



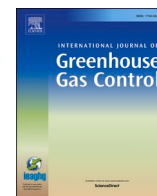
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Capturing or offsetting carbon emissions: techno-economic evaluation of decarbonising industries with CO₂-lean flue gas

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ABSTRACT

This work investigates the technical feasibility and costs of carbon capture from industrial emission sources with CO₂ concentrations of 0.5–4.0%. Three post-combustion capture schemes are considered: monoethanolamine absorption (MEA-ABS); alkali absorption combined with calcium looping (ALK-ABS); and temperature vacuum swing adsorption (TVSA), the latter two commonly investigated for direct air capture. Capture and avoidance costs are compared across industrial sectors, varying the flow rate and CO₂ concentration of the flue gas, and contrasted with the price of offsetting the emissions through purchases of carbon dioxide removal credits at current voluntary-market prices.

The results show that MEA-ABS has the lowest levelized cost of capture (LCOC) for flue gas streams with larger flow rates (>200 t/h) and CO₂ concentrations ≥1%. TVSA presents the lowest LCOC at lower flow rates (especially <100 t/h), while ALK-ABS has the lowest LCOC only within a small window of low CO₂ concentration (0.5–1%) and 100–200 t/h. Accounting for the carbon footprints of the input energy changes this picture: the carbon avoidance costs (assuming a transport and storage cost of 105 \$/tCO₂) reach their lowest values with ALK-ABS, unless heat with a low carbon intensity is available for MEA-ABS. Comparing the cost of carbon avoidance with carbon dioxide removal credits on the voluntary market, it can be more economical for industries with flue gases with CO₂ concentrations <2.6% to offset their emissions through the purchase of such credits than to install any of the post-combustion capture technologies considered here.

1. Introduction

In order to limit global warming to 2°C by the end of this century, both large-scale mitigation of fossil carbon emissions and carbon dioxide removal (CDR) are required (IPCC, 2022). Different mitigation options, including fuel shifting, electrification, efficiency measures, and carbon capture and storage (CCS), have been developed and have reached technological maturity, while CDR technologies are still under development. Presently, a well-established CCS technology is post-combustion capture using CO₂ absorption with monoethanolamine (MEA-ABS) (Kuramochi et al., 2012; Mikulčić et al., 2019; Dods et al., 2021; Dziejarski et al., 2023; CCS in Energy and Climate Scenarios - IEAGHG [Internet], 2026; Towards Zero Emissions CCS from Power Stations using Higher Capture Rates or Biomass - IEAGHG [Internet], 2026); it has reached the level of commercial operation (CCUS Projects Explorer – Data Tools - IEA [Internet], 2024) and is commonly used as a benchmark for CCS. Among the CDR technologies, direct air capture (DAC) refers to a set of technologies designed for the

removal of CO₂ from the atmosphere (María Erans et al., 2022), i.e., with application to CO₂ concentrations of around 430 ppm. Among the several processes proposed for DAC, two have been studied more than the others: i) temperature vacuum swing adsorption (TVSA) (Gebald et al., 2011; Gebald et al., 2014; Wurzbacher, 2015; Wurzbacher et al., 2016; Beuttler et al., 2019; Deutz and Bardow, 2021; Sabatino et al., 2021; Young et al., 2021; Wiegner et al., 2022; Young et al., 2022; Young et al., 2023; Bertoni et al., 2024; Postweiler et al., 2024; Rezo et al., 2024; Sievert et al., 2024; Postweiler et al., 2026; Stampi-Bombelli et al., 2024), which captures the CO₂ via cyclic adsorption on solid sorbents and regeneration at higher temperature and under vacuum conditions (Wurzbacher, 2015); and ii) alkaline absorption with subsequent calcium looping (ALK-ABS) (Mahmoudkhani and Keith, 2009; Holmes and Keith, 2012; Keith et al., 2018; De Jonge et al., 2019; Madhu et al., 2021; McQueen et al., 2021; Sabatino et al., 2021; An et al., 2022; Young et al., 2022), which captures the CO₂ through liquid absorption with KOH, followed by causticisation to obtain CaCO₃ and a subsequent calcium looping cycle to regenerate the

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CO₂ (Keith et al., 2018).

1.1. Industries with CO₂-lean flue gases

The greenhouse gas emissions from industries with CO₂-lean (<4%) flue gases are often described as hard-to-abate, owing to the thermodynamic principles of entropy and separation work. In this work, the term *hard-to-abate* is used in two study-specific senses that differ from its conventional meaning, which refers to sources whose emissions are intrinsically difficult to eliminate because they arise from process chemistry or feedstock rather than from energy use. An emission source is termed economically hard-to-abate when the levelized cost of avoidance through on-site point-source capture exceeds the cost of offsetting the same emissions by purchasing CDR credits, under the specific cost assumptions adopted here (Dufour-Décieux et al., 2025); and thermodynamically hard-to-abate when the minimum separation work (or exergy demand) required to capture its CO₂ exceeds that of direct air capture i.e., when capturing from that source is thermodynamically less favourable than capturing the same CO₂ from the atmosphere. Examples

of such industries are the production of aluminium and silicon, the pelletization of iron ore, natural gas combined cycles (NGCC), paint shops, breweries, as well as transformed industries engaged in propane dehydrogenation or methanol-to-olefin processes. These industries emit around 200 Mt CO₂ annually in Europe (NGCC ~188 Mt, aluminium production ~7 Mt, silicon production ~3.5 Mt, iron ore pelletization 2 Mt). Fig. 1 shows the location of these industries over Europe, showing that the majority of silicon and aluminium production and a significant share of the iron ore pelletization occurs in the Nordics while, on the other hand, the general spread of NGCC plants over Europe is not seen in the Nordics.

Fig. 2 maps, in terms of flow rate and CO₂ concentration, the typical flue gas ranges for major industries emitting CO₂-lean flue gases. The sources for CO₂ emissions vary across the specific industrial sector. In the aluminium and silicon production processes, carbon raw material and electrodes are used, which are consumed during the process and emitted as CO₂ in the flue gases. These net carbon emissions can be significantly reduced by replacing the fossil carbon in the electrodes with a biogenic material. While the literature focuses on methods to

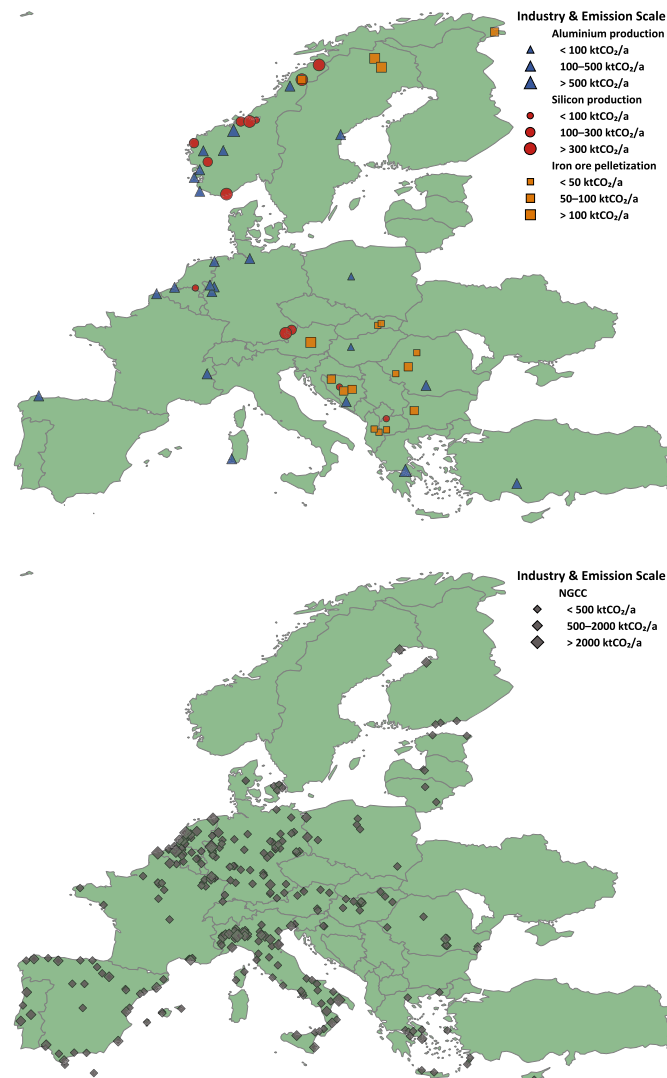


Fig. 1. Location and scale of point source emissions for identified industries with CO₂-lean flue gases over Europe (Industrial Reporting under the Industrial Emissions Directive 2010/75/EU and European Pollutant Release and Transfer Register Regulation (EC) No 166/2006 [Internet], 2025; Mineral Facilities of Europe [Internet], 2025; Elkem | Elkem.com [Internet], 2025). Data is collected from European Environment Agency report 2025, USGS database on mineral facilities in Europe 2010 and single plant websites.

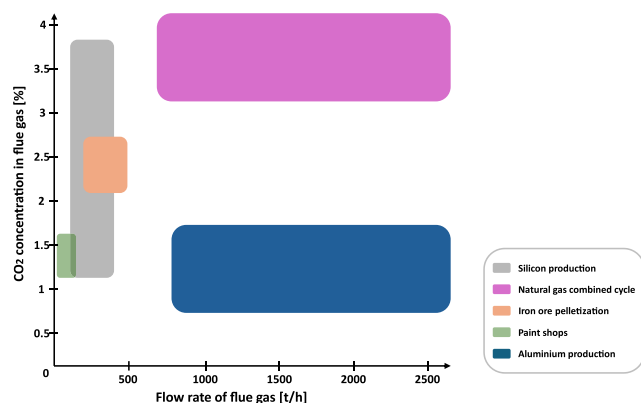


Fig. 2. Mapping of the flue gas ranges in terms of flow rate and CO₂ concentration for industries emitting CO₂-lean flue gases. The ranges are retrieved from the literature and linearly extrapolated based on the scale of the available plants (Normann et al., 2019; Wang and Song, 2020; Beiron and Johnsson, 2024; Andreasson and Emme, 2025; Hjartarson, 2025; Polysilicon Manufacturers, 2025) therefore the sharp edges in the Fig. should be seen as approximates.

increase CO₂ concentrations for both industries (up to 15% for silicon production and up to 10% for aluminium production) (Lassagne et al., 2013; Jilvero et al., 2014; Normann et al., 2019; Hoff et al., 2024), the concentration of CO₂ in the flue gases currently remains low. In paint shops, carbon emissions mainly originate from the fossil-fuel burners used in ovens and from the combustion of volatile organic compounds. Electrifying the burners can substantially reduce these emissions, although a small fraction of the volatile-related emissions may still remain. Iron ore pelletization involves heating the ore to approximately 1,200°C to facilitate the oxidation of magnetite. This process is often supported by coal combustion, contributing to CO₂ emissions together with the calcination of limestone additives. Transitioning to alternative heat sources would reduce the carbon emissions from fuel combustion, but the calcination emissions will remain. Breweries both emit and consume CO₂ during the fermentation process, which makes carbon capture and utilisation an attractive option for process optimisation. In the cases of future propane dehydrogenation or methanol-to-olefin processes, the only option is the use of feedstocks of biogenic origin.

