Evaluation of the Performance and Economic Viability of a Novel Low Temperature Carbon Capture Process

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Abstract

A novel Advanced Cryogenic Carbon Capture (A3C) process is being developed due to its potential to achieve high CO_2 capture efficiencies using low cost but high intensity heat transfer to deliver a much reduced energy consumption and process equipment size and cost. These characteristics, along with the absence of process chemicals, offer the potential for application across a range of sectors. This work presents a techno-economic evaluation for applications ranging from 3% to 35% CO_2 content.
Keywords: low temperature; carbon capture; anti-sublimation; cryogenic separation

1. Introduction

Reduction of energy penalty and costs of existing or emerging CCS technologies are driving current advances. Amongst these technologies, low temperature CO_2 separation relies on phase change, separating the CO_2 from the gas as liquid or solid, without chemical absorbents. However, low temperature separation is often seen as an energy-intensive choice due to the high cooling duty required [1]. Desublimation processes separate CO_2 as a solid frost which is then warmed to sublimation conditions. They can operate at nearly atmospheric pressure, achieve high CO_2 capture ratios while delivering high-purity CO_2 streams with integral dehydration to ppb levels, avoiding product drying stages [2-4]. Proper integration with low-cost cold sources is key for their competitiveness [1].

Few research groups have investigated desublimation for post-combustion CO_2 capture [1-5]. Clodic et al. studied CO_2 frost formed on the surface of a heat exchanger [2]. However, this hinders heat transfer, reducing process efficiency [3]. Packed beds processes using periodic switching between CO_2 capture, sublimation and bed cooling conditions [1, 3], are complex and suffer adverse switching losses. The A3C process described in this paper overcomes these limitations by using a moving bed of metallic beads as heat transfer medium and frost capture surface. This allows intensive heat transfer and avoids the adverse effects on heat transfer of heavy frost deposition, with a much reduced energy consumption and process equipment size. This work details the process concept and its performance for applications ranging from 3% to 35%vol. CO_2 content in the process streams.

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Nomenclature

A3C   advanced cryogenic carbon capture
CCGT   combined cycle gas turbines
CCS   carbon capture and storage
LCCC   levelised cost of carbon capture
MEA   monoethanolamine

2. The A3C Separation Process

The A3C separation process has two stages, a cooling and drying step, and CO_2 separation by desublimation, as in figure 1. The raw gases, quenched if necessary, are cooled conventionally to 274K to condense most of the water vapour. The residual water content is removed in the cold end of a circulating packed bed of small metallic beads, moving in counterflow to the gas. By cooling to about 190K, the water content is reduced to below 50ppb. The ice bearing bed material is carried out of the raw gas stream, warmed slightly and transferred to a section where it moves in counterflow to the cold lean gas leaving the core process. The warm bed is then cooled and dried by the dry lean gas and returned to the raw gas section at about 185K.
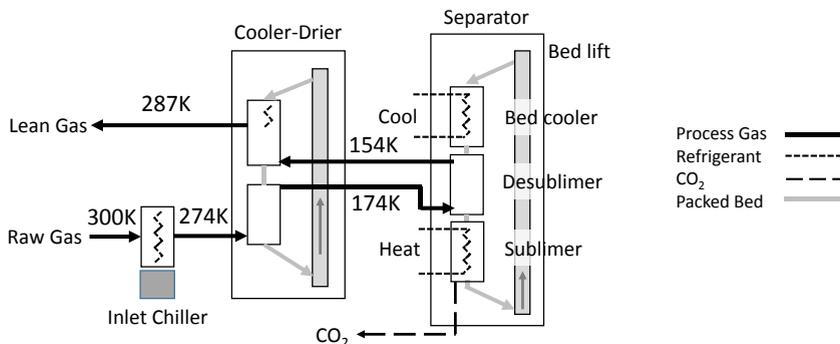

Fig. 1. Outline of the two stages of the A3C carbon dioxide separation process.

