GHGT-12

Energy and cost evaluation of a low-temperature CO_2 capture unit for IGCC plants

David Berstad[a],*, Simon Roussanaly[a], Geir Skaugen[a], Rahul Anantharaman[a], Petter Nekså[a], Kristin Jordal[a]

[a]SINTEF Energy Research, P.O. Box 4761 Sluppen, NO-7465 Trondheim, Norway

Abstract

The application of CO_2 capture by liquefaction has been investigated for an integrated gasification combined cycle (IGCC). Two configurations of the process are developed – one supplying CO_2 at conditions suitable for pipeline transport and the second one producing liquid CO_2 suitable for ship transport. The liquefaction process for CO_2 capture is more efficient and compact compared to Selexol process for providing CO_2 suitable for ship transport as the separation and liquefaction units are integrated in the process presented in this work. An economic analysis performed shows that CO_2 capture by liquefaction is more cost efficient than corresponding Selexol-based separation processes by 9–11% in terms of the levelized cost of electricity and 35–37% in terms of CO_2 avoidance costs.

Keywords: Pre-combustion; CCS; CO_2; Capture; IGCC; Energy; Cost; Liquefaction

1. Introduction

The major point sources of greenhouse gas emissions vary substantially in capture conditions for CO_2 [1]. Sources with high CO_2 concentration and/or CO_2 partial pressure are generally more attractive for CO_2 capture than those with more diluted CO_2. Integrated gasification combined cycles (IGCC) have rather high CO_2 concentration and partial pressure, in some cases as high as 20–30 bar, and such high CO_2 pressure can be utilised to achieve highly energy- and cost-efficient CO_2 separation and capture.

In the present work, a high-efficiency CO_2 capture unit for IGCC power plants based on CO_2 liquefaction is presented. The capture method is based on low-temperature phase separation (liquefaction), where condensed liquid

---

* Corresponding author. Tel.: +47-930-02-784; fax: +47-735-93-950.
E-mail address: david.berstad@sintef.no

CO_2 is separated from the uncondensed hydrogen-rich vapour phase. The shifted syngas under consideration consists of mainly hydrogen and CO_2, with a concentration of the latter close to 40%.

At GHGT-11, an energy benchmarking was presented [2] for the CO_2 capture by condensation, against Selexol as baseline technology. At 85% CO_2 capture ratio (CCR), the net electric efficiency of the IGCC plant was calculated to be 0.8–1.0 %-point higher using CO_2 condensation relative to the baseline cases with Selexol.

As a follow-up to the GHGT-11 paper [2] this work elaborates on a further techno-economic benchmarking, including the comparison of cost data. Moreover, energy and cost calculations also include a modified version of the CO_2 condensation process producing cold, liquid CO_2 for ship transport.

2. Methodology

2.1. CO_2 separation concept

Different process designs are possible for the low-temperature CO_2 capture and condensation unit. The proposed baseline process flow diagram for producing high-pressure CO_2 for pipeline transport is shown in Fig. 1. Cleaned and shifted syngas from a coal gasifier is first dehydrated and compressed to specified separation pressure and subsequently cooled to the specified separation temperature in a heat exchanger network consisting of process-to-process heat exchangers as well as utility heat exchangers. From the main separator, the liquid CO_2 product is further purified in a secondary flash separator, pumped and heated to transport state. The hydrogen-rich fuel product is heated and expanded sequentially until the gas turbine combustor feed state is reached. The process is not autothermal and requires auxiliary refrigeration, which can be provided by standard vapour-compression cascade refrigeration units. In this work propane (C3) and ethane (C2) are used as refrigerants although more sophisticated and efficient cycle designs are possible.
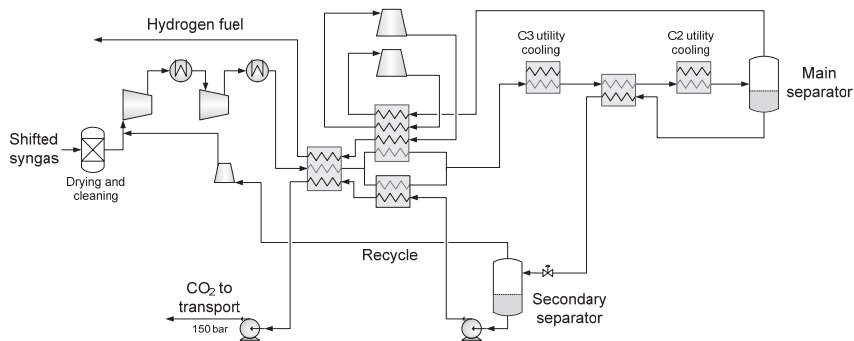

