

Article

More Energy-Efficient CO₂ Capture from IGCC GE Flue Gases

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Abstract: Carbon dioxide (CO₂) emissions are one of the main reasons for the increase in greenhouse gasses in the earth's atmosphere and carbon capture and sequestration (CCS) is known as an effective method to reduce CO₂ emissions on a larger scale, such as for fossil energy utilization systems. In this paper, the feasibility of capturing CO₂ using cryogenic liquefaction and improving the capture rate by expansion will be discussed. The main aim was to design an energy-saving scheme for an IGCC (integrated gasification combined cycle) power plant with CO₂ cryogenic liquefaction capture. The experimental results provided by the authors, using the feed gas specification of a 740 MW IGCC General Electric (GE) combustion power plant, demonstrated that using an orifice for further expanding the vent gas after cryogenic capture from 57 bar to 24 bar gave an experimentally observed capture rate up to 65%. The energy-saving scheme can improve the overall CO₂ capture rate, and hence save energy. The capture process has also been simulated using Aspen HYSYS simulation software to evaluate its energy penalty. The results show that a 92% overall capture rate can be achieved by using an orifice.

Keywords: IGCC GE process gases; expansion; cryogenic separation; process system modeling

1. Introduction

CO₂ capture and storage is an effective method for the substantial reduction of CO₂ emissions produced from the utilization of fossil fuels such as in power generation. Primary approaches aimed at CO₂ separation from flue gas currently available are: absorption (including chemical and physical absorption), adsorption, use of membranes, chemical looping and cryogenic processes [1]. CO₂ capture by cryogenics is a unique separation method which requires a low temperature to cool, condense and purify CO₂ from a gas stream. This technique is a vapor liquid separation (CO₂-rich liquid phase) namely cryogenic condensation, vapor solid separation namely the hydrate precipitation, and a combination of both such as in CO₂ slurry separation. For hydrate precipitation, clathrate hydrates are crystalline solid compounds in which gaseous CO₂ molecules (guest) are trapped in water cavities (host) by hydrogen bonds under appropriate thermodynamic conditions (such as high pressure and/or low temperature). Hydrate-based CO₂ capture (HBCC) has been receiving attention due to its moderate operating temperature range, the ease of recycling for aqueous solution containing additives, the capability for continuous operation which allows large-scale treatment and its unique separation mechanism.

For cryogenic condensation, flue gas under high pressure is cooled to sub-zero temperatures which allow only CO₂ to condense. The process of liquefying CO₂ can provide two types of product conditions: one is liquid CO₂ for pipeline transport, and the other is liquid CO₂ for ship

transport. Many researchers have conducted experiments based on this method and have accomplished significant results. For example, Zhang et al. studied a liquefied natural gas-fueled quasi-combined power plant using supercritical CO₂ Rankine-like and Brayton cycles [2–4]. Zanganeh et al. introduced and discussed the challenges, development stages, and commissioning of a pilot-scale CO₂ capture and compression unit, which can separate CO₂ as a liquid phase from the flue gas of oxy-fuel combustion [5]. Amann et al. evaluated the technical performance of a natural gas combined cycle converted for oxy-fuel combustion using a O₂/CO₂ cycle with a cryogenic CO₂ recovery process [6]. Hart et al. developed a cryogenic CO₂ separation process, in which CO₂ is cooled into a three-phase mixture and the solid CO₂ is melted by a heater, and then separated with the liquid [7].

CO₂ capture by cryogenic condensation and expansion possesses several economic and technological advantages such as:

- (1) Liquid CO₂ is directly produced at a relatively low pressure, avoiding huge energy consumption in compressing gaseous CO₂ to a very high pressure to reach the supercritical state.
- (2) The cryogenic separation method is based on relatively mature industrial processes, such as compression and refrigeration, indicating easy expansion to industrial-scale utilization.
- (3) In cryogenic processes, CO₂ can be handled in the liquid phase, making it easy to pump to the high pressures required for pipeline transportation and sequestration with a minimum energy requirement.

However, to achieve CO₂ condensation, the mixture requires a reduction to a very low temperature, which requires substantial energy consumption for refrigeration, leading to a high energy penalty for CO₂ capture, and it requires CO₂ solidification at low temperatures, which causes several operational problems, such as the blockage of pipes or heaters. Therefore, the key point in improving the cryogenic method lies in how to achieve CO₂ liquefaction at relatively higher temperatures above -55 °C.

