Partial carbon-capture by absorption cycle for reduced specific capture cost

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ABSTRACT

For a sustainable energy system, the industrial carbon emission should be zero – or close to. Partial capture of CO_2, i.e. capturing only a share of the CO_2, is discussed as an option to initiate the transition towards decarbonization of industry by reducing the CO_2 mitigation cost at industrial sites. This work models two approaches to achieve partial capture based on amine absorption – 1) capturing 90 % CO_2 from a split stream of the flue gas or 2) capturing less CO_2 (<< 90%) from the total flue-gas flow. A techno-economic analysis is carried out that considers scale, CO_2 concentration, and process configurations (absorber intercooling, rich solvent splitting) when comparing cost of partial capture to full capture, i.e. capturing close to all CO_2 from the entire gas. Besides lowering absolute costs, the study shows that partial capture from CO_2-rich gases may lower also specific cost (€ per tonne CO_2 captured) compared to full capture, despite economy of scale, during certain market conditions. Operating expenditures, especially the cost of steam, are found as dominating cost factor for partial capture even for capture down to 200 000 t per year.

Keywords: Partial capture, CCS, process industry, intercooling, rich-split, MEA, cost estimation

INTRODUCTION

Anthropogenic carbon dioxide emissions will have to be reduced drastically in order to limit global warming. To reach vast reductions and meet the Paris Agreements, emission intensive industries such as steel, cement or oil refineries typically will require to apply carbon capture and storage (CCS) [1,2]. Also, most integrated assessment models constrained to 2 °C do not converge without CCS [3]. CCS is associated with high energy requirement and investment costs. Hence, large-scale and cross-sector deployment of CCS is behind in expectations of an estimated 4,000 plants by 2030 required to meet emission scenarios [4] if compared to the approximately 20 plants currently in operation. An early implementation at large high-concentration point sources in process industry can be a way to lower specific investment costs [5,6], and thus facilitate the ramp-up in CCS deployment. The capture cost (excl. transport & storage) per tonne avoided CO_2 for *full capture* at a short to mid-term deployment of CCS in process industry has been estimated to be between 50 – 90 € for iron and steel, 160 – 310 € for cement-production, and 150 – 190 € for petroleum refineries and chemical sectors [7]. In this work, we define full capture as carbon capture applied to the entire flue-gas flow with an as high as possible capture efficiency, or capture rate, provided that a minimum in specific capture cost is reached. We suggest this definition of full capture to be focused on specific CAPEX (€/per tonne CO_2 captured) since CAPEX is a global cost, and less plant, location, and policy dependent than OPEX. In practice, the most cost-effective capture rate for full capture (including both CAPEX and OPEX) is found to range between 85 – 96% for coal power plants [8]. Most studies in academia on CCS in process industry commonly apply a capture efficiency of 90% for full capture.

An option to facilitate the implementation of CO_2 capture towards decarbonization of process industry is so called *partial capture*. We define partial capture as CCS concept where only a share of the accessible CO_2 from a flue gas or process gas is captured for storage with the extent of this share being governed by economic factors, such as energy prices, and policy-driven requirements. In all cases, partial capture reduces the absolute energy penalty and the required investment. The following examples illustrate in which scenarios partial capture may be feasible, i.e. preferable over full capture:

1) for plants that have access to low-cost energy (excess heat) or that can vary their product portfolio to meet time-varying market conditions [9,10] Capturing an amount of CO_2 equal to the available heat or when conditions are favorable may lead to a more cost-effective (CAPEX & OPEX) solution than full capture.

2) for plants or industry sectors that require to reach defined Emission Performance Standards (EPS); an example is the Clean Power Act in the U.S., which requires newly constructed super-critical pulverized coal power plants to capture approx.16 - 23 % of their CO_2 emissions [11], depending on coal type, to meet an EPS of 1,400 lb CO_2/MWh-g;

3) in order to reduce investment risks in CCS deployment for power generation from coal (PC & IGCC) [10,12] or from process industry;

4) for plants with multiple stacks – targeting the most suitable stack(s) instead of capturing from all, even diluted or remote stacks, which are less feasible [13];

5) for sites where the adjacent storage facility's capacity is insufficient to store all CO_2 emitted from the site during its lifetime [10];

6) for sites where partial capture can be used cost effectively in combination with other mitigation measures, such as fuel change, improvements in energy efficiency, and the use of biomass.

Figure 1 illustrates how partial capture could be a pathway towards full CO_2 reduction in process industry. From a systems perspective, partial capture could co-exist next to full capture CCS sites, bio-energy CCS sites and carbon-free/new technology sites. However, these mitigation options will not be deployed at the same time – new carbon-free technologies are often long-term technologies that are still in early-phase development (e.g. hydrogen based systems), whereas a short-term deployment of partial capture is possible today by applying off-the-shelf technology such as post-combustion capture using amine absorption, e.g. using monoethanolamine (MEA). Also, partial capture may evolve towards full capture on site via technology advancement such as solvent developments and de-bottlenecking of unforeseen flaws in design and operation over time (site specific learning curve) in case of first-of-a-kind plant.