As summarized above, industries emitting CO₂-lean flue gas have several emission reduction paths to consider (e.g. electrification, fuel shift), and this work sets its scope in the path of applying post-combustion carbon capture, thus providing values for comparison to other works that explore the techno-economic feasibility of alternative mitigation paths.

1.2. Point-source capture and direct air capture: two ends of the concentration spectrum

Post-combustion capture and DAC occupy opposite ends of the CO₂-concentration spectrum: the former benefits from the favourable thermodynamics of concentrated flue gases (typically 4–20% CO₂), whereas the latter is designed for the atmospheric concentration of around 430 ppm. Table 1 compares the main characteristics of the two approaches.

Table 1
Characteristics of CCS compared with DACCS.

	CCS	DACCS
CO ₂ Concentration	4–20 %	0.043 %
Capture rate	90% or above	Flexible
Location	Bound to CO ₂ source	Flexible
Scale	Limited to CO ₂ source	Limited to infrastructure
Operation	Bound to CO ₂ source	Weather dependent

The thermodynamic advantage of point-source capture comes with constraints on location, scale, capture rate, and operation. The location of a point-source capture unit is bound to the emission source, whereas the siting of a DAC plant is flexible and can be cost-optimised for example, through co-location with a storage site, which reduces the cost of transport infrastructure and removes the deployment constraint imposed by its absence. Similarly, the scale of point-source capture is set by the emission source, while the scale of DAC is limited by the availability of infrastructure, such as energy supply, space, and grid capacity. Since the purpose of point-source capture is to avoid emissions, units are typically designed for capture rates of 90% or above; for DAC, the purpose is to remove CO₂ at the lowest cost, so the capture rate is instead an outcome of cost optimisation. Finally, the operation of a point-source capture unit follows the operational pattern of the emission source, whereas DAC operation follows weather conditions. Hence CCS experiences fewer fluctuations in inlet conditions compared to DAC.

Industries with CO₂-lean flue gases (0.5–4.0% CO₂) fall between these two domains: their concentrations lie up to two orders of magnitude above the atmospheric level, yet below the range for which conventional post-combustion capture has been developed. It is therefore a priori unclear whether such streams are best served by the established post-combustion benchmark or by DAC-derived technologies applied as point-source capture systems a question that has received little attention in the literature.

1.3. Previous studies on CO₂ capture from CO₂-lean flue gases

The literature on post-combustion capture focuses on flue gas streams with CO₂ concentrations >4% (the value corresponding to natural gas combined cycle power plant), and often on CO₂ concentrations >10% (e.g., in coal power plants and industrial processes such as cement manufacture and refineries (Kuramochi et al., 2012; Leeson et al., 2017; Bui et al., 2018)). There is limited literature on carbon capture from CO₂-leaner flue gas streams. High and ultra-high capture rates for MEA-based absorption (MEA-ABS) have been demonstrated as technically feasible through both modelling and experimental studies (Abu-Zahra et al., 2007; Feron et al., 2019; Hirata et al., 2020; Jiang et al., 2020; Du et al., 2021; Mullen and Lucquiaud, 2024; Mullen et al., 2024; Michailos et al., 2025; Homaei et al., 2026; Towards Zero Emissions CCS from Power Stations using Higher Capture Rates or Biomass - IEAGHG [Internet], 2026). However, these high capture rates have primarily been investigated in the power sector, while industrial applications particularly those involving CO₂-lean streams remain comparatively underexplored.

Brandl et al. (2021) provided the broadest parametric assessment to date, evaluating MEA-ABS capture costs for CO₂ concentrations of 2–30% and demonstrating the economic feasibility of capture rates above 90%; they further showed that, at low concentrations, capture costs become increasingly sensitive to both concentration and flow rate. Their study is, however, limited to a single technology, excludes transport and storage, and leaves out the carbon footprint of the energy supplying the capture process. The remaining studies are single-case assessments for MEA-ABS without comparison to other technologies that might be competitive for capturing from low CO₂ concentrations: Jilvero et al. (2014) and Lassagne et al. (2013) each evaluated MEA-ABS for one specific aluminium plant (400 t/h at 3% CO₂ and 2,744 t/h at 1.2% CO₂, respectively), reporting capture costs of 111–125 €/t CO₂ and 115 €/t CO₂ for the studied design point only, and Normann et al. (2019) assessed one silicon plant (~370 t/h, 1–4.4% CO₂), reporting a capture cost of 130 €/t CO₂. Recently, Dorado-Alfaro et al. (2024) addressed the process thermodynamics of capture from CO₂-lean streams using absorption with ionic liquids, with the assessment confined to process performance rather than cost.

Three gaps emerge from this body of work. First, it is confined to a single capture technology, MEA-ABS; DAC-derived technologies such as ALK-ABS and TVSA whose modularity and design for dilute streams

make them natural candidates for CO₂-lean sources remain to be evaluated as point-source capture systems. Second, the carbon footprint of the energy supply is left out of all these studies, although at low CO₂ concentrations the energy demand per tonne captured, and thus the indirect emissions, is particularly high, so that capture and avoidance costs can differ substantially. Third, the resulting costs are yet to be benchmarked against the alternative of offsetting through the purchase of CDR credits, even though the higher capture cost from CO₂-lean flue gases after including transport and storage may surpass current CDR credit prices, making offsetting the more economic route depending on flow rate and concentration (Brandl et al., 2021; Du et al., 2021). Thus, while the available literature on capture from CO₂-lean flue gases remains limited, to the best of our knowledge no previous study has provided a comprehensive and consolidated comparison of full chain for these capture technologies under a consistent techno-economic framework for these streams, nor benchmarked the resulting costs against CO₂ offsetting.

The aim of this work is to investigate the economic performances of different post-combustion CO₂ capture technologies applied to CO₂-lean flue gas streams under a net-zero direct emissions design constraint. This work investigates the economic and technical performances of technologies originally developed for DAC when applied as first-of-a-kind point-source capture systems, and compares these Fig.s to the performances of the current benchmark, MEA-ABS. The impact from the carbon footprint of the energy input is also considered. Finally, the economic performances of the technologies considered are compared to the current prices of CDR credits in voluntary markets, to identify the most economic decarbonization path via carbon capture technologies for industries with CO₂-lean flue gas streams. Further, results from this work allow for evaluating the role of CO₂-lean industries as potential candidates for further deployment (and thus maturing and cost reduction) of the ALK-ABS and TVSA technologies as well as possible demands from these industries for CDR credits. Also, results allow for a better understanding of which CO₂-lean industries can be considered as hard-to-abate, both from an economic and a thermodynamic perspective.

2. Methodology

Fig. 3 shows the overall CO₂ capture and storage chain considered in this work, which consists of the direct contact cooler (DCC), capture unit, compression and liquefaction step, and transportation and storage. As seen, Fig. 3 also depicts the system boundaries used for the calculation of the levelised cost of capture (LCOC) and the levelised cost of carbon avoidance (LCOA), which are described in further detail below. The process models cover the whole chain, from the DCC to the compression and liquefaction stage. The inlet boundary is set as the incoming dry flue gas at 150°C, which is cooled to 40°C in the DCC. Then, the flue gas enters the capture stage, in which CO₂ is removed by

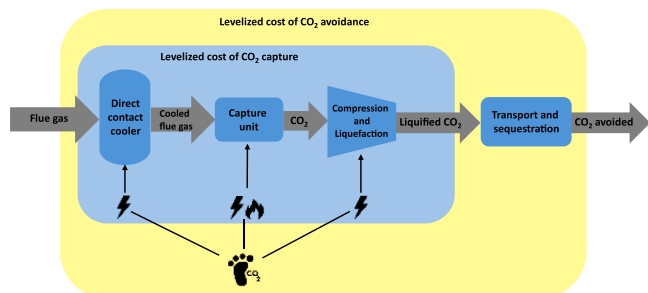


Fig. 3. System boundaries for calculating the cost of CO₂ capture (blue frame) and avoidance (yellow frame). The flue gas enters the DCC at 150°C and is cooled to 40°C before the capture unit. The liquefied CO₂ leaves the plant at 16 bar and -26°C, i.e., conditions compatible with medium-pressure ship transportation.

means of a specific capture technology. In this work, the closures of the energy and mass balances for the capture step are acquired from process models specific for each capture technology but where the outlet flow is in all cases set to have the same CO₂ concentration as atmospheric air (here taken as 400 ppm), in order for the process to reach net-zero direct CO₂ emissions by itself. Finally, the CO₂ gas stream is pressurised and liquefied in the compression and liquefaction unit to obtain conditions that are suitable for medium-pressure ship transport (16 bar, -26°C (Norlights.com [Internet], 2024)) and that comply with the purity limit required by the Northern Lights project. Considering the mapping shown in Fig. 2, this work considers flue gas streams that are characterised by flow rates in the range of 25–2,700 t/h, i.e., from the value for a small paint shop up to that for a large-scale aluminium production plant (Lassagne et al., 2013), and CO₂ concentrations in the range of 0.5–4.0%vol, where the chosen upper bound represents the flue gas from a natural gas combined cycle power plant.

Fig. 4 shows the overall framework for the techno-economic modelling of the three technologies considered. In this Fig., arrows represent data flows, while boxes represent models: the process models (blue boxes) close mass and heat balances to provide as output the sizing of the different process elements, which the cost model (yellow box) uses as input data. As illustrated, the capture step uses different process models for each of the three technologies considered (described in greater detail below), while all processes share the same models for the pre-capture and post-capture steps.