Fig. 1. Process flow diagram for the case of high-pressure pipe CO_2.

In addition to producing high-pressure CO_2 for pipeline transport, the process scheme can be modified to producing liquid CO_2 for ship transport by extracting the liquid product directly from the secondary flash separation unit. The process flow diagram for this option is shown in Fig. 2. In an early phase of CCS deployment where seaborne bulk transport of liquid CO_2 from source to sink may be required in the case of absent pipeline infrastructure, the low-temperature capture unit producing liquid CO_2 can provide a flexible solution, including a real option for a smooth future modification of the capture unit enabling production of high-pressure CO_2 for pipeline transport. The process scheme in Fig. 2 can in principle be retrofitted to the Fig. 1 scheme by adding the CO_2 pumps and reconfiguring a few heat exchangers.
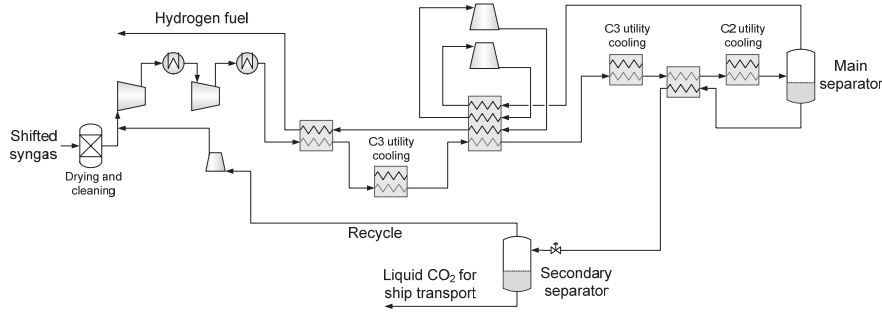
Fig. 2. Process flow diagram for the case of liquid CO_2 for ship transport.

2.2. Technical methodology

The syngas composition used in this work is given in Table 1. Details of the process as well as parameters and assumptions are given in Berstad et al. [2]. In order to obtain a high condensation ratio, and thus CCR, a sufficiently high pressure level is required. Additionally, the separation should be carried out at lowest practical temperature. Operation fully avoiding any solidification of CO_2 is assumed and consequently, the minimum separation temperature is governed by the CO_2 freezing point of the gas mixture at any given separation pressure – between 217 K (-56°C) and 220 K (-53°C) in this work, depending on desired temperature safety margin. Experimental vapour–liquid equilibrium (VLE) data for the H_2/CO_2 system in the temperature range 220–290 K is available from [3]. Based on the experimental VLE data, the obtainable CCR can be estimated, and the CCR for separation at 220 K is plotted as function of pressure in Fig. 3. As can be observed, approximately 85% CCR can be targeted at around 100 bar separation pressure. In this pressure range the discrepancy between results for experimental data and Peng–Robinson equation of state (EOS) is low, indicating that this EOS can be used for predicting CCR.

Steady-state simulations of the low-temperature CO_2 capture process were performed in Aspen HYSYS version 8.3 with Peng–Robinson equation of state.

Table 1. Syngas composition after H_2S removal and drying. Temperature and pressure is assumed to be 30°C and 35 bar, respectively.