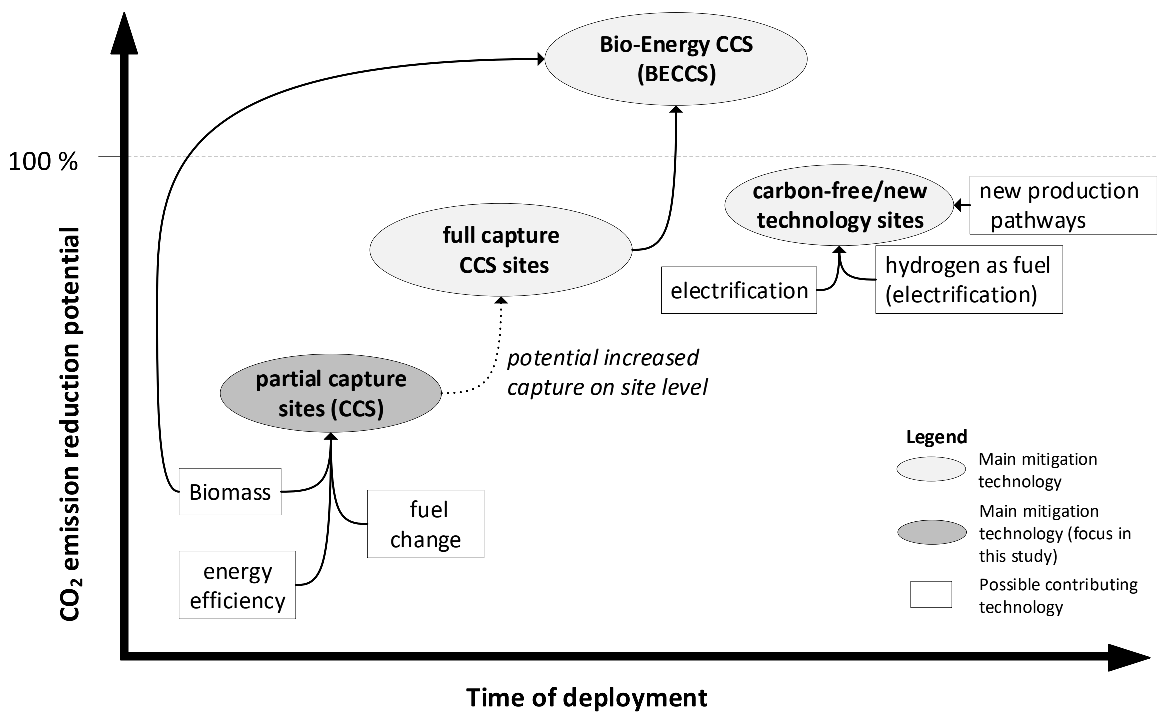


Figure 1. Contextualization of partial capture on industrial system level. Dotted arrows indicate possible development paths of partial capture.

This work focuses on partial capture motivated by economic considerations and cost of energy as explained in point 1) above. The work evaluates design considerations for a partial capture amine absorption cycle with respect to size and nature of the industrial CO_2 source with focus on continuously operated capture units with constant load. A generic study on two principle designs for partial capture is conducted in Aspen Plus that quantifies specific energy demand and capture costs in relation to the capture rate, which is varied between 30 – 97 %. The study considers the impact of CO_2 concentrations, flow rate of the CO_2

source as well as the effect of process configuration on the partial capture designs. Cost results, based on Aspen Cost Estimator and an individual detail factor method, are compared to a reference design for full capture (assumed 90% capture rate) from the entire flow of a CO_2-rich industrial flue gas (single stack site). Finally, an analysis on the balance of CAPEX and OPEX in dependence of scale and CO_2 source is carried out to map and demonstrate the viability of partial capture as cost-efficient mitigation measure for process industries.

FULL AND PARTIAL CAPTURE

Reflections on full capture

Most studies on amine-based CCS follow the approach of full capture, as defined above. In this section, an example on full capture is given for a coal fired power plant to illustrate the use of specific CAPEX (€/t CO_2 captured) to determining a cost-optimal capture rate. Figure 2 shows an estimation of specific CAPEX depending on capture rate for Nordjyllandsverket, a coal power plant (410 MW_e) in northern Denmark. For a single-train absorption cycle, the minimum specific CAPEX is reached at around 90% capture rate and 90% capture rate is, thus, defined as full capture for this specific plant.

In their study on CCS from pulverized coal (PC) power, Rao and Rubin [8] argue for optimizing the capture rate towards cost-effectiveness measured in cost of CO_2 avoided and find capture rates of 81% and 87% representing minimum cost for multi-train absorption units for PC plants with 1000 MW_e and 650 MW_e , respectively. They stress plant design, choice of solvent, maximum absorber train size, and plant size as most influential parameters. For industrial sites in particular, the cost of CO_2 avoided, especially OPEX, is industry and site specific [9] and depends on raw material flows, such as crude oil, limestone or iron ore, and energy flows, mainly governed by fossil fuel type and electric energy system. These flows also impact the carbon balance in dependence of how the battery limits are defined [14]. As suggested in the introduction, defining the capture rate for full capture by the minimum specific CAPEX only makes the result less dependent on market, policy, site and industry specific variations.

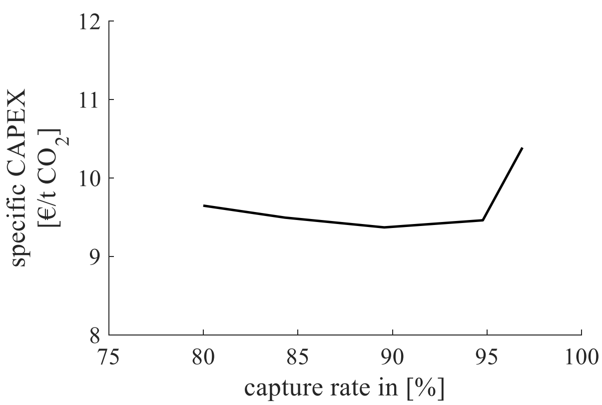


Figure 2. Specific CAPEX per tonne CO_2 captured in single train absorption cycle in standard configuration (no process modifications) dependent on capture rate – illustrates the definition of full capture. Cost estimation based on the method in this work, battery limit around MEA capture unit, capture from gas flow of Nordjylland coal power station (370 kg/s with 13 % CO_2). Note that the ordinate does not start from 0.