2.1. Process models

The overall process model (blue frame in Fig. 4) includes models which, by closing the mass and energy balances, provide the sizing of the three steps, i.e., pre-capture (DCC unit), capture (MEA-ABS, ALK-ABS, or TVSA process), and post-capture (compression and liquefaction unit). Regarding pre-capture, the DCC unit is designed as a column with wet packing, the size of which is determined by the varying water mass flow required to cool the flue gases to 40°C. At the other end of the process, the compression and liquefaction step is sized and modelled as a three-stage intercooled process that incorporates a refrigeration cycle that uses ammonia as the refrigerant, according to the description of Deng et al. (Deng et al., 2019). For more process information see *Supplementary Information S1-S4*. As mentioned above, the different processes considered here differ only with respect to the technology chosen for the capture step.

The MEA-ABS capture step is designed according to the description of Roshan Kumar et al. (2025a), whereby multiple absorption columns (if needed) and one stripper are used and based on the approach used for the flooding method (Biermann et al., 2022b) (for more design parameters see Table S1 in *supplementary information*).

The ALK-ABS capture step is designed according to the layout described by Sabatino et al. (2021), where the air contactor is sized according to the required capture rate, while the remaining components are sized assuming the optimised on-design parameters given by Keith

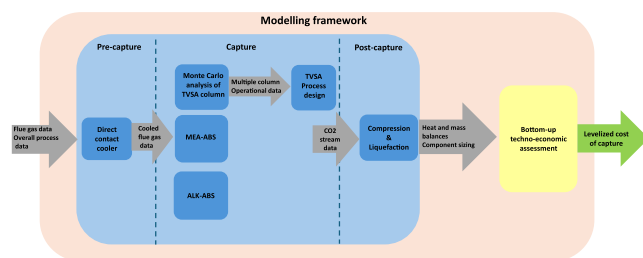


Fig. 4. Overall modelling framework for assessing the LCOC through closures of the mass and heat balances and sizing of the process. The boxes represent models (blue for process models and yellow for the economic model) and the arrows represent data flows.

et al. (2018). In line with the reference design (Keith et al., 2018), the calciner is fired with natural gas under oxy-fuel conditions, with oxygen supplied by a cryogenic air separation unit. The combustion CO₂ is thereby released internally together with the calcination CO₂ and routed to the compression and liquefaction unit. Consequently, the heat supply of the ALK-ABS process does not give rise to direct atmospheric CO₂ emissions, at the expense of a higher natural gas demand and the capital and energy costs of the air separation unit. This assumption has a decisive influence on the avoidance cost results presented in Section 3. To ensure a fair comparison, the power island included in the original design is here removed, so all the processes are assumed to cover their whole electricity need from the grid.

The TVSA capture step is modelled in four stages, involving the: i) development of a transient 1-dimensional model (implemented via the gProms software) (gProms Digital Process Design and Operations - Siemens Global [Internet], 2025) to describe the cyclic operation of one single adsorption column using Lewatit VP OC 1065 as sorbent (note that the inherently dynamic operation of adsorption columns (Do, 1998) requires a dynamic modelling framework); ii) running of Monte Carlo simulations (about 30,000 samples per concentration case; ranges and variables are presented in Table S2 in *supplementary information*), to map the performance of the column under different sets of design parameters; iii) sizing the other components of the capture process for each simulation; and iv) assessing the cost of the sized process for each simulation, so as to identify the cost-optimal design and operation.

The thin-plate contactor used in commercial TVSA-DAC units is designed for short gas–solid contact times, which suffices for the moderate capture rates targeted in DAC. In this work, the fixed outlet concentration of 400 ppm implies capture rates of 92–99% at inlet concentrations of 0.5–4%, requiring substantially longer contact times and larger sorbent inventories per unit of gas flow. Thus, this work considers three contactor designs alternative to the common thin-plate DAC design (Caram et al., 2020; Sabatino et al., 2021; Young et al., 2021): a) thicker plates, b) a classical adsorption column with greater bed height, and c) a stacked zigzag configuration, in which several thin plates are stacked to increase the contact thickness. When moving from atmospheric towards higher CO₂ concentrations, the plate thickness is increased by stacking (configuration c), which lengthens the gas residence time; once the stack thickness exceeds 1.5 m, the design transitions to a classical adsorption column (configuration b). Rather than being fixed a priori, the geometry follows from the cost-optimal bed depth of each case, so that each Monte Carlo design point is realised in whichever geometry corresponds to its optimal thickness.

The bed is heated indirectly, via immersed heat-exchanger tubes, during the pre-heating and desorption steps, avoiding direct steam contact with the sorbent, which would otherwise saturate the amine-functionalised Lewatit VP OC 1065 with water. The low heat transfer of this indirect heating lengthens the desorption step and lowers the volumetric productivity, which raises the number of parallel columns required and thus the cost of TVSA. The adsorption column model is validated against the DAC-condition model data of Young et al. (2022); the geometry is then scaled up to the flow conditions required here, up to a maximum column diameter of 3 m, the upper limit recommended by Subraveti et al. (2021). Adsorption columns must operate below the minimum fluidisation velocity, which limits the flow rate a single column can handle.

Although costs are the main criterion for decision-making in commercial operations, process efficiencies complement the picture with a long-term perspective that is less sensitive to, for example, variations in the prices of materials and energy inputs. Therefore, efficiency values on the basis of energy Eq. (1) and exergy Eq. (2) are calculated for the technologies considered as:

$$\eta_{en} = \frac{W_{min}}{EL_{in} + Q_{in}} \quad (1)$$

$$\eta_{exg} = \frac{W_{min}}{EL_{in} + Q_{in} * \left(1 - \frac{T_{env}}{T_{reg}}\right)} \quad (2)$$

where EL_{in} (MJ/kg) and Q_{in} (MJ/kg) denote the electricity and heat requirements, respectively, of the capture technology, and T_{env} (K) and T_{reg} (K) denote the ambient temperature and regeneration temperature, respectively, of the capture technology. Finally, W_{min} (MJ/kg) is the work of direct separation of CO₂ from an ideal mixture using a selective CO₂ membrane, given by Eq. (3) (Caram et al., 2020) as a function of the capture rate (CR) and the molar concentration of CO₂ in the flue gas (y_{CO2}):

$$W_{min}(CR, y_{CO_2}) = RT_0 \left[\frac{1}{CR} \ln(1 - CR) - \ln \frac{y_{CO_2}(1 - CR)}{1 - CR * y_{CO_2}} - \frac{1}{CR * y_{CO_2}} \ln(1 - CR * y_{CO_2}) \right] \quad (3)$$

2.2. Techno-economic assessment model

A bottom-up cost assessment model was developed based on the guidelines by Roussanaly et al. (2021), to estimate the capital expenditure (CAPEX) and operational and maintenance (O&M) expenses using as input data the mass and energy balance closures and the process sizing provided by the process models (Section 2.1). Table 2 presents the assumed parameter values for the bottom-up cost assessment model.

The CAPEX values of each component were calculated using cost functions from the literature (CEMCAP Action full title: CO₂ capture from cement production Type of action: D4.4 Cost of critical components in CO₂ capture processes, 2015; Pieper et al., 2018; De Lena et al., 2019; Subraveti et al., 2021; Biermann et al., 2022a; Young et al., 2022) and are reported in USD₂₀₂₁/tCO₂, adjusted to 2021 values using the Chemical Engineering Plant Cost Index, CEPICI (The Chemical Engineering Plant Cost Index ® - Chemical Engineering [Internet], 2025). For the ALK-ABS process, given the limited published data, the novel components were treated as first-of-a-kind and were scaled based on their sizing and the reference values reported by Keith et al. (2018), using the cost-scaling factors from the DOE/NETL power-plant recommendations and Eq. (4) (NETL, 2011). For modular components (in the TVSA and ALK-ABS processes), costs were scaled linearly. It should be noted that the ALK-ABS and TVSA cost estimates are first-of-a-kind values, whereas MEA-ABS is costed as a commercially mature technology; the implications of this maturity difference for future cost trajectories are discussed in Section 3.

The scaling relation and its parameters for the novel ALK-ABS items are given in Eq. (4) and Table 3. The 'other equipment' entry was estimated as 20% of the sum of the remaining components.

$$C = C_0 * \left(\frac{S}{S_0}\right)^n \quad (4)$$

The remaining equipment costs were calculated using the cost functions given in Table 4.

Regarding O&M costs, the variable share covers the costs for fuel, heat, electricity, chemical material and cooling water. For the ALK-ABS process, the heat and electricity demand of the slaker, calciner, washer, and pellet reactor are obtained by scaling the values reported by Keith et al. (2018) according to the respective mass flows. Regarding the fixed O&M costs, to calculate the labour costs for TVSA and ALK-ABS processes, the number of employees is assumed to increase proportionally to the total CO₂ captured per year, since a higher number of components is required due to the linear nature of the scale-up of these technologies (Young et al., 2022). However, this is not the case for MEA-ABS, where the number of employees is assumed to remain

Table 2

Assumptions made in the cost assessment model.

Parameter	Value	MEA-ABS	ALK-ABS	TVSA
Cost year	2021			
Life-time	25 years			
capture unit				
Cost of electricity	95.35 \$/MWh ([nrg_pc_205] Electricity prices for non-household consumers - bi-annual data (from 2007 onwards) [Internet] 2025)			
Cost of heat	36.4 \$/MWh (ADVANCED MANUFACTURING OFFICE Benchmark the Fuel Cost of Steam Generation)			
Utilisation factor	0.913			
Discount rate	8%			
Process		10%	50%	10%
contingency		TDC ¹		
Project		30%	35%	35%
contingency		EPC ²		
System		0% TPC ⁴	10%	10%
contingency ³				
Indirect costs		15%	15%	15%
		TDCPC ⁵		
Spare parts	0.5% TPC			
Owner cost	7% TPC			
Interest during construction	9% TPC			
Start-up capital	2% TPC			
Start-up labour	2.5% total labour			
Start-up energy	2% total yearly energy			
Indirect labour share	30% direct labour			
Maintenance	1.5% TPC			
Insurance	0.5% TPC			
Local tax fees	0.5% TPC			
Employee salary	64,900 \$/year			
Euro/USD	1.11			
Coal price (\$/GJ)	3.46 (Fuel Commodities and Carbon Prices – TYNDP 2020 — ENTSOs [Internet], 2025)			
Natural gas price (\$/GJ)	8.65 ([nrg_pc_203_c] Gas prices components for non-household consumers - annual data [Internet], 2025)			

¹TDC stands for total direct costs.²EPC stands for engineering, procurement and construction costs.³Only considered for ALK-ABS and TVSA as they combine new components (Roussanaly et al., 2021)⁴TPC stands for total plant costs.⁵Total direct costs with process contingency

constant with increasing plant size (Biermann et al., 2022a) since only the size of the equipment changes. Other fixed O&M costs, such as tax fees, insurance, and maintenance, are calculated according to the available guidelines in the literature (NETL, 2011; IEA GREENHOUSE GAS R&D PROGRAMME [Internet], 2017; Roussanaly et al., 2021) (see Supplementary Information Table S3).