1.1. Improvement of the Cryogenic Separation Method

Figure 1 shows the three-phase diagram of CO₂. The liquefaction temperature of CO₂ is determined by its pressure. The increase in the CO₂ pressure enhances its liquefaction temperature, which can lower the energy consumption in the refrigeration and prevent it from freezing. Thus, the increase in CO₂ pressure can be an improvement technique for the cryogenic separation method. In general, two factors can directly affect the CO₂ pressure in a gas mixture: one is the CO₂ concentration, and the other is the total pressure of the gas mixture.

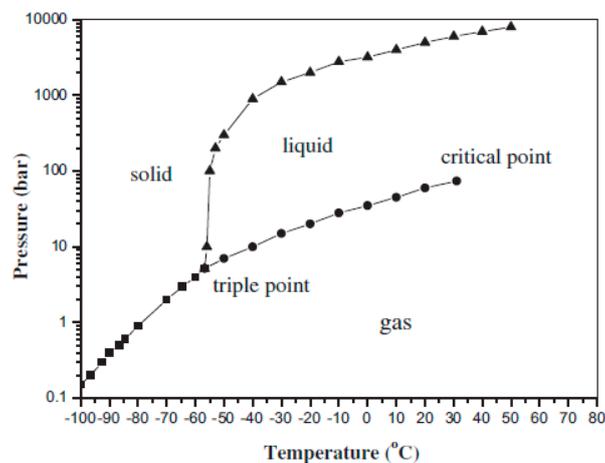


Figure 1. Three-phase diagram of CO₂ [8].

Xu et al. conducted the experiments that gave all the data as shown in Figures 2–4 [8]. Figure 2 shows that increasing the CO₂ concentration in the mixture to 80 mol % or higher before initiating the separation process prevents CO₂ solidification and decreases cryogenic energy consumption. Effectively increasing the CO₂ concentration in the gas mixture significantly improves the performance of the cryogenic separation method.

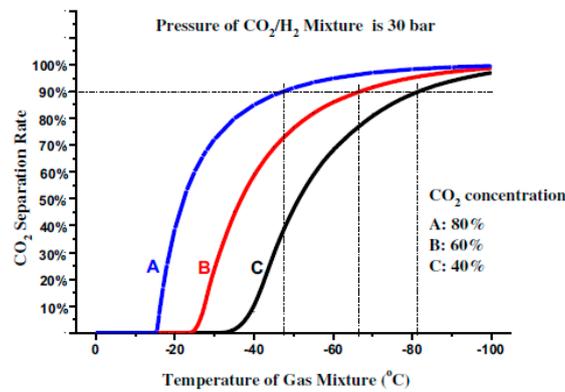


Figure 2. Variation in CO₂ concentration and separation ratio of CO₂/H₂ mixture with temperature [8].

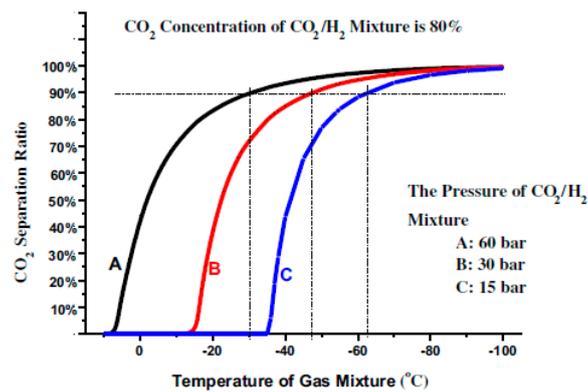


Figure 3. Variation in pressure and separation ratio of CO₂/H₂ mixture with temperature [8].

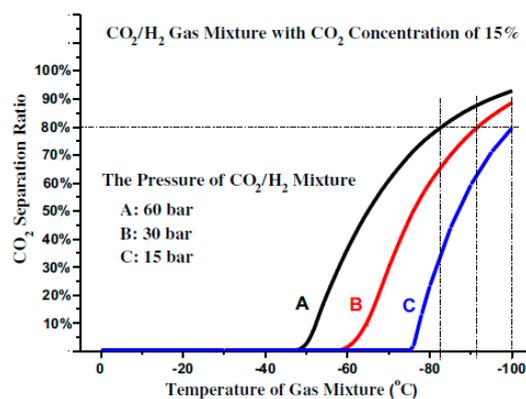


Figure 4. Variation in pressure and separation ratio of CO₂/H₂ mixture with temperature at 15 mol % CO₂ concentration [8].