Design paths for partial capture

For a single-stack*, there are two *design paths* to implement partial capture: The *slip stream path* (SSP) - separation of CO_2 from a *slip* stream with a high CO_2 separation in the absorber (e.g. 90%), or *separation rate path* (SRP) - separation of a lower fraction of CO_2 (significantly less than 90%) from the *full* stream. To give an example, Figure 3 illustrates the two design paths with capture of 45 % of the stack emissions – which is half the amount of CO_2 captured in the reference full capture (90 % of the CO_2 from the full stream) shown in Figure 3b. For the SSP, see Figure 3a, 50 % of the stream is fed to the capture unit. The SSP design is, thus, merely down-scaled from full capture and has the same specific heat demand (MJ/kg CO_2), same liquid-to-gas ratio and same *separation rate*, i.e. fraction of CO_2 removed from the feed gas entering the absorber. The change in investment relative full capture is, thus merely determined by the economy of scale. The SRP design path for partial capture, see Figure 3c, resembles that of an oversized absorber since the entire stream of CO_2-rich gas has to pass the column. The liquid-to-gas ratio and the absorber are designed in such a way that only a fraction of the CO_2 is separated – in the example the *separation rate* of the absorber is 45 % in order to capture the same amount of CO_2 as the SSP design. The reduced liquid-to-gas ratio reduces the size of the remaining equipment, which contributes to lower CAPEX. Similarly to SSP, the SRP is expected to have higher specific CAPEX (€/t CO_2) than full capture. The over-sized absorber gives a higher rich loading and reduced temperature difference in the cross-flow heat exchanger, which reduces the reboiler heat duty and lowers the specific heat demand (MJ/t CO_2), relative full capture [15]. Since the heat demand is a key driving force in operating costs, the SRP potentially lowers operating costs and total cost as shown in a recent partial capture study by Øi et al. [16]. In scenarios where OPEX dominates the capture costs, partial capture may lower the specific capture costs (€/t CO_2).

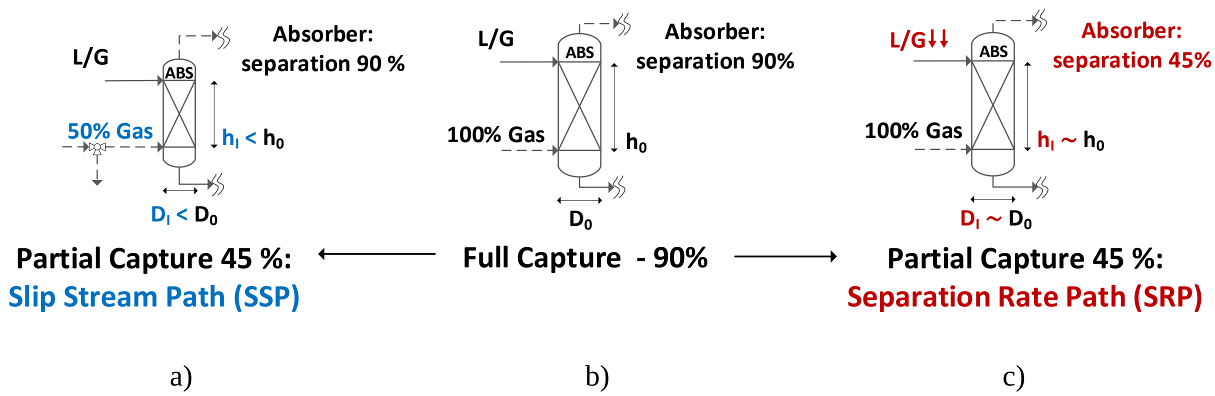


Figure 3. Design paths for partial capture a) and c) from a designated CO_2 source compared to full capture b); simplified process scheme (only absorber shown; dimensions of remaining equipment are adapted as well). Indices '0' refer to full capture reference dimensions, indices '1' refer to partial capture design dimensions.

* These design paths are also valid for a multi-stack facility (not shown for simplicity)

PROCESS CONFIGURATIONS FOR PARTIAL CAPTURE

In literature, various configurations of the amine capture process to reduce the energy penalty caused by the temperature swing for regeneration of the solvent have been proposed for full capture (predominantly based on modelling) [17–22]. To match the scope of this work, namely near-future deployment of partial capture, rather simple, easy to operate, non-capital intensive, four well-known modifications of the original process are considered for each of the two partial capture design paths SRP and SSP described above. Figure 4 shows flow diagrams of a) the simple absorption cycle, b) *intercooled absorber (ICA)* or intercooling/inter-stage cooling, and c) *rich solvent splitting (RSS)* or rich split, and d) the combination of the ICA and RSS configurations. In the following paragraphs, a short literature review on ICA and RSS for full capture is given.