The cold dry gas is passed into a second circulating packed bed cascade of similar design to the cooler-drier. Here it flows counter to a colder bed, so that the CO_2 in the gas stream deposits as a frost on the bed material. The lowest gas temperature, around 150K, is chosen to correspond to the CO_2 saturation temperature at the desired residual content. The bed carries the CO_2 frost to the gas inlet end of the bed and through a gas lock into a submerged tube heat exchanger where it is warmed to 195K to recover the CO_2 by sublimation. The bed is then recirculated through a further heat exchanger for cooling to the desired inlet temperature.

2.1. A3C Modelling and analysis

Deriving data for benchmarking the A3C process against an absorption-based carbon capture process required the modelling of the process behavior and energy performance, preliminary engineering of the process equipment and the costing of the equipment. Each of these steps was repeated for the selected applications.

Modelling the thermodynamic behavior of the A3C process in Aspen Plus® presented several challenges. The solid bed was represented by a non-reactive liquid, with direct contact heat exchange represented by indirect heat exchange. The progressive CO_2 desublimation was modelled in small steps following the approach in [6]. The validity of these representations was checked by a spreadsheet model of the core process developed using finite temperature step analysis with detailed CO_2 low temperature properties [7].

Preliminary engineering of the A3C process relied partly on Aspen Plus® functions for conventional elements and spreadsheet design for the submerged tube heat exchangers to assess surface area and physical arrangement. Heat transfer coefficients between bed and gas and between bed and submerged tubes were derived from Ref. [8, 9].

Costs were derived primarily by using the Aspen Plus® cost evaluation functions. Costs for the submerged tube heat exchangers were obtained by adjusting the Aspen costs for equivalent shell and tube implementations. The costs of the direct contact beds and material handling systems drew on references to comparable industrial systems.

2.2. Applications evaluated

The A3C process was compared with a reference amine case for three applications which offered a range of scale and CO_2 content of the process gases as detailed in Table 1. The conventional amine-based reference process is based on Ref. [10] with additional energy recovery, using a 30% wt. of MEA solvent.

Table 1: Key process gas inlet conditions for the three application cases.

	Fired Boiler	CCGT	Biogas upgrading
Gas flow rate (t/h)/(kNm^3/h)	612 / 472	493 / 388	0.83 / 0.77
CO_2 content (vol. %)	12.2%	3.23%	35.0%
Capture rate	90%	90%	94%

2.3. Levelised cost of carbon capture (LCCC)

A cost model based on conventional methods was used to produce a levelised cost of carbon capture excluding the costs of transport and storage or carbon emission credits. The model used costs for heat, power and MEA make-up derived from the process models and fixed costs estimated from capital costs.

3. Benchmarking of the A3C process

The LCCC model compared the A3C process with the reference MEA system for a range of heat and power costs, with a baseline assumption that the steam would otherwise be used to generate electricity in a steam turbine. Table 2 shows that the A3C process can offer a modest advantage over MEA for the larger applications and is radically better for the biogas case due to its lower equivalent power consumption and capital cost.

Table 2: Comparison of MEA with A3C for the different applications at baseline energy cost

	Utility Boiler		CCGT		Biogas Upgrading	
	MEA	A3C	MEA	A3C	MEA	A3C
Heat (MJ/s)	114	0	28.6	0	0.49	0
Power (MW)	12.2	34.4	2.95	17.2	0.06	0.12
Total equivalent (MWe)	38.6	34.4	9.9	17.2	0.19	0.12
Capex (£m)	81.3	82.3	37.5	33.2	4.9	0.65
Opex excl. energy (£m)	6.4	5.1	3.6	1.6	0.69	0.25
LCCC (£/te CO_2)	39.7	34.9	76.8	79.3	395.8	120.9

4. Conclusions

The A3C process has been shown to be feasible for a range of scales and CO_2 concentrations. Techno-economic evaluation of the process shows a modest cost advantage over MEA technology in the larger selected applications and radically lower costs in the smaller application. It should be noted that A3C is an immature technology and while extensive regenerative energy recovery has been used, there are significant opportunities for further improvement and optimization with additional potential for trade-offs between energy consumption and capital cost.

Acknowledgements

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