The relationship between the CO₂ separation ratio and the temperature of the CO₂/H₂ mixture under different pressures, at a CO₂ concentration of 80 mol %, is shown in Figure 3. Under the mixture pressures of 15, 30, and 60 bar, the temperature should be lowered to about −63 °C, −48 °C,

and 30 °C, respectively, to separate 90% of the CO₂ from the mixture. In other words, enhancing the mixture pressure can also effectively increase the separation temperature of CO₂, thereby improving the performance of the CO₂ cryogenic method.

If the concentration of the CO₂ in the gas mixture is very low, the increase in the CO₂ liquefaction temperature will be constrained by the enhancement of the gas mixture pressure as shown in Figure 4.

1.2. CO₂ Capture from IGCC GE Process Gases by Expansion after Cryogenic Separation

Cryogenics and hydrate techniques can be used for CO₂ capture from process streams containing mainly CO₂ and hydrogen. The baseline approach for the cryogenic process is to cool down the gas under pressure to temperatures where CO₂ can be condensed and removed in liquid form at an elevated pressure while the hydrogen stays in the gas phase. Under the project funded by The Cooperative Research Centre for Greenhouse Gas Technologies (CO₂CRC), a laboratory pilot plant was designed in 2007 and built in 2008 at the Clean Gas Technology Australia of Curtin University to capture CO₂ by a combined cryogenic-hydrate method. Experimental results using this method have verified that 79% of CO₂ can be captured by cryogenics, plus another 11% by hydrates using a newly identified promoter for the IGCC GE process gas [9]. Another more efficient method of capturing 11% CO₂ after the cryogenic section was proposed by expanding the depleted gas overhead of the separator to 24 bar, which will be described in this paper, including an energy-saving scheme.

This work is to show and discuss the experimental results obtained from a series of tests carried out using the laboratory pilot plant in order to achieve the CO₂ capture rates by expansion after the cryogenics section using nozzles with various sizes of orifices. The IGCC GE gas compositions, as shown in Table 1 below, were originally prepared according to the design report [10].

Table 1. The IGCC GE gas composition and conditions [10].

Component	Process stream composition (mol%)	
	IGCC Texaco	Experimental values
H ₂	55.15	58.13
CO	2.85	-
CO ₂	40.30	40.40
N ₂	0.68	0.68
CH ₄	0.02	-
H ₂ S	0.22	-
Ar	0.79	0.79
Inlet pressure	57 bar	57 bar

Notes: (1) CO, CH₄ and H₂S components were omitted for this experimentation due to safety considerations since the experimentation was done in a confined space environment (laboratory). This safety consideration, however, does not affect the industry in applying the energy-saving scheme as presented in this work, as the IGCC GE gas is originally composed of CO, CH₄ and H₂S and these are typically removed by a CO₂ capture process. (2) For IGCC GE gases: MW = 19.45 kg/kg mole, total flow rate for the power plant = 37205 kg mol/hr; (3) IGCC GE gases overhead of the separator after the cryogenics section (depleted gas) contain 14.7 mol % CO₂ by simulation (design) and 12 mol % CO₂ by experimental work [11].

The objective of this work is to perform a modification of the existing laboratory pilot plant so that the energy available in the system at high pressure can be used to capture another 11% of CO₂ from the outlet gas of the cryogenics section containing 12–15 mol % CO₂ to achieve an overall CO₂ capture rate of 90 mol %. An energy-saving scheme was also developed for the combined capture/recovery scheme using HYSYS simulation software for the IGCC GE power plant with a gross power generation of 740 MW [12]. The mechanism and potential of the energy-saving scheme in other cryogenic CO₂ capture processes for a coal-based hydrogen production process were also presented by Xu et al. [8]. They proposed that the total energy penalty in the CO₂ capture process via the novel CO₂ cryogenic liquefaction and separation method for a gas mixture composed of 80% CO₂ and 20% H₂ is 35.81 MW. However, this cryogenic energy-saving scheme will not be discussed in this work.