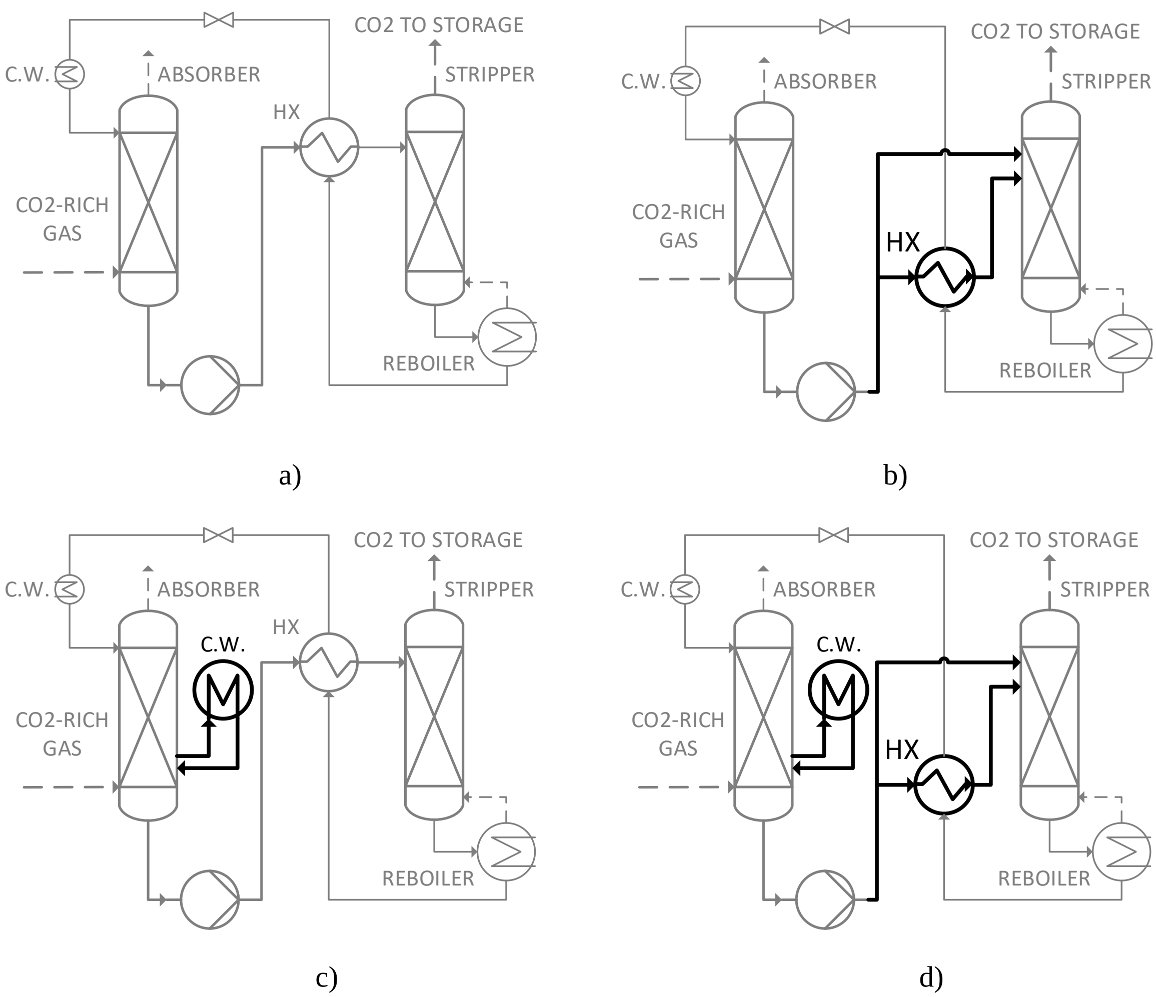


Figure 4. Process flow diagrams for amine CO_2 capture evaluated for partial capture: Standard configuration (a), Rich solvent splitting configuration (b), Intercooled absorber configuration (c), Rich solvent splitting and intercooled absorber configuration (d)

Cooling the liquid phase in the absorber (ICA, Figure 4c) reduces the solvent temperature and causes a shift in gas-liquid equilibrium (the CO_2 partial pressure at equilibrium is lowered leading to a larger CO_2 partial pressure difference between gas bulk phase and local equilibrium[23]). For gases with high CO_2 content, this is in general beneficial for the absorber outlet CO_2 loading capacity in the solvent - despite slightly reduced mass transfer rates at lower temperatures. The increased loading capacity means less solvent recirculation, which reduces the reboiler duty. The influence of ICA on reboiler duty for MEA regeneration from power plants (approx. 12 – 14 vol.% CO_2 in the flue gas) have been investigated: Knudsen et al. could not find any obvious advantage of intercooling in experiments [24], whereas modelling studies found reductions in reboiler duty of 2.84 %[25] and 6.4% [26]. A model by Li et al.[27], validated against pilot plant data, calculated savings of 1.8 %. Garðarsdóttir et al. [23] state the importance of CO_2 concentration in the feed gas and reported 4.2 % and 9.3 % savings in reboiler duty when applying intercooling for feed gas CO_2 concentrations of 30 mol.% and 40 mol.%, respectively. Intercooling is used in commercial processes like Fluor's Ecoamine FG Plus [28] (based on MEA) or Shell's CANSOLV CO_2 capture process [29]. The implementation of ICA requires an additional heat exchanger, an extra pump, and piping and instrumentation. In terms of operability, the intercooling modification can be by-passed [17] in case of failure caused by a trip in the pump or heat exchanger.