As depicted in Fig. 3, the LCOC (\$/tCO₂) Eq. (5) considers the costs of the on-site capture steps and the carbon captured. To allow comparison to prices for CO₂ offsetting, the LCOA (\$/tCO₂) Eq. (6) offers a more

holistic view by including the costs for transport and storage and by relating the resulting total cost to the carbon avoided rather than the carbon captured (accounting for the carbon footprint of the energy input, see Eq. (7)). This work considers the carbon footprint associated with the heat and electricity demand of pre-capture, capture and post capture units (thus, excluding that associated to transport and storage), which are the major contributors to the indirect emissions of capture and storage processes (Sendi et al., 2024).

$$LCOC = \frac{C_{\text{annualized, capex}} + C_{\text{fixed O\&M}} + C_{\text{variable O\&M}}}{m_{\text{CO}_2, \text{captured}}} \quad (5)$$

$$LCOA = \frac{C_{\text{annualized capex}} + C_{\text{fixed O\&M}} + C_{\text{variable O\&M}} + m_{\text{CO}_2, \text{captured}} * C_{\text{t \& s}}}{m_{\text{CO}_2, \text{avoided}}} \quad (6)$$

$$m_{\text{CO}_2, \text{avoided}} = m_{\text{CO}_2, \text{captured}} - m_{\text{CO}_2, \text{electricity}} - m_{\text{CO}_2, \text{heat}} \quad (7)$$

where $C_{\text{annualized capex}}$ (\$/a) is the annualized capital expenditure using discount rate of 8% and economic lifetime of 25 years; $C_{\text{fixed O\&M}}$ (\$/a) and $C_{\text{variable O\&M}}$ (\$/a) are the annual fixed and variable operating and maintenance costs; $C_{\text{T\&S}}$ (\$/tCO₂) is the specific cost of CO₂ transport and storage; and $m_{\text{CO}_2, \text{captured}}$ (tCO₂/a) and $m_{\text{CO}_2, \text{avoided}}$ (tCO₂/a) are the annual masses of CO₂ captured and avoided, with $m_{\text{CO}_2, \text{avoided}}$ given by Eq. (7), where $m_{\text{CO}_2, \text{electricity}}$ (tCO₂/a) and $m_{\text{CO}_2, \text{heat}}$ (tCO₂/a) are the indirect emissions associated with the electricity and heat supplies.

The heat-related emissions $m_{\text{CO}_2, \text{heat}}$ in Eq. (7) are calculated as the product of the heat demand of the capture process and the carbon footprint of the respective heat supply option (Table S5). For coal and natural gas boilers, the carbon footprint comprises the direct combustion emissions (0.475 and 0.277 tCO₂/MWh_{heat}, respectively) plus the upstream emissions of the fuel supply chains (0.020 and 0.022 tCO₂/MWh). The assumption for MEA-ABS and TVSA is that there is an auxiliary boiler where fuel is combusted and emissions are released into atmosphere. For natural gas boilers equipped with CCS, only the residual (non-captured) direct emissions and the upstream emissions are counted (0.0425 tCO₂/MWh in total). Waste heat is assigned a zero-carbon footprint, as its emissions are attributed to the primary process. For electric boilers, the carbon footprint equals that of the country-specific electricity mix (Table S4). In the NGCC case, process heat for MEA-ABS is extracted as low-pressure steam from the steam cycle; the associated combustion CO₂ forms part of the flue gas treated by the capture unit, so the heat supply carries no additional direct emissions beyond the upstream natural gas emissions. For ALK-ABS, the calciner heat is supplied by oxy-fuel natural gas combustion with internal capture of the combustion CO₂, so that only the upstream emissions of the natural gas supply enter $m_{\text{CO}_2, \text{heat}}$.

For calculations of the capture cost, this work assumes as reference a cost for electricity of 95.35 \$/MWh and a natural gas price of 31 \$/MWh (Roussanaly et al., 2017). However, since energy costs and carbon footprint vary significantly depending on location and provision method (Roussanaly et al., 2017), this work considers a sensitivity analysis with four different heat provision methods (coal boiler, natural gas boiler, electric boiler, and waste heat) and four different electricity scenarios (corresponding to the prices and carbon footprints in Sweden, Spain, Germany, and Poland (Husebye et al., 2012; Carbon intensity of electricity generation, 2024; [nrg_pc_205] Electricity prices for non-household consumers - bi-annual data (from 2007 onwards)

Table 3

Scaling parameters for novel equipment items priced via Eq. (4).

Process unit	Technology	Cost basis / scaling variable	Scaling exponent (n)	Base scale (S ₀)	Reference CAPEX (C ₀) (MUSD ₂₀₂₁)	Source
Air contactor	ALK-ABS	Plant scale (tCO ₂ /y)	1 (modular)	980000(tCO ₂ /y)	212/1667	(Keith et al., 2018)
Pellet reactor	ALK-ABS	Plant scale (tCO ₂ /y)	1 (modular)	980000(tCO ₂ /y)	130.8	(Keith et al., 2018)
Fines filter	ALK-ABS	Fines output (t/hr)	0.68	29.4 (t/hr)	30.9	(Keith et al., 2018)
Buildings	ALK-ABS	Plant scale (tCO ₂ /y)	0.55	980000(tCO ₂ /y)	6.7	(Keith et al., 2018)

Table 4

Cost functions used for the remaining equipment items.

Process unit	Technology	Cost basis / scaling variable	Cost function	Source
Columns (DCC, Absorber, Stripper)	All	Column volume (m ³) (packing)	$C = 91764 \cdot x^{0.6154}$ (€ 2015)	(Biermann et al., 2022a)
Reboiler	MEA-ABS	Heat exchange area (m ²)	$C = 1675.8 \cdot x^{0.8794}$ (€ 2015)	(Biermann et al., 2022a)
Cross heat exchangers	MEA-ABS	Heat exchange area (m ²)	$C = 2862.6 \cdot x^{0.7988}$ (€ 2015)	(Biermann et al., 2022a)
Flue gas fan	MEA-ABS	Power (kW)	$C = 2774.5 \cdot x^{0.7814}$ (€ 2015)	(Biermann et al., 2022a)
Pumps	MEA-ABS	Power (kW)	$C = 32147 \cdot x^{0.6029}$ (€ 2015)	(Biermann et al., 2022a)
Reclaimer	MEA-ABS	Mass flow rate HSS (kg/hr)	$C = 68483 \cdot x^{0.9315}$ (€ 2015)	(Biermann et al., 2022a)
Tanks and vessels	MEA-ABS	Volume (m ³)	$C = 66927 \cdot x^{0.5047}$ (€ 2015)	(Biermann et al., 2022a)
Pre- and post-filter	MEA-ABS	Number of filters	$C = 90000 \cdot x$ (€ 2015)	(Biermann et al., 2022a)
Active carbon filter	MEA-ABS	Number of filters	$C = 139000 \cdot x$ (€ 2015)	(Biermann et al., 2022a)
ASU	ALK-ABS	Oxygen mass per day (t/d)	$C = 22\,000\,000 \cdot (x / 432)^{0.6}$ (€ 2016)	(De Lena et al., 2019)
Calcliner	ALK-ABS	Heating power (kW)	$C = 2.07 \cdot 193000 \cdot x^{0.65}$ (€ 2014)	(CEMCAP Action full title: CO2 capture from cement production Type of action: D4.4 Cost of critical components in CO2 capture processes, 2015)
Slaker	ALK-ABS	Gas volume flow rate (m ³ /hr)	$C = 2.1 \cdot 89700 \cdot x^{0.5}$ (€ 2014)	(CEMCAP Action full title: CO2 capture from cement production Type of action: D4.4 Cost of critical components in CO2 capture processes, 2015)
Adsorption column	TVSA	Heat exchange area (m ²)	$C = 1.5 \cdot (70 \cdot x^{1.2} + 32000)$ (\$ at Cepci 532.9)	(Young et al., 2022)
Sorbent	TVSA	Mass (kg)	$C = 7641 \cdot x$ (\$ 2019)	(Young et al., 2022)
Vacuum pumps	TVSA	Volumetric flow (m ³ /hr)	$C = 423.9 \cdot x^{0.653} + 30000$	(Subraveti et al., 2021)
Compressors	TVSA	Volumetric flow (m ³ /hr)	$C = 121.412 \cdot x^{0.9} + 1.032 \cdot \text{inlet pressure (bar)}$	(Subraveti et al., 2021)
Heat exchangers	TVSA	Heat exchange area (m ²)	$C = (12003 \cdot x^{0.603}) \cdot 1.01118^{\text{pressure}}$	(Subraveti et al., 2021)
Switching valves	TVSA	Number of valves	$C = 6000 \cdot x$ (€2015)	(Subraveti et al., 2021)
Water removal	TVSA	Water mass flow rate (kg/hr)	$C = 230 \cdot x$ (\$ 2015)	(Subraveti et al., 2021)
Compressor stage 1	Compression & liquefaction	Power (kW)	$C = 831732 + 77.08 \cdot x^{1.5} - 0.676 \cdot x^2$ (€ 2015)	(Deng et al., 2019)
Compressor stage 2	Compression & liquefaction	Power (kW)	$C = 1040305 + 17.85 \cdot x^{1.5} - 0.101 \cdot x^2$ (€ 2015)	(Deng et al., 2019)
Compressor stage 3	Compression & liquefaction	Power (kW)	$C = 1316807 + 1.78 \cdot x^{1.5} + 0.0223 \cdot x^2$ (€ 2015)	(Deng et al., 2019)
Compressor stage 4	Compression & liquefaction	Power (kW)	$C = 1168183 - 51.28 \cdot x^{1.5} + 16476 \cdot x^2$ (€ 2015)	(Deng et al., 2019)
Refrigeration compressor stage 1	Compression & liquefaction	Power (kW)	$C = 736639 + 79.94 \cdot x^{1.5} - 0.769 \cdot x^2$ (€ 2015)	(Deng et al., 2019)
Refrigeration compressor stage 2	Compression & liquefaction	Power (kW)	$C = 889060 + 28.24 \cdot x^{1.5} - 0.225 \cdot x^2$ (€ 2015)	(Deng et al., 2019)
Recirculation compressor	Compression & liquefaction	Power (kW)	$C = 287.2 \cdot x + 270 \cdot \text{inlet pressure} + 29439 \cdot \text{pressure ratio} + 952751$ (€ 2015)	(Deng et al., 2019)
Pressurized vessel	Compression & liquefaction	Volume (m ³)	$C = 70693 \cdot x^{0.562} \cdot 1.01522^{\text{pressure}}$ (€ 2015)	(Deng et al., 2019)
Dehydration	Compression & liquefaction	Water mass flow rate (kg/hr)	$C = 90 \cdot x$ (€ 2015)	(Deng et al., 2019)