2. Materials and Methods

2.1. Flow Scheme, Equipment and Material Used

Figure 5 shows the flow scheme for capture of CO₂ from the IGCC GE process gases by expansion that was investigated in this work.

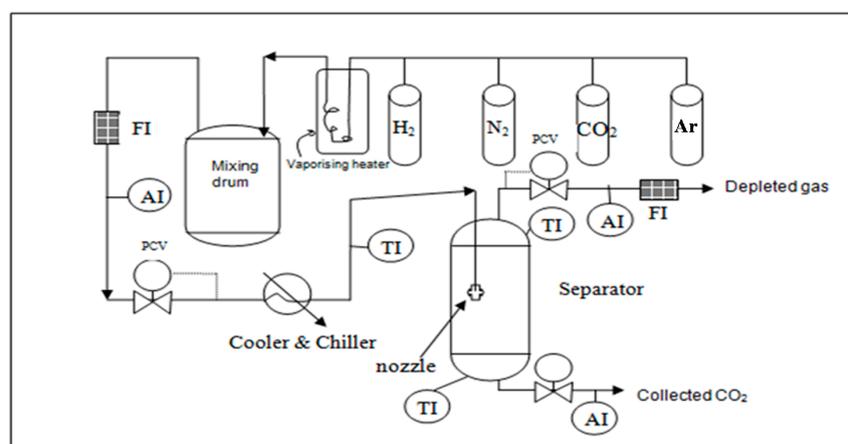


Figure 5. Laboratory pilot plant schematic (cryogenic condensation).

First, the nozzle with a particular size of orifice was installed at the inlet line of the separator in order that the inlet pressure of the separator could be reduced to 24 bar or lower as designed by CO₂CRC. In each experiment, a nozzle with one orifice size was selected from a range of 0.1, 0.2, 0.3 and 0.4 mm. The process gases were then prepared using the pressure regulators at the gas cylinders. The mix drum was filled by gas mixed to a composition of the IGCC GE process gases outlet overhead of the separator (after cryogenics section at 57 bar) which equaled 15 mol % CO₂ approximately [11]. The pressure of the gas mixture in the mix drum was filled to about 110 bar thereby ensuring that sufficient gas at the fixed composition was available throughout the experiment. The required composition of the gas mixture (at the IGCC GE process gas composition of 15 mol % CO₂ approximately as shown in Table 3) was verified using a CO₂ analyzer after the gases were mixed in the mixing drum.

The gas mixture was fed to the heat exchanger which was simulated by chilling coils submerged in methanol with dry ice bath. The feed pressure was controlled at an operating pressure (57 bar) and an operating temperature of $-55\text{ }^{\circ}\text{C}$ which was achieved before entry to the separator. At these conditions, the condensed CO₂ was separated from the overhead gas stream. The liquid CO₂ composition was analyzed at the exit. The lean CO₂ overhead gas was also analyzed at the exit from the separator. Flash calculations to estimate effectiveness of cryogenic separation of CO₂ were performed in HYSYS process simulation environment using Peng-Robinson equation of state.

The temperature of each stream was measured by inline sensors and monitored using a 6 channel Shimaden controller with an accuracy of $\pm 0.1\text{ }^{\circ}\text{C}$. The cold methanol-ice bath used as a heat exchanger was monitored by a Deluxe Dual Input Thermometer with an accuracy $\pm 0.1\text{ }^{\circ}\text{C}$. The inlet pressure was controlled by Swagelok high pressure regulators. Swagelok pressure transmitters with 0–120 bar range connected to a Shimaden controller were used for monitoring pressure with an accuracy of ± 0.1 bar. An Elster RVG G16 flow meter was used to measure and control gas flow with an accuracy of ± 1 . A gas alarm CO₂ analyzer (online MSR Polygard (R) analyzer) was installed on the venting line after the flow meter. All measurements were carried out after reaching steady flow and operating conditions for 15 min. All vessels used for experimental set-up were made of carbon steel (CS) and stainless steel (SS) of vertical orientation having the design pressure of 110 bar. Dimension of the vessels are given in Table 2.

Table 2. Dimensions of the vessels used.

