source of fuel combustion emissions are the steam boilers. The compressors and turbines running off the steam generated by the boilers do not themselves generate any additional CO₂ emissions.

Reservoir CO ₂	0.97 mtpa		
CO ₂ from fuel combustion	1.26 mtpa		
CO ₂ from flaring	0.06 mtpa		
Total CO ₂ emissions	2.30 mtpa		
LNG production (Note 1)	3.60 mtpa		
Specific CO ₂ emissions (Note 1)	0.64 tCO ₂ /tLNG		

Table 7. CC	emissions	from	Prelude	FLNG	(Ref	10)
	2				·····	•••

Notes:

1. Prelude facility will produce 3.6 mtpa of LNG plus 0.4 mtpa of LPG and 1.3 mtpa of condensate) The specific CO2 emissions from the facility are allocated to LNG production only.



7.0 EVALUATION OF CCS POTENTIAL

Evaluation of the potential implementation of CCS on SSLNG and FLNG plants is provided in this section together with applicable considerations for each.

A review of CCS routes (pre/post/oxyfuel combustion) and technologies (absorption, adsorption, cryogenic, membranes) and evaluation of CCS potential for baseload plants has been provided in Part 1 of this study (Ref 1) and it is applicable for SSLNG and FLNG.

7.1 CCS on SSLNG Plants

7.1.1 SSLNG Development

It is recognised that there will be an increase in the construction of SSLNG plants in regions where governmental policies in support of implementation of LNG are either already in place or are being developed. In Europe, the European Union Clean Power for Transport Package will ensure that member states develop national policy framework and LNG bunkering at ports and LNG fuelling stations for heavy-duty vehicles by 2030. Government agencies in the United States are introducing tax incentives to encourage the purchase and use of natural gas vehicles. In China there is significant attention on reducing emissions through the use of clean fuels particularly to tackle pollution within the major cities. While attention is on the use of alternative fuels, CO_2 emissions in general are being targeted, although for places such as China the focus is likely to be on CO_2 emissions associated with coal.

7.1.2 Technical Considerations

Of the three CCS routes, post-combustion capture is the most viable CCS route for SSLNG plants. Pre-combustion and oxyfuel combustion processes are likely to introduce significant complexity to the LNG plant impacting on both ease of operation and project schedule. As for baseload plants, LNG production would be impacted if the production of syngas (pre-combustion) or oxygen (oxyfuel combustion) was lost, impacting upon the availability of the SSLNG plant. Loss of operation of post-combustion capture would not impact on LNG production.

Post-combustion capture using solvents requires large plot areas, large column heights and high energy requirements all of which are likely to be unattractive for SSLNG unless sufficient plots of land and provision of power and heating are available.

SSLNG plants may provide a gateway for CO₂ capture technologies to the LNG industry by providing a stepping stone from demonstration scale tests to baseload LNG plant scale. SSLNG plants would then allow confidence to be gained in the process at a larger scale than demonstration plants before expanding to baseload LNG scale where some key items of equipment are likely to fall outside the limits of experience.

Simplicity, availability, reliability and quick project implementation are key selling points for SSLNG plants. It is likely therefore that the introduction of a CCS scheme which adversely affects these attributes would be unfavourable.

SSLNG plants located in geographically stranded regions are likely to experience difficulty in finding a suitable storage location for captured CO_2 unless transport infrastructure (i.e. CO_2 pipelines) and suitable geographical storage exists or potential for an 'over the fence' type agreement to be established whereby CO_2 can

be sold to a third party (industrial use of CO_2 as feedstock is considered to represent temporary/short term sequestration). SSLNG plants are typically located in these regions as it is deemed unattractive (for many possible reasons) to connect the stranded region to the main natural gas supply pipeline via a pipeline. It is therefore likely that transportation of CO_2 over large distances from these regions to suitable storage locations would encounter the same difficulties as a natural gas pipeline would. This is a significant barrier to the implementation of CCS schemes.

7.1.3 Commercial Considerations

If environmental constraints are in place such as emission charges (CO_2 tax), the economic attractiveness of implementing CCS on SSLNG plants may increase (in the long term). It may even be the case that developments of new SSLNG plants are benefitted by the inclusion of emission abatement strategies by reducing levelised LNG costs for the duration of the project in comparison with a plant with emissions charges.