Rich solvent splitting (RSS, Figure 4 *b*), improves the efficiency of the stripper [30]. In the simple absorption cycle, the warm rich solvent enters the stripper column at the top, and the vapour released leaves the stripper at a high temperature. Applying RSS, the rich solvent is split prior to the heat exchanger, resulting in a cold and warm stream. The cold stream is introduced at the top, meeting the vapour from the warm stream, which enters the column a few stages below. In this way, the vapour leaves the column at a lower temperature, leading to increased efficiency and a decreased reboiler duty by 7 – 10 %[17,26,27,31]. The implementation requires no additional equipment except for a split valve, its instrumentation and additional piping. Concerning operability, Le Moullec et al. [17] assessed RSS as non-critical to process performance, i.e. its installation does not increase the complexity of the capture unit. It rather improves operability as the valve represents an additional control variable which can smooth out process instabilities.

MODELLING

Process modelling and simulation of partial capture

The amine absorption process using a 30 wt.% monoethanolamine (MEA) aqueous solvent is simulated in Aspen Plus (V 8.8). The process model is based on the work of Garðarsdóttir et al. [23]. It uses the Bravo et al. [32] correlation from 1985 to predict mass transfer and interfacial area in the rate-based modelling of the structured packing in absorber and stripper. The correlation by Bravo et al. [33] from 1992 is used for liquid hold-up calculations. All capture designs in this study were simulated in design mode, i.e. equipment was sized accordingly for each capture rate. Also, each capture design was optimized towards minimum specific heat demand by varying the L/G ratio at a targeted capture rate.

In order to benchmark the partial capture concepts, a reference full capture (90 % separation rate at full feed gas flow) design is characterized based on the standard configuration, see Figure 4*a*. A generic CO_2-rich gas (called REF) resembling CO_2 sources from process industry is selected as feed. It is specified with a CO_2 concentration of 20 vol.% and a flow rate of 200 kg/s at 150 °C and pressure of 1 bar. Similar CO_2 concentrations are reached in lime kilns (Kraft process), in steam methane reformers (refineries), in hot stoves/ power plants (integrated steel mills) or in combined stacks for cement production [6]. The specifications of the full capture reference design are shown in

Table 1. The simulated holdup of the liquid phase in the absorber packing (residence time 2.4 min) is in the range of reported values of pilot plants (1.6 – 2.8 min; [34]), and in the same order of magnitude as what is representative for full-scale plants (4 – 14 min; [35,36]).

The partial capture design paths SSP and SRP (see Figure 3) are rendered from the full capture reference design by either decreasing the flow rate of the feed gas, $\dot{m}_{\mathrm{feed}}$, (SSP) or the circulation rate of the solvent, $\dot{m}_{\mathrm{lean}}$, (SRP) [8]. The separation rate of the SSP in the absorber is held at 90%, whilst for SRP it decreases with the solvent circulation. To allow for a comparison of the partial capture designs in capture performance and cost, several design parameters are held constant (see lower half of

Table 1). Note that the *gas* residence time in the absorber packing $\bar{\tau}_{\mathrm{hyd,abs}}$, is set to 13 seconds in all cases to define the absorber packing dimensions in relation to the feed gas flow according to Eq.(1).

$$\bar{\tau}_{\mathrm{hyd,abs}} = \frac{H_{\mathrm{pack,abs}} \cdot D_{\mathrm{abs}}^2 \cdot \pi}{4 \cdot \frac{\dot{m}_{\mathrm{feed}}}{\rho_{\mathrm{feed}}} \cdot z} \quad [\mathrm{s}] \tag{1}$$

with:

$H_{\mathrm{pack,abs}}$	Height of absorber packing material
D_{abs}	Absorber diameter; assumed cylindrical geometry
$\dot{m}_{\mathrm{feed}}$	Mass flow of feed gas into absorber
ρ_{feed}	Density of feed gas into absorber
ρ_{REF}	Density of reference feed gas REF
$z = \frac{\rho_{\mathrm{feed}}}{\rho_{\mathrm{REF}}}$	Correction factor to compensate for changes in density in case of variation of feed gas CO_2 concentration

Table 1. Design parameters for 30wt.% MEA partial capture designs: Upper half: valid for full capture reference design only. Lower half: valid for all designs (held constant if not explicitly mentioned otherwise).

Design parameter		unit	value
specific to full capture design	Liquid-to-gas ratio	kg/kg	5.2
	absorber height	m	20
	absorber diameter	m	11.9
	stripper diameter	m	7.9
	liquid hold up in absorber packing	min	2.4
general to all designs in this study	lean solvent loading	mol/mol	0.28
	flooding factor	-	0.8
	gas residence time in absorber packing	s	13
	packing material	SULZER	MELLAPAK Y250
	absorber pressure	bar	1
	cross heat exchanger temperature difference	°C	10
	stripper pressure	bar	2
	max. Stripper sump T	°C	122
	stripper height	m	15
	absorber inlet temperature (liq/gas)	°C	40
	stripper head-product condenser temp.	°C	20
	CO_2 pressure after compression	bar	110
	max. MEA slip after washer	ppm	1

For the separation rate design path (SRP), a constant gas residence time, $\bar{\tau}_{\mathrm{hyd,abs}}$, implies that the feed gas (100% of the reference) always sees the same volume of packing material [**] when varying the solvent circulation rate to obtain different separation rates. For the slip stream design path (SSP), the constant gas residence time $\bar{\tau}_{\mathrm{hyd,abs}}$ allows for scale-down from full capture when reducing the feed gas flow rate to obtain different capture rates while keeping the separation rate in the absorber at 90 %. In detail, the reduced feed gas flow rate is matched by a lower column that together with a constant L/G ratio gives the same mean residence time for the liquid phase – a criteria for packed column scaling [37].