	Mixing drum	Separator
ID (in.)	7	2
Length (in.)	41	15
Material	CS	SS

2.2. Operating Procedure

Gas mixtures with a composition of 15 mol % CO₂ approximately were prepared in a separate pressure vessel referred to as the mixing drum (shown in Figure 5). Mixtures were prepared on a pressure basis using industrial grade gases supplied by BOC Australia. CO₂ was passed through a hot water bath in order to ensure that all of it was in the gas phase. Pressure of each gas to be added was calculated in terms of the desired composition taking into account compressibility factor. Hydrogen, argon, and nitrogen were first mixed together. After that, CO₂ was added slowly allowing for faster stabilization to the total pressure of 110 bar. The mixture was left for 2–3 h to reach equilibrium.

The actual compositions of the gas mixtures were verified using the gas alarm CO₂ analyzer by taking two samples, from top and bottom respectively, of the prepared mixture were taken to ensure even distribution of CO₂ concentration along the mixing drum. Dry ice was constantly added into the methanol chilling bath, and temperature was controlled at about –55 °C. Hydrogen from a gas cylinder was passed through the chilling coils and separator at a flow rate of about 0.1 m³/h at supplying pressure of 57 bar in order to bring the temperature of the entire system to the desired level. The separator; with two temperature indicators located at the inlet and bottom lines, was gradually chilled to below –50 °C by pouring liquid nitrogen on the outer shell of the vessel inside the insulation. Once the required equilibrium temperature (–55 °C) for the separator was reached, the hydrogen feed was immediately switched to the synthetic gas from the mixing drum at 57 bar. The pressure of the separator was controlled to be at about 24 bar or lower by adjusting the regulator valve of the vent system. Liquid CO₂ bottom product was frequently drained to a collecting vessel in order to prevent an overflow of CO₂ from the inner separator. The concentration of CO₂ in the liquid was analyzed after the equilibrium condition was reached. Lean CO₂ overhead gas was passed through the analyzer at near atmospheric pressure (vent system). The actual equilibrium condition in the separator could then be observed when a stable CO₂ composition was reached; as measured by the CO₂ analyzer, was obtained at the overhead gas off the separator. The inlet separator temperature, at which the CO₂ composition became stable was recorded. This value was used in a HYSYS simulation to estimate the amount of CO₂ captured at the specified pressure and temperature of the separator.

The quantity of mixture prepared in the mixing drum was sufficient to run the cryogenic experiment for 20–30 min of each run.

2.3. Energy-Saving scheme for the Capture and Recovery Scheme of CO₂ from a 740 MW IGCC GE Combustion Power plant

The HYSYS software was used to simulate the IGCC GE power plant. Consider an energy saving scheme for CO₂ capture by cryogenics for the IGCC GE power plant having a gross power generation of 740 MW as shown in Figure 6.

The scheme as shown in Figure 6 is based on the feed gas chilling at 57 bar to –55 °C (subject to the conditions attained from the experiments) causing separation and capture of about 75%–80% of the CO₂ contained in the feed. The captured CO₂ in the form of liquid is then pumped by Pump 1 at –55 °C to 110 bar causing the temperature to rise to about –50 °C which is used to chill the feed gas in the recovery exchangers E5, E3 and E1.

The uncondensed part of the gas composing mainly of hydrogen is then expanded to a lower pressure (supposedly 24 bar) using the nozzles installed at the inlet line of the Separator B which

causes solidification of more CO₂ at the bottom and chilling to a temperature below $-55\text{ }^{\circ}\text{C}$ which could be down to $-83\text{ }^{\circ}\text{C}$ approximately. The vent gas mainly comprising of hydrogen is used to chill the feed in the recovery exchangers E6, E4 and E2. The solid CO₂ is then heated at the bottom of Separation B to a temperature of $-55\text{ }^{\circ}\text{C}$ by an exchanger providing heat load Q. Liquid CO₂ is then pumped by Pump 2 to 110 bar causing a temperature rise to about $-50\text{ }^{\circ}\text{C}$. Liquid CO₂ from Pumps 1 and 2 are combined and used to chill the feed gas in exchangers E5, E3 and E1.