[Internet], 2025; Carbon Footprint of Heat Generation [Internet], 2025; Renewable Energy Sources and Climate Change Mitigation — IPCC [Internet], 2025), as presented in Table S4 and S5 in *Supplementary information*). The coal and natural gas calciners were assigned the same CAPEX, as they share essentially the same reactor and differ mainly in the fuel-handling system (McQueen et al., 2021), while the electric calciner was assigned a 50% higher CAPEX to account for the electrical heating system (Quevedo Parra and Romano, 2023). The LCOA also includes the material consumption and the indirect emissions incurred by sorbent replacement presented in Table S6 in *Supplementary information*. However, the embodied emission for plant construction is not included as it has been regarded less significant (Deutz and Bardow, 2021; Sendi et al., 2024). Moreover, since the transport cost and carbon footprint are dependent upon the distance to the storage site, a transport and storage cost of 105 \$/tCO₂ is assumed (i.e., the mean value of the range reported for Europe, 35–175 \$/tCO₂ (Karlsson, 2024)), while as mentioned above the carbon footprint of transport and storage is neglected.

3. Results and discussion

Fig. 5 shows the LCOC for the three technologies considered over the flue gas window relevant to industrial cases (0.5%–4.0% CO₂, 500–2,700 t/h; see Fig. 2). The results for lower flue gas flow rates are provided in the *Supplementary Information*, Fig. S8. The results indicate that, among the three technologies considered, MEA-ABS yields the lowest LCOC over the whole flue gas window shown in Fig. 2, while the TVSA process has the highest cost as high numbers of columns are needed to handle the flue gas flow rates considered – implying a high level of capital investment. The cost-optimal TVSA designs operate with long cycle times (39,400–49,700 s, Table S7), which follow from the near-complete capture requirement combined with the low heat transfer of the indirect bed heating, and translate into low volumetric productivities (93–141 kg/(m³·day)). These are reflected in the reported costs through the large number of parallel columns required, which is why TVSA has the highest LCOC over the flue gas window shown in Fig. 5. Its modular scaling, however, avoids the steep small-scale cost

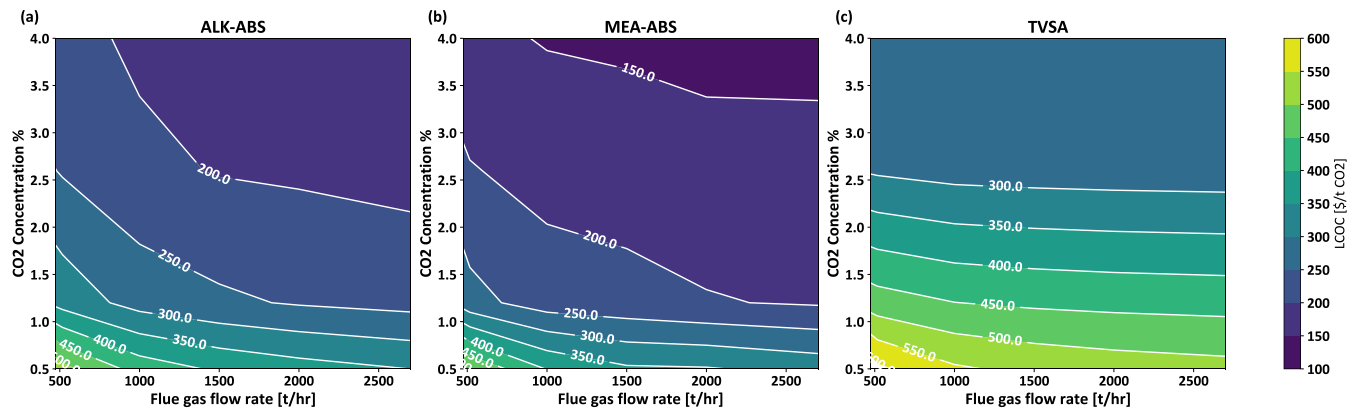


Fig. 5. Levelised costs of capture (LCOC) as a function of CO₂ concentration and flue gas flow rate, for: a) ALK-ABS; b) MEA-ABS; and c) TVSA. Contour lines indicate constant LCOC values, generated by linear interpolation between the simulated design points.

penalty of the monolithic technologies, so that TVSA becomes the lowest-cost option at small flue gas flow rates, as discussed below. Regarding TVSA, it should be noted that while this work considers only one TVSA configuration consisting of continuous capture in a configuration of parallel columns, there are alternative TVSA implementations (such as the moving bed configuration described by Svante (Svante: Carbon Capture Solutions for a Cleaner Future [Internet], 2025)) which might yield lower costs for TVSA than those reported here. In addition, it should be noted that this work only considers Lewatit VP OC 1065 as sorbent which is the commonly recommended for carbon capture purposes, while other sorbents might deliver better economic performances. Furthermore, this work uses cost expressions for Year 2021, i.e., at which time-point TVSA (and also ALK-ABS) had not been deployed at large scale and thus had greater potentials for cost reductions in the future compared with MEA-ABS.

The LCOC data presented in Fig. 5 can be explained by two main drivers: the flue gas flow rate, and the capture rate corresponding to each inlet CO₂ concentration (given the outlet CO₂ concentration set to 400 ppm). Regarding the flow rate, the extent of the cost decrease associated to higher flow rates depends on the scale-up pattern of each capture technology. The scale-up of MEA-ABS and ALK-ABS follow roughly monolithic patterns (significantly reduced cost with scale, especially for small sizes) even though in the ALK-ABS process the air contactor and pellet reactor (capture cycle) sections are modular. In contrast, following the modular nature of TVSA, the specific LCOC is almost constant with scale (the higher specific cost at the smaller scales is due to the monolithic scaling of the DCC and the compression and liquefaction units).

An increase in inlet CO₂ concentration is expected to yield decreased LCOC in regular carbon capture processes (Husebye et al., 2012; Brandl et al., 2021), as capture from a flue gas with a higher CO₂ concentration will benefit from the more-favourable thermodynamics and kinetics (thereby yielding reduced sizing of the equipment and thus lower costs). Capture rate has the opposite effect on LCOC, since the last amounts of CO₂ are at the lower concentration (Biermann et al., 2018). Hence higher capture rates yield higher LCOC. However, it is crucial to note that this work sets as design constraint an outlet CO₂ concentration of 400 ppm, which generally implies that higher inlet CO₂ concentrations require higher capture rates (thereby pushing for larger sizing of the equipment and thus higher costs). Thus, the impact of inlet CO₂ concentration on the LCOC is not obvious a priori of the calculations. Results in Fig. 5 show that the LCOC decreases as the inlet CO₂ concentration is increased, for all technologies. Thus, the positive effect of favoured thermodynamics transcends the higher requirement in terms of capture rate.

While Fig. 5 shows MEA-ABS as the economically most favourable option for carbon capture at flue gas flow rates above 500 t/h, zooming

into the lower flow-rate region (Fig. 6) allows identification of conditions under which the other two capture technologies considered, TVSA and ALK-ABS, become competitive. The solid red border lines in Fig. 6 represent LCOC values derived from a finite number of simulated design points, with the remainder of the space filled by piecewise-linear interpolation. Because parts of the response surface are interpolated linearly, the precise contour intersections should not be interpreted as sharp economic transition boundaries; moreover, even small cost differences can shift a boundary in favour of another technology. Hence, the dashed black regions indicate broader and more conservative domains in which TVSA or ALK-ABS consistently exhibit lower LCOC values than the competing options across the evaluated parameter space. These boxes thus represent regions where the respective technology is competitive and should be considered as an option.

According to Fig. 6, at flow rates below 200 t/h, TVSA can offer the lowest capture cost particularly at lower inlet CO₂ concentrations mainly due to the modular nature of the technology, which reduces scale penalties at small throughputs. ALK-ABS shows the lowest costs of all three technologies only within a narrow operating window (CO₂ concentrations <1% and flow rates between 100–200 t/h). This region corresponds to specific industrial emitters such as large-scale breweries (Brewery CO₂ Safety Monitors – CO₂ Meter [Internet], 2025; Looking

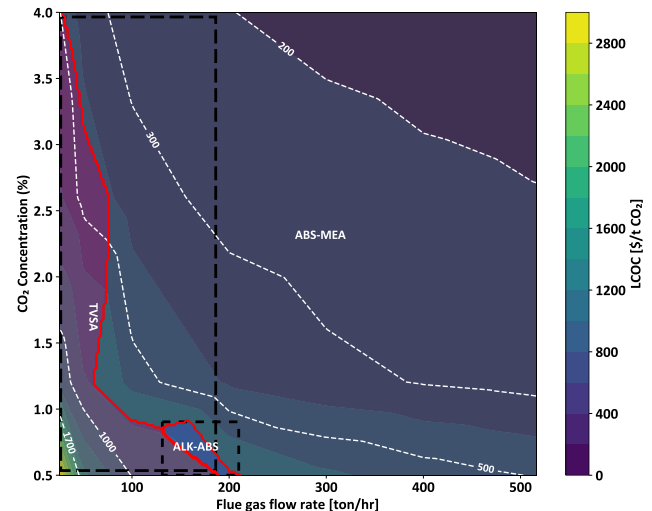


Fig. 6. LCOC as a function of flue gas flow rate for the three technologies investigated. Each region is labelled according to the technology exhibiting the lowest LCOC within that region. Contour lines indicate constant LCOC values, generated by linear interpolation between the simulated design points.