As a result of economies of scale, the specific capture cost (\notin /t CO₂) for a capture plant on a SSLNG plant will be higher than the cost of a capture plant on a baseload LNG plant. This will make financing and economic project feasibility overall more challenging for SSLNG plants than for baseload LNG plants (or large-scale capture plants in general)

Despite specific emissions being significantly higher as noted in Section 5.7.2, total CO_2 emissions (t CO_2 /day) of SSLNG plants will be lower than the emissions of a baseload plant, but this is only due to reduced size of plant. Based on the example plant capacities used in Table 8, the size of the capture plant for the SSLNG plant is about 30% of the capture plant capacity of the baseload LNG plant. Expected CAPEX for the small-scale capture plant is lower, although still considerable and to some extent comparable to the CAPEX of a capture plant on a baseload liquefaction plant.

		Baseload LNG	Small
		(Note 1)	Scale LNG
LNG Plant Capacity (example)	mtpa	4.6	0.8
Specific CO ₂ Emissions (Note 2)	tCO ₂ /tLNG	0.24	0.36
CO ₂ Capture Plant Capacity	mtpa	1.0	0.3
Total Capital Requirement	€ million	755	300-500

Table 8. CO₂ Emissions, Capture Plant and Cost Comparison

Notes:

- 1. Data for Baseload plant from Reference 1
- 2. Emissions from combustion processes only (gas turbines/boilers)

7.1.4 Potential for CCS Implementation

SSLNG plants are not likely to be a main priority for implementation of CCS schemes for emission abatement, at least in the short or medium term. The priority for implementation of CCS as means to achieve global decarbonisation targets follows the same considerations as per industrial CCS (i.e. capture schemes implemented on emission-intensive industrial operations like cement production,

refineries, steel industries, chemical plants and other emitters, excluding power generation). The key reason is that the global capacity of SSLNG plants is low and as such, associated net emissions from the key emission sources are small (and distributed in a large number of plants) in comparison with emissions from baseload LNG plants. Net emissions are even lower when compared to emissions from other industries such as large combustion plants, particularly coal-based power plants.

- Global CO₂ emissions from power generation plants are in the order of 10,000 mtpa. This corresponds to about one third of the total emissions from fossil-fuel use.
- The emissions associated to the global baseload LNG liquefaction capacity (300 mtpa of LNG) are in the order of 75-100 mtpa of CO₂ (excluding emissions of CO₂ separated in feed gas pre-treatment).
- The emissions associated to the global SSLNG liquefaction capacity (20 mtpa of LNG) are in the order of 8-10 mtpa of CO₂.

It is likely that other CO_2 abatement strategies will be the focus of attention for SSLNG plants before the decision is made to implement CCS schemes. This includes means to improve the efficiency (or reduce the specific emissions) such as the use of aeroderivative gas turbines or electric motors. A particular consideration for SSLNG plants is the potential for power to be supplied from the local grid due to the reduced power demand in comparison with a baseload LNG plant. This may provide more effective means of CO_2 capture from a larger centralised source (i.e. power plant).



7.2 CCS on FLNG Plants

7.2.1 FLNG Development

Although a relative large number of FLNG projects have been proposed in the last decade, only a small number of projects have seen actual implementation in recent years, with projects moving from the concept design and FEED to project execution. A number of associated technical and commercial hurdles exists, such as the technical challenges that a marine environment poses in design (e.g. impact of motion in equipment design and performance and in operations) and the difficulties to access financing due to the associated risks.

It would be expected that following technical and commercial demonstration of projects currently in execution, the development of the industry can be accelerated, subject to LNG market dynamics.

7.2.2 Technical Considerations

Specific emissions in a FLNG process are expected to be higher than the specific emissions of a baseload LNG plant depending in part on the liquefaction technology. Plants using nitrogen expander technology are expected to have the highest specific emissions due to the low liquefaction efficiency. FLNG plants using mixed refrigerant liquefaction technology can potentially achieve lower specific emissions (than nitrogen expander technology). The lowest specific emissions would be expected for DMR liquefaction technology, which has specific emissions similar to the AP-C3MR[™] liquefaction technology commonly used on baseload LNG plants (Ref 11).

It is noted that these remarks are only based on typical efficiency of liquefaction technologies. Optimisation of technologies for the specific project conditions including compressor driver selection may lead to improvements in overall thermal efficiency. Final selection of technology may be driven by other factors, including reliability, safe and robust operation, with a liquefaction efficiency penalty being acceptable. It is unlikely that differences between achievable efficiencies of process technologies for FLNG plants will be the deciding factor in the decision to implement CO_2 capture.