The intercooling, in the ICA configuration (see Figure 4c), is modeled as pump-around with a re-entry temperature of 40 °C. The draw position (stage) is varied for optimization – however, it is generally found in the bottom part of the absorber (roughly at ¼ to ⅕ of the packing height) in accordance with reported optimum locations [25]. The RSS configuration is incorporated in the Aspen Plus model as stream splitter block – the split ratio f_{RSS} is defined according to Eq. (2) as fraction of the hot rich stream $\dot{m}_{\mathrm{RICH,hot}}$ and the total rich stream $\dot{m}_{\mathrm{RICH,tot}}$. A constant value for f_{RSS} of 0.8 is used unless stated otherwise. The cold rich stream is fed into the stripper at the top – one stage below the condenser reflux. The feed stage for the hot rich stream is stage 5 (of 20) unless stated otherwise.

$$f_{\mathrm{RSS}} = \frac{\dot{m}_{\mathrm{RICH,hot}}}{\dot{m}_{\mathrm{RICH,tot}}} \quad [\text{-}] \tag{2}$$

Cost estimation

The cost estimation for the MEA based capture unit comprises capital expenditures (CAPEX) and operational expenditures (OPEX). The partial capture processes are compared in their capture cost, € per captured tonne of CO_2. The technology maturity is assumed to be high and the CAPEX reflect an "*n*th-of-kind" (NOAK) approach. The plant is assumed to be installed at an existing site (retro-fit). Figure 5 gives an overview of the system and represents the detail of the study, i.e. the equipment included in the cost estimation. Minor heat exchangers, pumps and other utilities, e.g. for the washer unit, have been neglected since it should be a fair assumption that these will have little influence on the results. CAPEX for each piece of equipment is estimated with Aspen In-Plant Cost Estimator and multiplied with an individual installation factor. The CAPEX of the reclaimer unit was estimated from the IEAGHG report on reclaimer and sludge disposal [38]. Empirical formulas from an in-house collected industry cost database [39] are used to determine the installation factors, which account for type and size of individual equipment and its installation depending on the type of site. In addition, a contingency of 20 % is included in CAPEX. It is further assumed that all equipment, exempt major vessels such as tanks and columns, is placed in non-insulated buildings. Not included is the purchase of land, piling or costs for secondary buildings. This method of CAPEX estimation has normally an uncertainty of ± 40% (80% confidence interval).

[**] the packing height is used as degree of freedom in a Design Spec in Aspen Plus; the absorber diameter is governed by the hold-up correlations and the flooding factor

The underlying financial assumptions for the cost estimation and utility costs are given in Table 2. The cost of steam has the highest impact on the capture cost, next to plant life time and rate of return [6]. Therefore, this work includes a cost range for the steam cost, besides an experience based reference value. The lower end of the span, 2 €/t, represents cost of steam from waste heat recovered [40] whereas the higher end of the span, 25 €/t, represents steam from a natural gas fired boiler (adapted from [40]).

Table 2. Assumptions for the cost estimation

parameter	unit	value
project life time	yr	25
construction time	yr	2
discount rate	%	7.5
cost year		2015
yearly operation	h/a	7000
maintenance (annualized; % of investment)	%	4
labor (6 operators, 1 engineer)	kEUR/a	821
location	-	generic location (Rotterdam)
electricity	€/MWh	55
cooling water	€/m^3	0.022
MEA	€/m^3	1867
NaOH	€/t	370
sludge disposal	€/t	190
steam (reference value)	€/t	16.7 ***