Closer at CO₂ Recovery at Your Brewery - First Key Consulting [Internet], 2025) and smaller silicon production sites. For breweries in particular, the calculated capture cost is comparable to the current market price of food-grade CO₂, suggesting a potentially viable niche application despite the limited parameter space in which ALK-ABS is economically favourable.

Fig. 7 shows, for CO₂ concentrations representative of the four different industrial applications in Fig. 2, the LCOC as a function of flue gas flow rate for each of the three capture technologies considered. The shaded regions represent the $\pm 30\%$ error span of the cost estimate used as the standard sensitivity range of the LCOC (Pickett et al., 2025). As shown, for very low flow rates, all three technologies end up in a similar capture cost range, especially when considering the sensitivity ranges. This makes it difficult to decide which technology is more suitable based on capture costs. Similarly, at concentrations $<1\%$ (Fig. 7, a and b) MEA-ABS and ALK-ABS have highly overlapping cost values. Therefore, for these conditions, case-specific variables such as heat availability, potential for process integration, space requirement (Roshan Kumar et al., 2025b), flexibility of the processes, and impact of the carbon footprint of energy supply, can determine the most-suitable process.

Fig. 8 shows the LCOC breakdown into annualized CAPEX, fixed O&M expenses, and variable costs (consumption of heat, electricity, and materials) for each technology at different CO₂ concentrations and a flow rate of 1,000 t/h. Normalized cost breakdowns are provided in Supplementary Figs S6 and S7. The major share (30%–50%) of the cost for most of the cases corresponds to annualized CAPEX, partially due to the low concentrations of CO₂ in the flue gas leading to larger sizing of

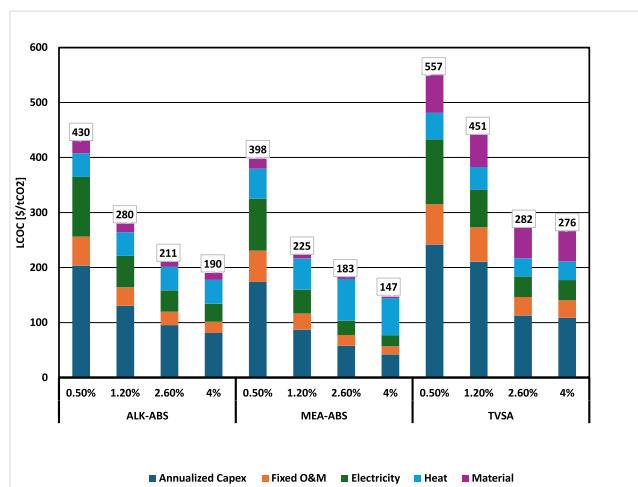


Fig. 8. Breakdown of the levelised cost of capture for the three capture technologies investigated, for different CO₂ concentrations (0.5 – 4%) and for a flue gas flow rate of 1,000 t/h.

equipment – it is noteworthy that the share of capital costs gradually decreases significantly with increased CO₂ concentration. It should be noted that these results represent first-of-a-kind cost estimates for these technologies, so the share of annualized CAPEX can be expected to be substantially lower upon wider deployment.

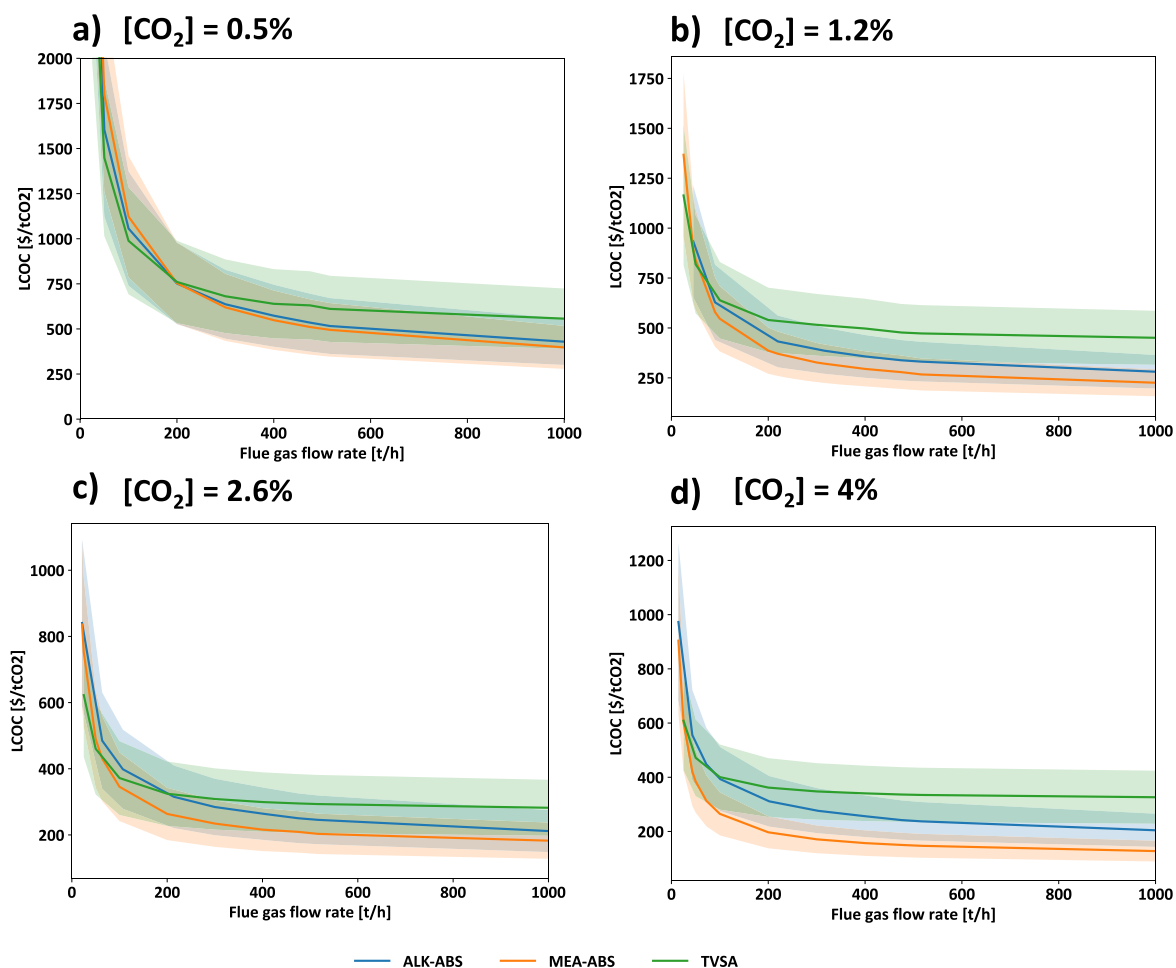


Fig. 7. Levelised cost of capture provided by each technology for CO₂ concentrations of a) 0.5%; b) 1.2%; c) 2.6%; and d) 4.0%, representative of different industrial cases.

Similar to annualized CAPEX, the cost share of fixed O&M expenses decreases with increasing CO₂ concentration for all three technologies. This trend is more pronounced for MEA-ABS due to its monolithic scaling behaviour, in contrast to TVSA which scales linearly on all capture components, and ALK-ABS, for which the most capital-intensive components are modular and thus scales linearly overall.

The LCOC share associated with electricity consumption follows a similar trend across all technologies, exhibiting its highest contribution at lower CO₂ concentrations. This is primarily due to fan power requirements remaining approximately constant while the amount of CO₂ captured decreases. The cost share of heat remains constant for ALK-ABS and varies only slightly for TVSA, reflecting their approximately constant heat of desorption. In contrast, the heat cost share for MEA-ABS increases with increasing capture rate, driven by higher reboiler duties.

Material costs represent a minor share of total costs for MEA-ABS and ALK-ABS but constitute a significant fraction (13–24%) for TVSA due to the high specific cost and make-up rate of the sorbent material (Deutz and Bardow, 2021). While such material consumption levels may be justifiable for CO₂ capture from air where higher CDR credit prices are anticipated they pose a substantial challenge for the economic feasibility of CO₂ capture from flue gas streams.

Fig. 9 presents the levelized cost of CO₂ avoidance (LCOA) for the three capture technologies applied to four industrial processes, shown as a function of the carbon intensity of the heat and electricity supplies. The industrial case studies were selected based on data availability and represent different plant scales: a small-scale silicon production facility (100 t/h, 10 kt/a, [CO₂] = 1.2%, small-scale), a large-scale iron ore pelletization plant (500 t/h [CO₂] = 2.6%, approximately 5 Mt/a (Kiruna Iron Ore Mine, 2025), large scale), and average-scale aluminium production (1,000 t/h, [CO₂] = 1.2%, approximately 125 Kt/a aluminium¹, medium scale) and natural gas combined cycle (NGCC) plants (2,700 t/h, [CO₂] = 4%, approximately 500 MW²). The carbon intensity ranges correspond to different heat generation technologies and country-specific electricity mixes. Boxes marked with a red cross in Fig. 9 indicate combinations for which no net CO₂ avoidance is achieved. Cells with costs exceeding 1,500 \$/tCO₂ are shown in grey, as these values are above the highest CDR credit prices currently available. Incongruous combinations such as NGCC paired with coal boilers, electric boilers, or waste heat, are indicated by hatched cells. The result data of the sensitivity analysis of transportation and storage costs are given in the Supplementary Information.

As shown in Fig. 9, TVSA exhibits the highest avoidance costs (consistent with its performance in capture cost) and shows a strong sensitivity to the carbon intensity of the energy supply, with a particularly pronounced dependence on heat-related emissions and a weaker dependence on electricity. ALK-ABS yields the lowest LCOA for aluminium production, iron ore pelletization, and silicon production. This is mainly attributable to the use of oxy-fuel combustion, which provides heat with integrated CO₂ capture and thereby avoids direct emissions from heat generation. However, this result is contingent on the oxy-fuel configuration. Were the calciner air-fired and its combustion CO₂ vented, the heat supply would carry direct emissions of 0.277 tCO₂/MWh_{heat}, equivalent to 0.38 tCO₂ per tCO₂ captured at the calciner duty assumed here. The avoided fraction would fall by 38 percentage points, from 85–96% to 47–58% depending on the electricity supply, raising the ALK-ABS LCOA by 65–80% if the remaining costs were held constant. Under that configuration ALK-ABS would lose its LCOA advantage in all four industrial cases, consistent with the limited removal potential reported for solvent-based DAC with fossil-fired regeneration (Bouaboula et al., 2025a). In contrast, the LCOA of both MEA-ABS and TVSA is strongly influenced by the carbon footprint of the

heat supply, so that MEA-ABS exhibits the lowest LCOA only in the case of NGCC combined with a low-carbon heat source. In this configuration, process heat is extracted from the low-pressure steam turbine, and the associated CO₂ emissions from natural gas combustion are captured. Due to this strong dependence on heat-related emissions, MEA-ABS achieves competitive LCOA values primarily when waste or excess heat is available on site a condition that is relatively uncommon, although partial heat integration may be feasible depending on a site-specific conditions.