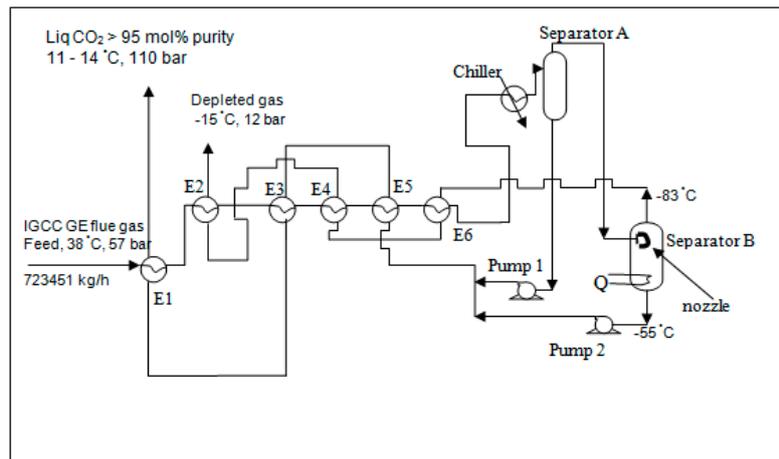


Figure 6. Energy saving scheme for CO₂ capture from a 740 MW IGCC GE power plant.

The total energy consumption for achieving liquid CO₂ at about $11\text{ }^{\circ}\text{C}$ and 110 bar is the sum of: Chiller + Pump 1 + Pump 2 + heat supplied at the bottom of Separator B (represented by Q) as shown in Table 4.

Conditions of data used in the simulations are as follows:

- The operating temperature of Separator B = the temperature measured at the inlet of the separator in the experiments.
- The operating pressure of Separator B = the pressure measured at the vent system of the separator in the experiments.
- In the actual experiments, nozzles were used as described earlier in this paper to reduce the pressure in Separator B. However, an expander was positioned at the inlet Separator B in each simulation for the same purpose in using nozzles due to no nozzles installation scheme available in HYSYS simulations.
- The efficiency of the chiller was assumed to be 75%.

The simulation data for each operating condition used and the equivalent power consumption for each process unit in Figure 6 are shown in Table 4.

2.4. The Capture Scheme of CO₂ from a 740 MW IGCC GE Combustion Power Plant without the Energy-Saving Scheme

In comparison with Section 2.3 where the energy saving scheme was performed, the typical 740 MW IGCC GE combustion power plant without process integration was modeled and simulated as shown in Figure 7.

The energy required for production of CO₂ in a dense phase (110 bar, $40\text{ }^{\circ}\text{C}$) would be

Compression (single stage): 122.75 MW;

Cooling from $630\text{ }^{\circ}\text{C}$ to $40\text{ }^{\circ}\text{C}$: 159.31 MW;

Total energy = compression + cooling = 122.75 MW + 159.31 MW = 282.06 MW = 38% of 740 MW

Therefore the total energy to be consumed for the compression and cooling stages would be 38% of the total power generated. The simulation data for operating compressor and cooler in association with the equivalent power consumption for the process in Figure 7 are also shown in Table 4.

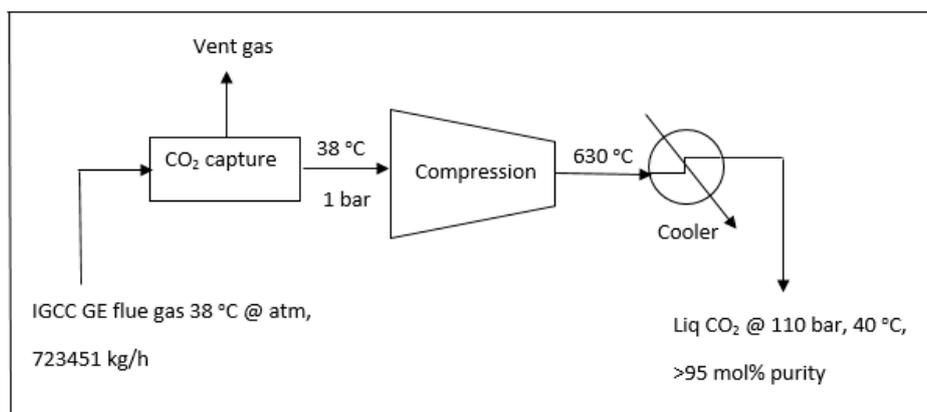


Figure 7. Typical flow scheme for CO₂ capture from a 740 MW IGCC GE power plant.

3. Results and Discussion

As shown in Table 3, the operating pressures at the outlet separator were not obviously related to the nozzle sizes. It was due to the fact that there was a valve installed at the vent system of the separator. Apart from the nozzles, this valve played an important role in the pressure control of the system.