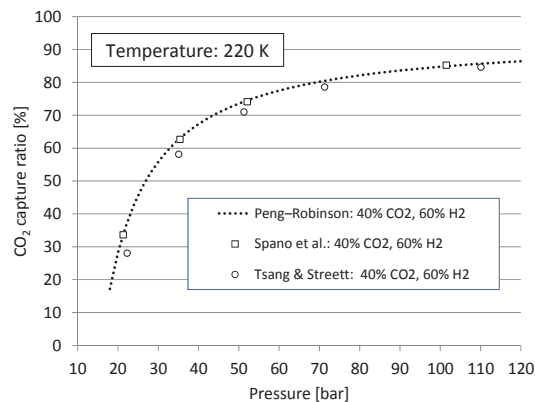| Component	H_2	CO_2	CO	N_2	Ar
Mol fraction	0.5375	0.3804	0.0160	0.0571	0.0090

Fig. 3. Estimated CO_2 capture ratio for phase separation of syngas at 220 K (-53.15°C).

2.3. Cost assessment

   In order to evaluate the potential of the low-temperature concept, the cost of an IGCC with CCS using low-temperature capture is compared to and IGCC without capture, as well as IGCC power plants with CCS using Selexol CO_2 capture for both pipeline and ship transport.

2.3.1. Cost evaluation methodology

   The cost of an IGCC power plant equipped with low-temperature capture is evaluated for an 85% capture rate and compared to the cost of an IGCC power plant with and without Selexol CO_2 capture. This study assumes costs of a "NOAK" (N^{th} Of A Kind) plant to be built at some point in the future when the technology is mature. Investment and operating costs are given in 2008 prices which correspond to the reference costs year in the EBTF report [4] used as reference for costs of the IGCC power plant with Selexol capture.

   The costs are based on the EBTF report [4] for the IGCC power plants and the Selexol capture facility while the low-temperature capture process is evaluated using Aspen Process Economic Analyzer for equipment costs and following the EBTF cost methodology in order to obtain consistent and comparable cost estimates. The EBTF average direct and indirect cost factors, the maintenance costs and utility costs of the IGCC plant with Selexol CO_2 capture are used to evaluate the investment cost of the IGCC with low-temperature CO_2 capture processes in this work.

   The technical characteristics and costs associated with CO_2 conditioning are modelled using the BIGCCS transport modules previously presented and illustrated [5, 6].

2.3.2. Key Performance Indicators

   The Levelized Cost Of Electricity (€/MWh) and the CO_2 avoidance cost (€/t_{CO2}) defined by Rubin et al. [7] are used as key performance indicators to measure the unitary cost of the electricity production and the unitary cost of CCS from an IGCC power plant, respectively, with and without CO_2 capture. The electricity production cost approximates the average discounted electricity price over the project duration that would be required as income to match the net present value of capital and operating costs for the project. The CO_2 avoided cost approximates the average discounted CO_2 tax or quota over the project duration that would be required as income to match the net present value of additional capital and operating costs due to the CCS infrastructure. The calculation of these KPIs considers a real discount rate of 8%[†], 7,500 operating hours per year and an economic lifetime of 25 years [4, 8].

   To evaluate the chain cost per ton of CO_2 avoided in a consistent way, it is necessary to also include the carbon footprint of the full CCS chain. To be consistent with the EBTF methodology, only direct greenhouse gas emissions from burning fuels (coal [4] and shipping fuel [9]) are considered.

3. Results and discussion

   In this section energy and cost results are summarised.

3.1. Energy results

   The absolute power requirement as well as the specific power requirement (energy per unit of CO_2 captured) for the investigated pipeline CO_2 and liquid CO_2 cases, are summarised in Table 2. As can be observed, the specific power requirement is higher for producing liquid CO_2 for ship transport than for high-pressure CO_2 for pipe transport. This is mainly due to the higher requirement for auxiliary refrigeration due to lower degree of process-to-process heat recuperation in the liquid CO_2 case. By energy optimisation of the auxiliary refrigeration cycles, however, a 5–10% reduction in power requirement should be possible for the process generating liquid CO_2.