The levelized cost of CO₂ avoidance (LCOA) can be compared with CDR credit prices to inform the choice between point-source CO₂ capture and CO₂ offsetting. To ensure a fair comparison, this work considers only CDR credits that provide permanent storage via direct air capture (DAC). At present, the main providers of such CDR credits are Climeworks, and Carbon Engineering, which employ TVSA-DAC, and ALK-ABS-DAC, respectively. Climeworks currently offers CDR credits at approximately 1,000 \$/tCO₂ for small-scale purchases (Climeworks, 2022) and around 600 \$/tCO₂ for large-scale purchases (Climeworks verifies that it captured and stored CO₂ for Microsoft, Stripe, and Shopify | The Verge [Internet], 2025). The corresponding prices reported by Carbon Engineering are approximately 395 \$/tCO₂ (Direct Air Capture [Internet], 2025). Based on these values, a CDR credit price range of 400–600 \$/tCO₂ is adopted in this work as an illustrative benchmark for DAC-based CO₂ removal.

Comparing the lowest avoidance cost for each industry in Fig. 9 with CDR credit prices, NGCC ([CO₂] = 4%, 500 MW) is the only case in which point-source capture is clearly more cost-effective than purchasing CDR credits: MEA-ABS with low-carbon heat achieves the lowest LCOA of the technologies considered, and ALK-ABS with coal or natural gas calciner also falls below the assumed CDR credit price range of 400–600 \$/tCO₂ for this case, though higher than MEA-ABS with low-carbon heat. For iron ore pelletization ([CO₂] = 2.6%, 5 Mt/a) and aluminium production ([CO₂] = 1.2%, 100 kt/a), the estimated LCOA lies below the CDR credit price range, but the overlap with the uncertainty ranges makes it difficult to determine conclusively whether capture is economically preferable to offsetting. For silicon production ([CO₂] = 1.2%, 10 kt/a), purchasing CDR credits appears to be the lower-cost option. These results correspond to specific plant sizes: larger silicon plants than the small-scale case considered here would have lower LCOA but are still expected to remain above current CDR credit prices, whereas iron ore pelletization plants smaller than the case considered here would have LCOA above CDR prices.

Therefore, from an LCOA (Scope 2) perspective and under the cost assumptions adopted here, emissions from silicon production, small-scale aluminium production, and small- and medium-scale iron ore pelletization can be classified as economically hard-to-abate in the sense defined in Section 1.1, i.e. their avoidance cost through on-site capture exceeded the cost of offsetting the same emissions with CDR credits (Dufour-Décieux et al., 2025). It should be emphasised that this capture-versus-offset comparison reflects current cost assumptions and market conditions. Future technological learning, infrastructure development particularly for CO₂ transport and storage and the maturation of the CDR market are expected to reduce the costs of both point-source capture and CDR, potentially at different rates. The relative competitiveness of LCOA and CDR credit prices, and hence the preferred mitigation option for a given source, may therefore change over time; the results presented here should be read as representative of the current state rather than as long-term general conclusions.

The above results and discussion must also be interpreted in view of several uncertainties. The assumed reference costs for transport and storage (105 \$/tCO₂) are subject to considerable uncertainty and depend strongly on the configuration of the transport and storage chain (Ouvray et al., 2024). A sensitivity analysis spanning these costs from 35 to 175 \$/tCO₂ (see Supplementary Information S9–S11) indicates that transport and storage can contribute with 10–50% of total CO₂ avoidance costs, making them a key cost driver for first-of-a-kind capture deployments. These findings underscore the critical importance of

¹ Estimated value based on the parameters reported by Lassagne et al. (2013).

² Estimated value based on the parameters reported by Du et al. (2021).

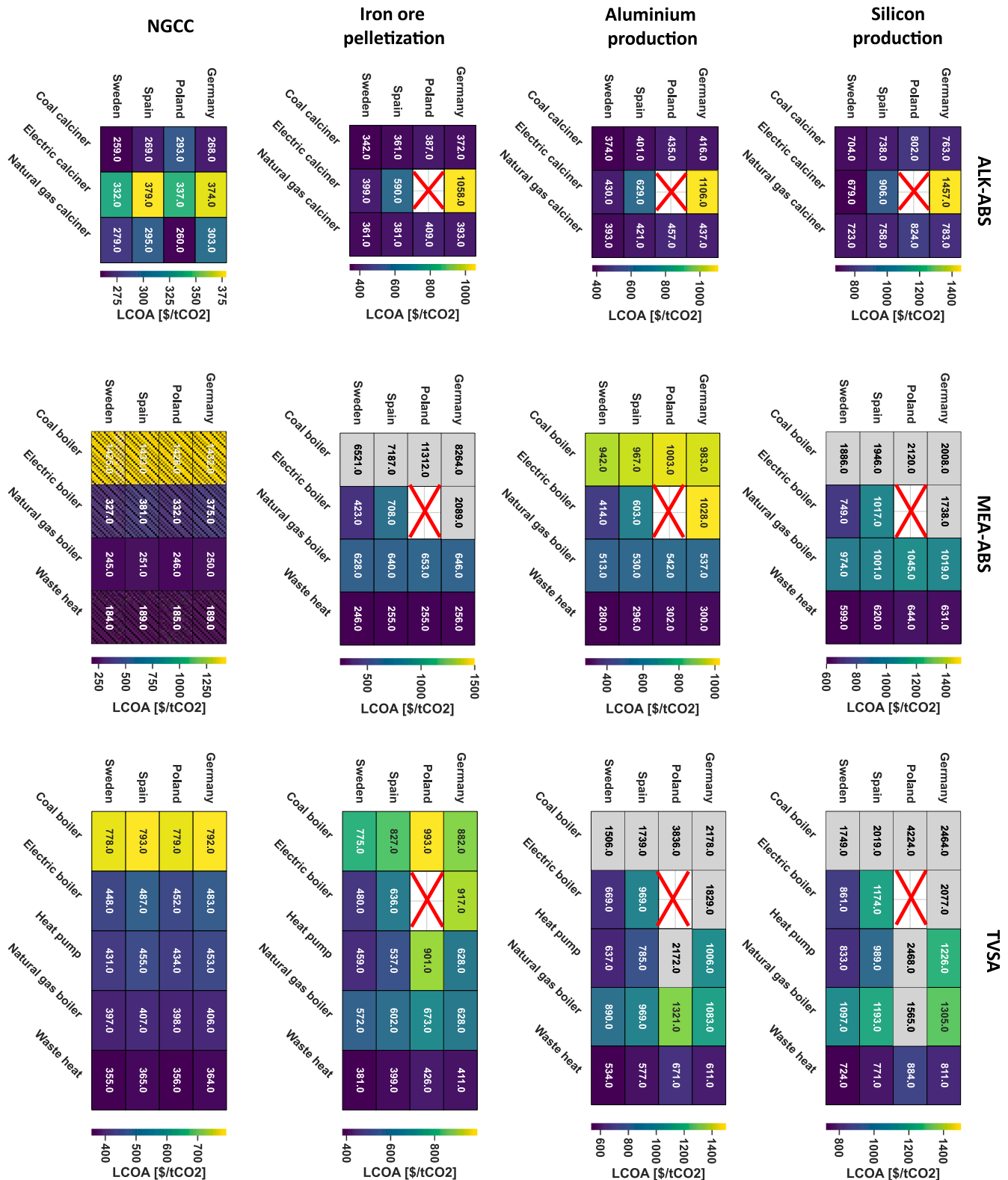


Fig. 9. LCOA as a function (María Erans et al., 2022) of the carbon footprint of the heat and electricity supply, for different countries and heat provisions, for the cases of silicon production (100 t/h, 10 kt/a, [CO₂] = 1.2%, small-scale), aluminium production (1,000 t/h, [CO₂] = 1.2%, approximately 125 Kt/a aluminium, medium scale), iron ore pelletization (500 t/h [CO₂] = 2.6%, approximately 5 Mt/a (Kiruna Iron Ore Mine, 2025), large scale), and NGCC (2,700 t/h, [CO₂] = 4%, approximately 500 MW). Cells with costs above highest CDR credit prices available (1500 \$/tCO₂) are coloured as grey. Cells corresponding to incongruous combinations are hatched. Note the different scales for the colour gradients.

developing low-cost CO₂ transport and storage infrastructure to incentivize CCS deployment in front of relying on offsetting. For example, for aluminium production, transport and storage costs in the lower end (Supplementary Information S15) yield a avoidance cost similar to that of the lower bound of the considered range for CDR credit prices. Conversely, high T&S costs may delay CCS deployment and instead incentivize reliance on CDR technologies.

Moreover, a baseline uncertainty of $\pm 30\%$ is applied to cost estimates; this can increase to $\pm 50\text{--}60\%$ for emerging and first-of-a-kind technologies, which makes it difficult to draw firm conclusions regarding economic competitiveness. In this sense, this work focuses on first-of-a-kind cost estimates, but it is important to distinguish between the maturity levels of the technologies considered: while MEA-ABS is commercially mature and therefore expected to experience only modest future cost reductions, ALK-ABS and TVSA are novel, DAC-derived technologies likely to benefit from substantial learning effects as deployment occurs. Studies of future DAC deployment suggest capital cost reductions of approximately 10–15% per cumulative doubling of installed capacity (Sievert et al., 2024). Under such learning trajectories, ALK-ABS and TVSA could become increasingly competitive relative to MEA-ABS over time, despite their higher initial costs.