Table 3. Experimental results and material balance for each equilibrium condition tested.

Orifice Nozzle (mm)	Operating pressure (bar)		Separator Inlet Temp. (°C)	Feed flow rate		Composition of CO ₂ (mol %)			CO ₂ capture rate (%)
	Inlet Separator	Outlet Separator		(m ³ /hr)	(cc/min)	Feed	Depleted gas	CO ₂ product	
0.1	57	10	-55	0.104	1725	14	6	92	61
0.2	57	11	-54	0.104	1738	13	5	94	65
0.3	57	17	-55	0.105	1756	15	10	95	37
0.3	57	13	-54	0.101	1689	15	9	93	44
0.4	57	16	-53	0.107	1781	15	12	97	23

Experimentally, the IGCC GE gas mixtures with up to 15 mol % CO₂ were tested using four different sizes of nozzles (0.1, 0.2, 0.3 and 0.4 mm) at the same operating conditions of pressure (57 bar), temperature (-55 °C) and feed flow rate (0.1 m³/hr). Using the orifice nozzle sized 0.1 mm resulted in up to 61% of CO₂ captured. The overhead stream contained 6 mol % CO₂ and the liquefied CO₂ obtained in this case was characterized by a 92 mol % purity on average. As can be observed, there was no confirmed impact of the orifice nozzle size on the concentration of CO₂ in the liquid phase. With the various sizes of the orifice nozzles, between 23% and 65% of CO₂ was captured from the depleted gas. The nozzle sized 0.3 mm was tested twice by varying the outlet separator pressures. The liquid bottom product contained between 92 and 97 mol % CO₂. The best CO₂ capture rate, i.e., 65%, was achieved with the use of the 0.2 mm nozzle.

As indicated before, all the experimental results were simulated using HYSYS software for the energy-saving scheme for the capture and recovery of CO₂ from a 740 MW IGCC GE combustion power plant. The details of the flowsheet and the parameters for the simulation are presented above in Section 2.3. The simulated results are shown in Table 4 and described as follows.

Table 4. Simulated results of the energy required for a typical 740 MW IGCC GE power plant with and without the energy-saving scheme and the experimental results of the CO₂ capture rate obtained from the expansion.

Nozzle size (mm)	Separator B Operating pressure (bar)	Separator B Inlet Temp (°C)	Energy Required (MW)							% of the total power generated (740 MW)	CO ₂ capture rate by expansion (%)	Total CO ₂ capture rate (%)
			Compressor	Cooler	Chiller	Pump 1	Pump 2	Q	Total			
0.1	10	−55	Not required		55.75	0.91	0.21	1.20	58.07	7.8	12.8	91.8
0.2	11	−54	Not required		54.27	0.90	0.20	1.10	56.47	7.6	13.6	92.6
0.3	13	−54	Not required		54.27	0.90	0.18	0.92	56.27	7.6	9.3	88.3
0.4	16	−53	Not required		52.77	0.89	0.16	0.70	54.52	7.4	4.8	83.8
0.3	14	−55	Not required		55.75	0.91	0.15	0.67	57.48	7.8	7.8	86.8
N/A	Typical scheme		122.75	159.3		Not required			282.1	38.0	N/A	90.0

For each condition, the total CO₂ capture rate = (CO₂ capture rate by cryogenics using IGCC GE flue gases as the feed to Separator A = 79% [9]) + (CO₂ capture rate by expansion using the depleted gas off Separator A as the feed to Separator B, which was shown in Table 3). For example: by using the nozzle size of 0.2 mm., Table 3 shows the CO₂ capture rate by expansion (using the depleted gas off Separator A as the feed to Separator B) = 65%. This means that the expansion section could capture 65% of the 21% of the remaining CO₂ after the cryogenic section (as 79% was already captured in the cryogenic section). Therefore, 65% of the remaining 21% CO₂ = 0.65 × 0.21 = 0.1365 = 13.6%. The total CO₂ capture rate for the entire process (cryogenics + expansion) is thus 79% + 13.6% = 92.6%.