─────────

[†] This real discount rate of 8 % corresponds to a nominal discount rate around 10% if an inflation rate of 2% is considered.

Table 2. Energy results for the two investigated cases.

Design	Compressors [kJ$_e$/kg$_{CO2}$]	Expanders [kJ$_e$/kg$_{CO2}$]	Auxiliary refrigeration [kJ$_e$/kg$_{CO2}$]	Cooling water pumping [kJ$_e$/kg$_{CO2}$]	Recycle compression [kJ$_e$/kg$_{CO2}$]	CO$_2$ pumping [kJ$_e$/kg$_{CO2}$]	Net specific power [kJ$_e$/kg$_{CO2}$]
Baseline[a]	279	119	67.6	2.55	14.0	18.2	275
Liquid CO$_2$[b]	276	110	125	3.6	13.9	—	321

[a] Producing gaseous CO$_2$ at 150 bar for pipeline transport.
[b] Producing liquid CO$_2$ at -56°C and 7.8 bar for ship transport.

3.2. Cost evaluation results

Based on the electricity output and the plant costs, the cost of electricity (LCOE) is used to compare the five power plant cases without or with capture, and for pipeline or ship transport as shown in Fig. 4. It is worth noting that at this stage neither transport nor storage costs are considered. The evaluation shows that the cost of electricity with low-temperature CO$_2$ capture is 23% more expensive than the plant without capture and is therefore 9% less expensive than the reference plant with Selexol CO$_2$ capture. When looking directly at the cost of capturing CO$_2$, the low-temperature concept for pipeline transport, with a cost of only 22 €/t$_{CO2,avoided}$, is 35% more cost-efficient than the Selexol process.

For the capture process producing liquid CO$_2$ for ship transport, the cost of electricity with low-temperature CO$_2$ capture and internal CO$_2$ liquefaction is only 23% more expensive than the plant without capture. It is therefore 11% less expensive than the reference plant with Selexol CO$_2$ capture and liquefaction by ammonia cycles commonly used in the literature [10, 11]. When considering the CO$_2$ avoidance costs, the evaluation gives a capture cost, including liquefaction, of 23 €$_{2008}$/t$_{CO2,avoided}$. This is 37% more cost-efficient than the Selexol process (38 €$_{2008}$/t$_{CO2,avoided}$). This is primarily due to the already existing cost advantage of the process in the case of a pipeline export, and the extremely energy- and cost-efficient low-temperature internal liquefaction for ship transport compared to liquefaction by ammonia cycles.

The capture costs in the cases including liquefaction of CO$_2$ for ship transport are higher than for the pipeline CO$_2$ cases for both capture technologies. However, these results cannot be compared directly as the transport systems cost are not included in the numbers.
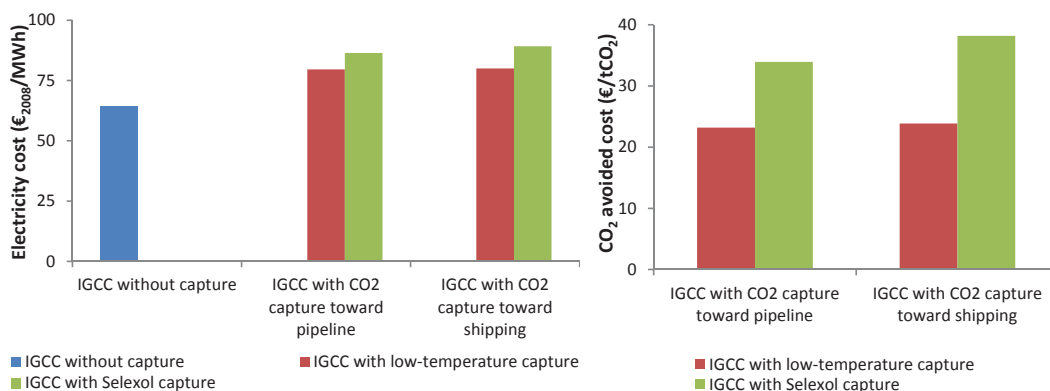

Fig. 4. Electricity cost (a) and CO$_2$ avoided cost (b) of the five IGCC power plant cases with and without CCS.