The three technologies assessed here are those for which consistent techno-economic data were available, while a broader set of processes is under development for capture from dilute CO₂ sources. These include electrochemically driven routes, such as pH-swing (Bouaboula et al., 2025a,b) or ALK-ABS with bipolar membrane electrodialysis (Sabatino et al., 2020; Young et al., 2022), and configurational variants of the processes considered here, for example moving-bed TVSA (Svante: Carbon Capture Solutions for a Cleaner Future, 2025). The electrochemically driven routes share a feature that bears directly on the cost drivers identified above: regeneration by electrical or electrochemical work rather than by heat, which removes the dependence on the carbon intensity of the heat supply that governed the LCOA results in Fig. 9. This shifts the energy demand towards electricity, so that the avoidance cost of such processes would depend on the price and carbon footprint of the electricity supply to a degree comparable with the dependence on the heat supply found here for MEA-ABS and TVSA. A quantitative benchmark against the values published for these processes falls outside the scope of this work, since their assumptions on energy prices, discount rate, plant lifetime, and the carbon footprint of the energy supply differ substantially from those adopted here; evaluating them within the present framework is a natural direction for future work.

Beyond cost considerations, several practical factors must also be considered when implementing point-source carbon capture, including challenges related to retrofitting existing plants, space constraints, and potential mismatches between the remaining lifetime of the host facility and that of the capture unit (Roshan Kumar et al., 2025b).

Finally, to complement the work with data that are less-exposed to the volatility of economic factors (e.g. capital and material costs, heat and electricity prices), Fig. 10 presents the energy and exergy efficiencies of the three capture technologies considered and compares them to the values for the benchmark DAC technology, s-DAC (House et al., 2011; Hoseinpouri et al., 2023). The efficiencies of all three processes are dominated by an increase with CO₂ concentration. For TVSA, this is due to the strong enhancement of the adsorption kinetics, which overcomes the negative impact of the required higher capture rate. For MEA-ABS, the non-monotonic trends of efficiency with CO₂ concentration arise from opposing effects on the specific heat demand: increasing the CO₂ concentration reduces it, while the implied increased capture rate raises it (see Fig. S6). The efficiencies of ALK-ABS increase only slightly with CO₂ concentration as the energy requirements change scarcely, although there are minor reductions in the fan and pump power demands.

From the energy perspective (Fig. 10a), TVSA shows the best performance of the three processes studied here, with s-DAC showing to be at least as competitive. From the exergy perspective (Fig. 10b), both MEA-

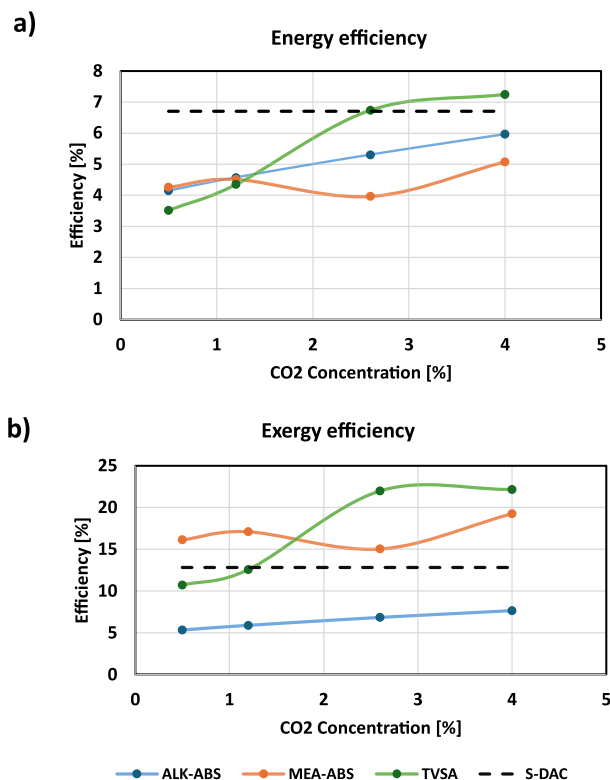


Fig. 10. Efficiencies of the capture step in terms of a) energy and b) exergy, for the three processes considered with direct air capture via solids adsorption. Data shown are for a flue gas flow rate of 1,000 t/h.

ABS and TVSA perform better than s-DAC, leaving ALK-ABS as the least-competitive option. Therefore, from exergy perspective none of the point source emissions with CO₂ concentration above air can be considered thermodynamically hard-to-abate emissions (Dufour-Décieux et al., 2025). In general terms of efficiency, TVSA is the most favoured process for CO₂ concentrations >1–2%, which is not in line with the cost assessment results and thus highlights the potential cost reductions for nth-of-a-kind applications of the TVSA process and discourages the use of cost of capture as the sole metric to compare technologies beyond the short-time perspective. Recalling the results of the economic analysis showing that the current pricing for CDR credits advocates for offsetting for CO₂ concentrations below 1%, it is here concluded that this CDR price is aligned in terms of energy efficiency but not exergy efficiency, i.e. higher prices for CDR credits are required to incentivize the abatement method with the highest exergy efficiency (capture via MEA-ABS). In this regard, it is worth noting that the energy and exergy efficiency values are in relatively narrow ranges (3–8% and 5–22%, respectively); thus, the impact of efficiency on cost becomes pronounced only in cases of high energy prices or limitations on energy resources (detailed mass and energy balances and sizing are presented in Tables S7–S9).

4. Conclusions

This work compares the levelized costs of carbon capture (LCOC) and avoidance (LCOA) of three post-combustion capture technologies, the commercial benchmark MEA-ABS and the DAC-derived ALK-ABS and TVSA processes, applied to CO₂-lean (0.5–4.0%) industrial flue gases, and benchmarks the results against offsetting via purchase of CDR credits. Four main conclusions are drawn.

First, no single capture technology is preferable across the whole CO₂-lean flue gas window; instead, a technology-selection map emerges. From the LCOC perspective, MEA-ABS provides the lowest capture cost

for flue gas flow rates above approximately 200 t/h thanks to its monolithic economies of scale. The modular scaling of TVSA makes it the cheapest option at small-scales (below 200 t/h, e.g., small silicon plants and paint shops), while ALK-ABS is cost-competitive only in a narrow window (0.5–1.2% CO₂, 100–200 t/h) corresponding to, e.g., large-scale breweries, where the co-produced food-grade CO₂ may add further value.

Second, the ranking changes fundamentally once the carbon footprint of the energy supply is included. The avoidance cost of MEA-ABS and TVSA is highly sensitive to the carbon intensity of the heat supply, whereas ALK-ABS by virtue of its oxy-fuel-fired calciner with internal capture of the combustion CO₂ largely decouples its avoidance cost from the heat supply, yielding the lowest LCOA for the silicon, aluminium, and iron ore pelletization cases under the assumptions of this study. MEA-ABS attains the lowest LCOA only where low-carbon heat is available, as in the NGCC case with steam extraction, or where waste heat can be utilised. This underlines that the availability and carbon intensity of on-site heat, rather than capture cost alone, should guide technology selection for CO₂-lean sources.

Third, comparing the avoidance costs with current voluntary-market CDR credit prices (400–600 \$/tCO₂) maps the capture-or-offset decision: point-source capture is clearly more cost-effective than offsetting only for the NGCC case, and likely so for large-scale iron ore pelletization and aluminium production, although the overlap of the uncertainty ranges prevents a firm conclusion for the latter two. For small-scale sources exemplified here by silicon production offsetting through CDR credits is currently the cheaper route to net-zero direct emissions. Under the assumptions of this study, emission sources with CO₂ concentrations below ~1.2%, and sources in the 1.2–2.6% range occurring at small-scale (<400 t/h), can thus be characterised as economically hard-to-abate. These thresholds are specific to the scenarios analysed and are not intended as a general classification.

Fourth, the efficiency analysis provides a complementary, price-independent perspective: from an exergy viewpoint, point-source separation from any of the streams considered is more efficient than DAC, so none of the industrial cases is thermodynamically hard-to-abate. The divergence between the exergy ranking (favouring MEA-ABS and TVSA) and the FOAK cost ranking indicates that the current cost position of TVSA is driven by capital and current CDR market prices rather than by an intrinsic thermodynamic penalty, and cautions against using present-day capture cost as the sole long-term technology selection metric.

Finally, the conclusions of this work are valid under the assumptions adopted and reflect current cost conditions. The economic analysis relied on literature-derived techno-economic models and, for the DAC-derived technologies (ALK-ABS and TVSA), on first-of-a-kind cost estimates; costs realised in commercial-scale projects will additionally reflect detailed engineering, vendor quotations, financing and project development costs, contingencies, and commercial risk allocations. Both point-source CCS costs and CDR credit prices are expected to evolve through technological learning, the build-out of CO₂ transport and storage infrastructure, and the maturation of the CDR market, all of which may alter the relative competitiveness of capture and offsetting; the reported comparisons therefore represent current cost assumptions. Furthermore, the modelled LCOA and market CDR credit prices are quantities of a different nature: the former is a modelled cost estimate, whereas the latter are commercial prices that also embed project financing, business overheads, risk premiums, long-term liabilities, and monitoring and verification costs. Lastly, the analysis covers the three technologies investigated; emerging alternatives (e.g., pH-swing processes (Bouaboula et al., 2025b)) and future improvements to MEA-ABS and the DAC-derived technologies such as enhanced process integration, waste heat recovery, heat pump integration, or alternative process configurations (e.g., moving-bed TVSA) may change the techno-economic picture and constitute natural directions for future work.

Declaration of generative AI and AI-assisted technologies in the writing process

During the preparation of this work, the author(s) used ChatGPT in order to improve the language and readability. After using this tool/service, the authors reviewed and edited the content as needed and take (s) full responsibility for the content of the publication.

Nomenclature

ALK-ABS Alkaline absorption with subsequent calcium looping.
CCS Carbon capture and storage.
CDR Carbon dioxide removal.
DAC Direct air capture.
s-DAC Solid direct air capture.
LCOA Levelised cost of avoidance.
LCOC Levelised cost of capture.
MEA-ABS Absorption by monoethanolamine.
NGCC Natural gas combined cycle.
TVSA Temperature vacuum swing adsorption.

CRedit authorship contribution statement

Sina Hoseinpoori: Writing – review & editing, Writing – original draft, Visualization, Validation, Methodology, Investigation, Formal analysis, Data curation, Conceptualization. **David Pallarès:** Writing – review & editing, Supervision, Methodology. **Filip Johnsson:** Writing – review & editing, Funding acquisition.

Declaration of competing interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

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Supplementary materials

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Data availability

Data will be made available on request.

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