*** Price obtained from a large Swedish process industry

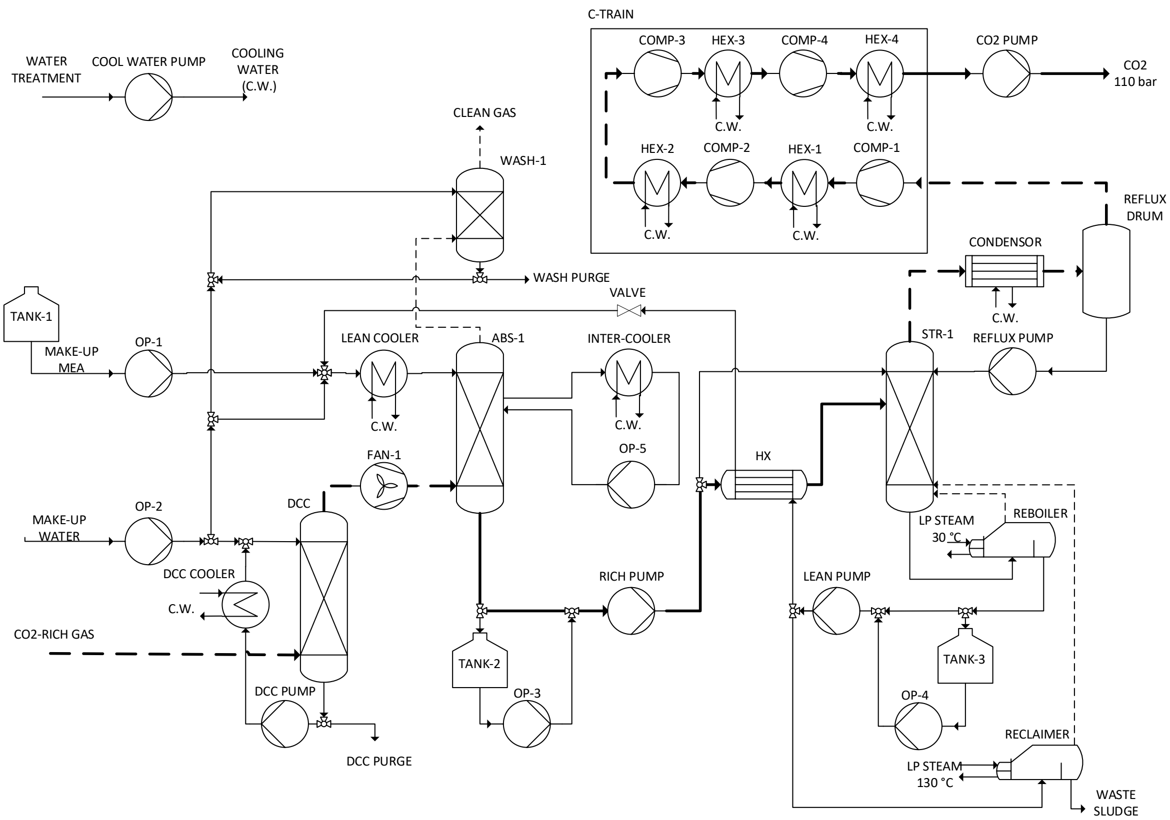


Figure 5. Illustration of process flow diagram of capture unit. Included equipment represents battery limit valid for cost estimations in this study.

RESULTS AND DISCUSSION

The effect of a reduced CO_2 capture efficiency on the design of the capture process is evaluated with respect to the technical and, subsequently, to the economic performance.

Technical performance of partial capture

Figure 6 compares the specific heat demand of partial capture of the SSP and SRP. Per definition, the specific heat demand of the SSP is the same as for the reference design and is not discussed further. The heat demand of the SRP path drops with decreasing capture rate. At a capture rate of 45 % the heat demand is 12.3 % lower than in the full capture reference design.

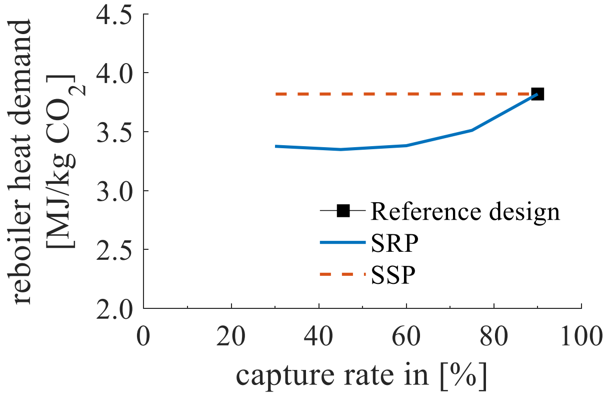
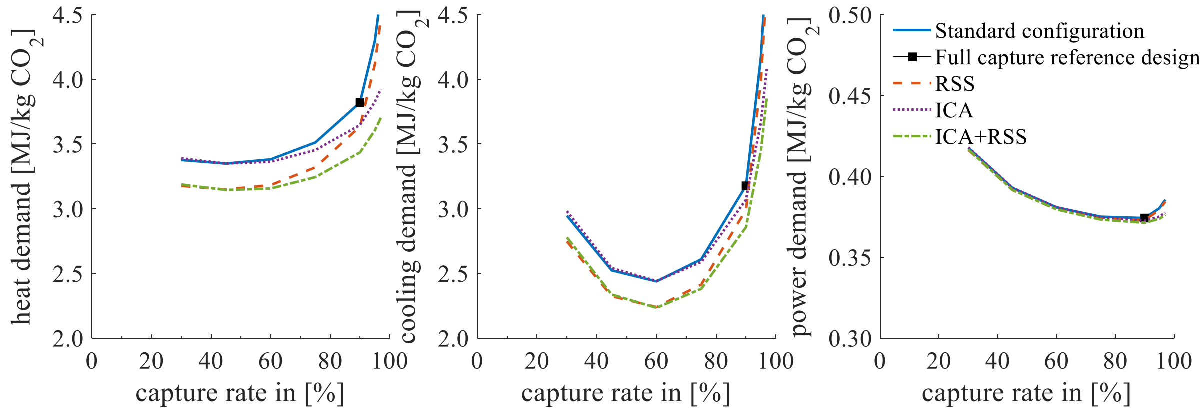


Figure 6. Specific heat demand for partial capture design paths relative to a reference full capture (90%) design: Separation Rate Path (SRP) and Slip Stream Path (SSP). Capture rate refers to total CO_2 in feed gas REF (20 vol.% at 200 kg/s). Note that the ordinate does not start from 0.

Process configurations for SRP partial capture. The heat demand, cooling requirement, and power demand of the standard, RSS, ICA, ICA + RSS configurations are shown in Figure 7. The more advanced cycles outperform the standard configuration in terms of specific heat demand for all separation rates. The combination of ICA+RSS has the lowest energy demand with a reduction relative to the standard configuration ranging from 6 % (45% capture) to 21% (97% capture). As could be expected, the RSS and ICA improves different parts of the process and may be super-imposed. Relative to the standard configuration, the RSS is increasingly beneficial at lower separation rates, while the ICA has a similar effect for separation rates below approximately 60 %. The cooling demand has a minimum at roughly 60 % capture. The RSS reduces the required duty of the condenser downstream of the stripper moderately - 3-8% compared to the standard configuration. The cross-heat exchanger becomes smaller with a higher hot-outlet temperature compared to the standard configuration, yet similar cold-outlet temperature (lean solvent). The implementation of intercooling can save 0-17% in cooling demand, with larger savings at high capture rates, as cooling in the absorber is more effective than cooling of the lean solvent for these separation rates. Again, the combination of ICA+RSS demonstrates the greatest potential. Changes in the power demand are less prominent – roughly within the range of 0.04 MJ/kg CO_2. All cycles require more electricity per kg capture CO_2 at lower separation rates, as the SRP passes the entire feed gas through the absorber maintaining the same power demand in the flue gas blower while capturing less CO_2. Deviations above separation rates of 90% are explained by larger cooling water flows that have to be pumped.



a) b) c)

Figure 7. Specific energy demands for process configuration SRP – heat demand (a), cooling requirement (b) and power demand(c). Note that the ordinate does not start from 0.