Post-combustion capture based on chemical absorption has been deemed the most appropriate for onshore baseload LNG plants. However, it is unlikely that this or any other capture route from combustion processes will be feasible for FLNG plants.

Chemical absorption based capture has a significant footprint for the treatment of large quantities of low pressure flue gas. For a FLNG plant, there are space and weight limitations, optimised for the gas pre-treatment, liquefaction and utilities. There is no spare space for further topsides equipment, which makes integration unfeasible. The options to have either a dedicated floating capture plant (e.g. on a barge) or an onshore capture plant treating flue gas for a FLNG unit located at shore also appear to be unfeasible.

Considering the concept of a floating capture plant, this would be subject to wave motions, which will impact on the design and operation of a chemical absorption based capture plant. Use of absorption technology involves the use of tall, heavy columns containing liquids, therefore introduces issues associated with equal distribution of two phase and liquid flow. Wave induced motion, which is a function of sea motion and vessel size, introduces acceleration forces and increased mechanical fatigue, increased complexity in design and higher cost for equipment. Incorporating the tall columns associated with post-combustion chemical solvent based capture into the design of the vessel (layout, size and weight constraints) is also likely to be a significant challenge.

Regeneration of the chemical solvent used in post combustion CO₂ capture requires a significant heat duty. The most cost effective and thermally efficient way of providing heat is via waste heat recovery from the LNG gas turbines, with interfaces representing a significant design challenge.

Additional power is likely to be needed to meet the requirements of the capture plant which will impact FLNG vessel design further in terms of space availability, layout and weight constraints etc. LNG production economics and feasibility may also be impacted due to the requirement of fuel gas for the additional power negatively impacting LNG production.

It might be considered that post-combustion capture prospects may improve for FLNG located at the shore line, with capture plant installed onshore. Although space and weight limitations can be removed, the interfaces required (e.g. to transfer flue gas from the FLNG plant to the capture plant) still make the scheme likely unfeasible.

Oxyfuel and pre-combustion do not offer any real prospects to make feasible the implementation of CO_2 capture from combustion processes, due to the issues inherent to these routes and technologies (Ref 1) and the issues discussed above for post-combustion.

Sequestration of reservoir CO₂ separated in the pre-treatment acid gas removal process appears to be the only feasible CCS scheme. The scheme may still pose significant technical challenges for design and operation of FLNG plants, and will impact the overall project economics.

Sequestration of captured reservoir CO_2 is limited to compression, potentially liquefaction for pumping (either standalone or integrated with the LNG process) and injection into storage. FLNG plants do offer the potential for sea water to be used as cooling medium for compression aftercoolers, which leads to equipment size reductions. Still, a considerable plot area and weight are added to install the required equipment (compressor, coolers, vessels, pumps, large bore piping and associated power generation capacity).

Fields with high CO_2 content will provide the greatest potential for CCS. However, from the liquefaction perspective (and project economics and hence commercial feasibility) it is likely that fields with a low CO_2 content would be selected for development with a FLNG scheme, which reduces pre-treatment duties and associated costs.

Transfer of CO_2 to a suitable storage site is likely to impose significant challenges and financial and environmental implications to the project. Interfaces with subsea systems pose a significant challenge in the design of a FLNG plant. Introducing the requirement to inject CO_2 will increase complexity and cost of design. Modifications to the design of the turret to allow for additional CO_2 return pipework would be required.



Locality of suitable storage is a key consideration. Storage located far from the FLNG plant will require extensive transport infrastructure (pipelines, etc.) which will add complexity, cost and environmental risks to the project. Transfer of liquid CO_2 via tankers is likely to be unviable due to the need for intermediate liquid CO_2 storage, high volumes of CO_2 overall, high pressure required for storage of CO_2 in liquid phase (which has density of more than double that of LNG), and impact on LNG offloading.

7.2.3 Commercial Considerations

As per SSLNG, the economic attractiveness of implementing CCS on FLNG plants may increase when environmental penalties (CO_2 tax) are in place. There does not appear to be potential for this in the short of medium term. Even if this was the case, alternative options (e.g. priority to develop fields with low reservoir CO_2) may be considered given the low technical feasibility of the CCS schemes.

7.2.4 Potential for CCS Implementation

The FLNG industry is still at an early stage of development, where technical and commercial feasibility is still to be proven at production scale, and only gas fields representing a low risk (e.g. fields in shallow water requiring little pre-treatment) will be considered.