4. Conclusions and further work

Process designs for low-temperature CO$_2$ capture from IGCC using CO$_2$ liquefaction have been presented. Steady-state simulations and techno-economic analysis have been performed to identify the performance of the

system. The specific power requirement for the base case has been calculated to 275 kJ_e/kg_{CO_2} for a CO_2 capture rate of 85%. The CO_2 product from this configuration is high-pressure gas at 150 bar for pipeline transport. Another process configuration was designed and simulated for producing cold liquid CO_2 at around 8 bar for ship transport. The specific power requirement for this configuration is calculated to 321 kJ_e/kg_{CO2}, which is roughly 17% higher than for the base case. However, this figure should be possible to improve by opimisation of the auxiliary refrigeration cycles.

An economic benchmarking of the two processes has been performed with Selexol process as the reference case. The levelized cost of electricity for the two CO_2 liquefaction processes are 9% and 11% lower than the corresponding Selexol cases while the CO_2 capture cost are 35% and 37% lower than the corresponding Selexol cases.

The results presented show that CO_2 capture by liquefaction in an IGCC process is both cost and energy efficient compared to the Selexol process. A systematic CCS chain approach, including capture, transport and storage, should therefore be performed in order to evaluate the full cost of this concept and if the low-temperature concept lead to additional synergy opportunities with the transport technologies.

Acknowledgements

References

[1] Berstad D, Anantharaman R, Nekså P. Low-temperature CO_2 capture technologies – Applications and potential. International Journal of Refrigeration. 2013;36:1403-16.
[2] Berstad D, Anantharaman R, Nekså P. Low-temperature CCS from an IGCC Power Plant and Comparison with Physical Solvents. Energy Procedia. 2013;37:2204-11.
[3] Tsang CY, Street WB. Phase equilibria in the H2/CO2 system at temperatures from 220 to 290 K and pressures to 172 MPa. Chemical Engineering Science. 1981;36:993-1000.
[4] Anantharaman R, Bolland O, Booth N, Dorst EV, Ekstrom C, Franco F, et al. D1.4.3 European best prectice guidelines for assessment of CO_2 capture technologies. DECARBit Project; 2011.
[5] Roussanaly S, Jakobsen JP, Hognes EH, Brunsvold AL. Benchmarking of CO_2 transport technologies: Part I—Onshore pipeline and shipping between two onshore areas. International Journal of Greenhouse Gas Control. 2013;19C.
[6] Roussanaly S, Brunsvold AL, Hognes ES. Benchmarking of CO_2 transport technologies: Part II – Offshore pipeline and shipping to an offshore site. International Journal of Greenhouse Gas Control. 2014;28:283-99.
[7] Rubin E, Booras G, davison J, Ekstrom C, Matuszewski M, McCoy ST, et al. Toward a common method of the cost estimation for CO2 capture and storage at fossil fuel power plants. Global CCS institute; 2013.
[8] Roussanaly S, Brunsvold AL, Hognes ES, Jakobsen JP, Zhang X. Integrated Techno-economic and Environmental Assessment of an Amine-based Capture. Energy Procedia. 2013;37:2453-61.
[9] EcoInvent. Ecoinvent life cycle inventory database v2.2: The ecoinvent Centre. Swiss center for life cycle inventorys. Available from: www.ecoinvent.org; 2012.
[10] Alabdulkarem A, Hwang Y, Radermacher R. Development of CO_2 liquefaction cycles for CO2 sequestration. Applied Thermal Engineering. 2012;33-34:144-56.
[11] Roussanaly S, Bureau-Cauchois G, Husebye J. Costs benchmark of CO_2 transport technologies for a group of various size industries. International Journal of Greenhouse Gas control. 2013;12C:341–50.