With the energy-saving scheme, as shown in Figure 6, the feed gases were gradually cooled down by a series of heat exchangers (E1 to E6) and finally chilled to $-55\text{ }^{\circ}\text{C}$ by a chiller before entering separator A. The overhead stream of separator A flowed to separator B with a nozzle installed inside for the purpose of expansion. Additional heat supplied was also installed to the bottom of separator B due to the ultra-cooled bottom liquid. The liquid products off the bottom of both separators were transferred through E5, E3 and E1 by pumps 1 and 2 while the overhead off separator B flowed through E6, E4 and E2 to chill the feed stream. The liquid CO_2 product was obtained at a more than 95 mol % CO_2 concentration at about 110 bar and $11\text{--}14\text{ }^{\circ}\text{C}$.

Based on the simulated results, the CO_2 capture rates at the expansion section were likely to drop along with the increasing size of the nozzles. The trial with the 0.3 mm nozzle was done twice, first at 13 bar giving the CO_2 capture rate of 9.3%, while the second was done at 17 bar giving a lower CO_2 capture rate at 7.8%. The operating pressure of 11 bar at $-54\text{ }^{\circ}\text{C}$ using a 0.2 mm nozzle was obviously the best condition for the purpose of CO_2 capture as it gave up to 13.6% of CO_2 captured by expansion, resulting in a total 92.6% capture rate for the entire cryogenic section. In terms of the total energy required for the process, using 0.1 and 0.2 mm nozzles could capture more than 90% of CO_2 in the total cryogenic section. However, the process using the 0.2 mm nozzle required slightly less energy, 7.6 MW. The simulated typical scheme required a total energy of 282 MW which was the highest energy required compared to all the other schemes.

Performance Comparison with Literature

According to the literature review on other CO_2 capture processes, it is obvious that dual-stage Selexol processes have been recognized as the most conventional processes for recovering H_2S and CO_2 simultaneously [13]. For the design of acid gas removal processes using Selexol solvents, Kohl and Nielsen [14] exhibited a simple two-stage Selexol process composed of one set of an absorber and steam stripper for H_2S removal followed by another set of an absorber and air stripper for CO_2 removal. The simple two-stage Selexol process was simulated by Robinson and Luyben [15]. Bhattacharyya et al. [16] implemented a comprehensive process simulation of an entire IGCC power plant integrated with a dual-stage Selexol process for a 90% overall carbon capture efficiency with the performance as shown in Table 5.

Table 5. Comparison of the energy consumption of the simulated dual-stage Selexol process and the simulated cryogenic energy-saving scheme.

Case	Integrated Dual-Stage Selexol unit	Cryogenic CO_2 Capture by Expansion with the Energy Saving Scheme (with 0.2 mm nozzle)
Carbon capture efficiency (%)	90.0	92.6 *
CO_2 product purity (mol%)	97.2	>95
H_2S stripper duty (MW_{th})	14.6	-
CO_2 compression power (MW_e)	32.1	-
Total auxiliary power consumption in dual-stage Selexol units (MW_e)	20.0	-
Power for chiller + pumps + coiled heater (MW)	-	56.5
Total energy consumption (MW)	66.7	56.5

Note: * experimental result.

Table 5 shows the energy consumption of the process, as extracted from Kapetaki et al. [13], as well as a comparison between the energy consumption of the dual-stage Selexol process and that of the cryogenic energy-saving scheme as described earlier in Table 4.

As shown in Table 5, the cryogenic CO_2 separation (with an expansion section) is apparently a competitive CO_2 capture technology for IGCC plants due to less energy consumption for the CO_2 capture process compared to other CO_2 capture processes such as the dual-stage Selexol process.

4. Conclusions

For the IGCC GE process gas scheme, an efficient method for capturing 11% CO₂ after the cryogenic section was proposed by expanding the depleted gas overhead of the separator to 24 bar (or less) using nozzles with various sizes of orifices (0.1–0.4 mm) at low temperatures (−55 °C). The experimental results obtained using the lab pilot plant showed that the best CO₂ capture rate (65%) was achieved with the use of a 0.2-mm-orifice nozzle, resulting in the overall CO₂ capture rate obtained by cryogenics and expansion at 92.6%.

Using the HYSYS software to simulate the typical flow schemes and the energy-saving scheme at 10, 11, 13, 16 and 17 bar showed that the energy-saving scheme consumed between 54 and 58 MW. The equivalent power consumption for the energy-saving scheme varied between 7.4%–7.8% of the total power generated at 740 MW.

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Conflicts of Interest: The authors declare no conflict of interest.

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