CCS in either form would introduce significant risks. CO_2 capture from combustion processes is technically unfeasible, with space/weight and interface requirements being the main constraints. Sequestration of CO_2 separated in feed gas pre-treatment introduces considerable design, operating, and economic challenges.

As per SSLNG plants, FLNG plants are not likely to be a main priority for implementation of CCS schemes for emission abatement due to the small contribution to global emissions. It is likely that other CO_2 abatement strategies will be the focus of attention to contribute to meeting environmental targets. Efforts will be focused to evaluate ways to improve the process efficiency and reduce the specific emissions of gas turbines.



8.0 CCS PROJECTS IN FLNG PLANTS

Potential for CCS and emissions abatement has been considered in the design of projects under execution, particularly by those projects where the relatively reservoir CO₂ contributed significantly to the plant emissions, i.e. Prelude FLNG and Browse FLNG.

8.1 Prelude FLNG

Due to the relatively high CO_2 in feed gas, large plant capacity (from the FLNG perspective) and low efficiency (e.g. due to the use of steam turbines), Prelude FLNG has significant net emissions, comparable and potentially higher to the emissions of a typical baseload onshore liquefaction plant.

At plateau production levels, Prelude FLNG is reported to be expected to emit approximately 1.26 mtpa CO_2 from flue gas and 1 mtpa from reservoir gas (Ref 12). Prelude FLNG opted against capturing CO_2 from reservoir gas or flue gas sources due to reasons including (Ref 9 and 10):

- Capture and storage of CO₂ from flue gas was considered unfeasible as the technology was not considered to be proven, required too much equipment space and deemed to be too expensive.
- Despite high reservoir gas CO₂ content (~9% CO₂), a major contributor to the overall CO₂ emissions, storage in geological formations was determined to be technically and commercially unviable due to the associated significant challenges, specifically regarding re-injection of CO₂ as no suitable storage site was available.
- The storage of reservoir CO₂ had significant cost and unresolved technical uncertainties which added complexity to the FLNG design.
- Storage of the reservoir CO₂ required the drilling of an additional well for the injection of CO₂ and the installation of a flow line from the FLNG facility to the injection well, increasing the project's footprint on the sea floor.
- Pre-treatment emissions represent a minor contribution to global atmospheric greenhouse gas overall with no local receptors impacted. The contribution of Prelude FLNG facility during the lifetime of the plant was not qualified to be high.
- Delays and uncertainties regarding the Carbon Pollution Reduction Scheme (CPRS) to incentivise reduction in carbon emissions in Australia. This scheme was deemed as temporary and not significant enough to warrant any change in strategy for the Prelude FLNG Project (i.e. did not merit the inclusion of CCS scheme)

Storage of the reservoir CO_2 was shelved until (if at a later stage in the project) sufficient progress has been made in overcoming the technical issues. Deck space was allowed in the FLNG facility design for a future CO_2 compression module to be installed, should the technical and other issues be overcome.



8.2 Browse FLNG

Browse FLNG identified reservoir gas and flue gas as the two potential sources for capture of CO_2 . However, it opted against capturing CO_2 from these sources due to considerations including:

- Capture and storage of CO₂ from flue gas was considered unfeasible as the technology was deemed unproven and requiring too much equipment space.
- Despite the reservoir gas containing on average 10 mol% CO₂ without importing additional CO₂ from adjacent reservoirs, the benefits achieved (recovering additional gas and managing subsistence) through the reinjection of CO₂ were deemed insufficient to compensate for the added technical and financial burdens and was deemed to be grossly disproportionate to the environmental benefit gained.
- Capture of the reservoir CO₂ would have required:
 - Additional subsea infrastructure to support the injection system, including wells and flow lines into a suitable reservoir
 - Additional infrastructure on the FLNG facilities including compression equipment and modifications to the LNG process to capture CO₂
 - Requirements for modification to the FLNG layout, and hull strengthening to support the required additional infrastructure.
- Storage of captured CO₂ within geological formations would introduce increased uncertainty regarding the selected reservoir (e.g. infectivity, capacity, containment and monitoring)
- Location of suitable geological storage was such that an extensive pipeline network would be required which introduced significant cost, technical challenges and environmental risks in the transportation of CO₂ from the FLNG plant to storage.



9.0 SUMMARY

9.1 Small Scale LNG

The global installed production capacity of SSLNG is of the order of 20 mtpa distributed across more than 100 plants. SSLNG Global capacity is expected to increase towards 30 mtpa by 2020. This constitutes only a fraction of the total global capacity from baseload LNG plants, which currently stands at over 300 mtpa distributed in about 100 liquefaction trains in over 40 export terminals.

SSLNG plants use single mixed refrigerant or nitrogen expander cycle liquefaction technologies. Thermal efficiencies are smaller than those achieved by baseload liquefaction processes, but this is generally less relevant to the overall economics of SSLNG plants.

As per baseload LNG plants, it is likely that post-combustion capture would remain the most viable CCS route for SSLNG plants. However, post-combustion capture using solvents requires large plot areas, large column heights and high energy requirements all of which are likely to be unattractive for SSLNG unless sufficient plots of land and provision of power and heating are available.

CCS schemes on SSLNG plants would not benefit from economies of scale. Therefore, the specific cost (\notin /t CO₂) of a capture plant on a SSLNG plant will be higher than the cost of a capture plant on a baseload LNG plant. Access to financing to cover the capture plant costs will be more challenging on the basis that overall project economics allow the inclusion of a CCS scheme.

SSLNG plants located in geographically stranded regions are likely to experience difficulty in finding a suitable destination for captured CO₂ unless transport infrastructure (i.e. CO₂ pipelines) and suitable geographical storage exists.

Simplicity, availability, reliability and quick project implementation are key selling points for SSLNG plants, therefore it is likely that the introduction of a CCS scheme which adversely affects these attributes would be unfavourable.

SSLNG plants are not likely to be a main priority for implementation of CCS schemes as means to achieve global decarbonisation targets, at least in the short or medium term. The key reason is that the global capacity of SSLNG plants is low and as such, associated net emissions from the key emission sources are small and distributed across a large number of plants. Global CO₂ emissions from SSLNG are an order of magnitude smaller in comparison with emissions from baseload LNG plants and three orders of magnitude smaller than global CO₂ emissions from power plants.

It is likely that other CO₂ abatement strategies will be the focus of attention for SSLNG plants before the decision is made to implement CCS schemes.

9.2 Floating LNG

Only four FLNG projects have been sanctioned and moved to the construction phase, with a total liquefaction capacity of 8.7 mtpa. 24 FLNG proposals have been announced with a total capacity of 157 mtpa.

DMR and nitrogen expander cycle processes have generally been evaluated and selected in FLNG projects. DMR has been selected due to its high efficiency (comparable to baseload LNG technologies) and large production capacity in a single train (as dictated by reserves and project economics), whilst nitrogen expander cycle processes have been selected based on safety considerations (elimination of flammable refrigerants) and insensitivity to FLNG

vessel motion. Safe and robust operation generally takes priority for FLNG, with efficiency being less relevant. It is unlikely that differences between achievable efficiencies of process technologies for FLNG plants will be the deciding factor in the decision to implement CO₂ capture.

Post-combustion capture based on chemical absorption has been deemed the most appropriate for onshore baseload plants. However, it is unlikely to be feasible for FLNG plants.

- Chemical absorption based capture has a significant footprint for the treatment of large quantities of low pressure flue gas. For a FLNG plant, there are space and weight limitations, optimised for the gas pre-treatment, liquefaction and utilities. There is no spare space for further topsides equipment, which makes integration unfeasible.
- Additional power is likely to be needed to meet the requirements of the capture plant which will impact FLNG vessel design further in terms of space availability, layout and weight constraints, etc.
- The options to have either a dedicated floating capture plant or an onshore capture plant treating flue gas for a FLNG unit located at shore appears to be unfeasible, particularly due to the interfaces required.

Sequestration of reservoir CO₂ separated in the pre-treatment process appears to be the only feasible CCS scheme. The scheme may still pose significant technical challenges for design and operation of FLNG plants, and will impact the overall project economics.

- Fields with high CO₂ content will provide the greatest potential for CCS. However, from the project perspective, it is likely that fields with a low CO₂ content would be selected for development with a FLNG scheme, which reduces pre-treatment duties, size and associated costs.
- Transfer of CO₂ to a suitable storage site imposes significant challenges and has financial and environmental implications. Introducing the requirement to inject CO₂ will increase complexity (e.g. interfaces with subsea systems) and cost of design. Storage located far from the FLNG plant will require extensive transport infrastructure (pipelines etc.) which will add complexity, cost and environmental risks.

As per SSLNG plants, FLNG plants are not likely to be a main priority for implementation of CCS schemes as means to achieve global decarbonisation targets due to the small contribution to global emissions. It is likely that other CO_2 abatement strategies (e.g. improving the process efficiency and reducing the specific emissions of gas turbines) will be the focus of attention.

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