Limits for intercooling on SRP. The ICA modification is not required at low separation rates (below 60% capture in Figure 7). The reason for this is the significant change in the absorber temperature profile during partial capture. For example, Figure 8 gives the absorber temperature profile for capturing 45% compared to full capture. The partial capture case shows roughly 20°C lower peak temperature. This marginalizes the temperature reduction by ICA for SRP partial capture, as indicated by the proximity of the two dotted lines (i.e. with/without ICA) in Figure 8. This is in agreement with the findings of Kvamsdal and Rochelle [41] that showed that for low L/G, the enthalpy of the absorption reaction mainly increases the temperature of the gas stream, lowering the maximum temperature and moves it higher up in the column. In Figure 9, the relation between maximum temperature and L/G is shown for the standard configuration (20% CO_2) in comparison to the results from Kvamsdal and Rochelle (17 % CO_2). Note the change in L/G from full capture (■) to partial capture (●) with SRP design. As reported in [23], the CO_2 concentration also has a significant effect on the maximum temperature in the absorber. Hence, ICA is most beneficial, at moderate/high CO_2 concentrations and high separation rates.

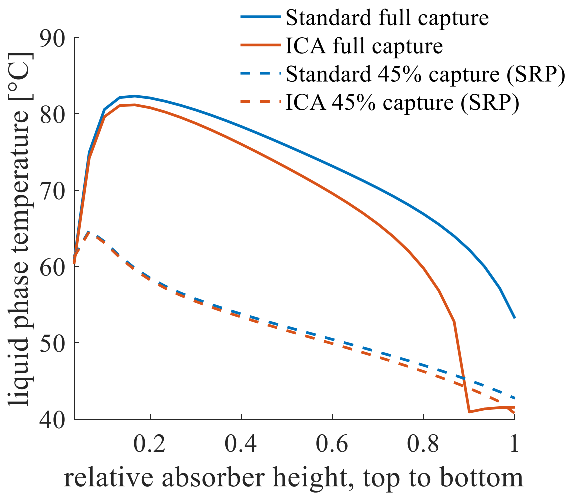


Figure 8. Absorber liquid temperature profile for full (90%) and partial (45%) capture (SRP).

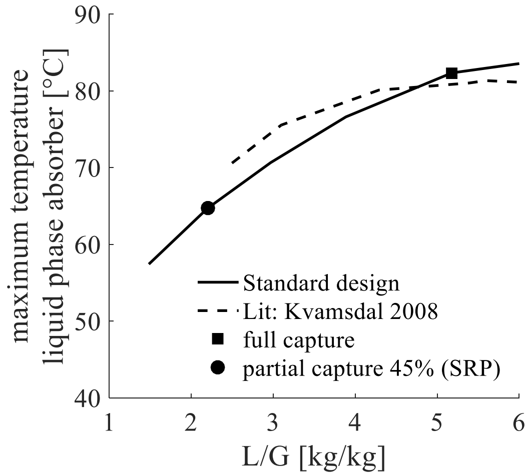


Figure 9. Liquid phase peak temperature in the absorber in dependence of liquid-to-gas ratio using standard design. Dotted line represents experimentally validated results from Kvamsdal and Rochelle [41] with 17% CO_2 in flue gas

Lean loading impacts rich-split on SRP. The positive effect of the rich-split (RSS) configuration shown in Figure 7 is based on a constant lean loading of 0.28. However, the benefit of RSS configuration is sensitive towards lean loading, as shown in Figure 10. A minimum in required heat (Figure 10 a), was found for a lean loading of 0.28 and 0.30 for full capture and partial capture (45%; SRP), respectively. In (Figure 10 b), the reduction potential for RSS compared to the standard configuration is highest for 0.30 lean loading. For 0.25 lean loading, there is no significant benefit. In fact, RSS shows a negative impact when feeding the hot rich below stage 5. For all lean loadings simulated, the partial capture SRP design shows a higher reduction in heat demand compared to full capture design. This is likely due to higher rich loadings in the solvent entering the stripper.

Influence of CO_2 concentration. The impact of CO_2 concentration on the heat demand of full capture has been discussed in literature, see for example [23,42]. Figure 11 shows the impact of CO_2 concentration depending on capture rate. The SSP (standard configuration) follows the performance of full capture, as it merely is downscaled. The SRP is favored by an increased CO_2 concentration more than the reference. The heat demand for 45 % and 67 % capture rates decreases until around 30 vol.%. For high CO_2 concentrations, e.g. 28 vol.%, capturing 45 % with an SRP design can reduce the specific heat demand by 19 %. The reason for this reduction in heat demand are absorber profiles with lower bulge temperatures experienced for partial capture SRP, as illustrated in Figure 8. Hence, process industries with high CO_2 concentrations would benefit from partial capture (SRP), because lower peak temperatures in the absorber lead to an enhanced absorption and, thus, more efficient capture than for full capture at these concentrations.

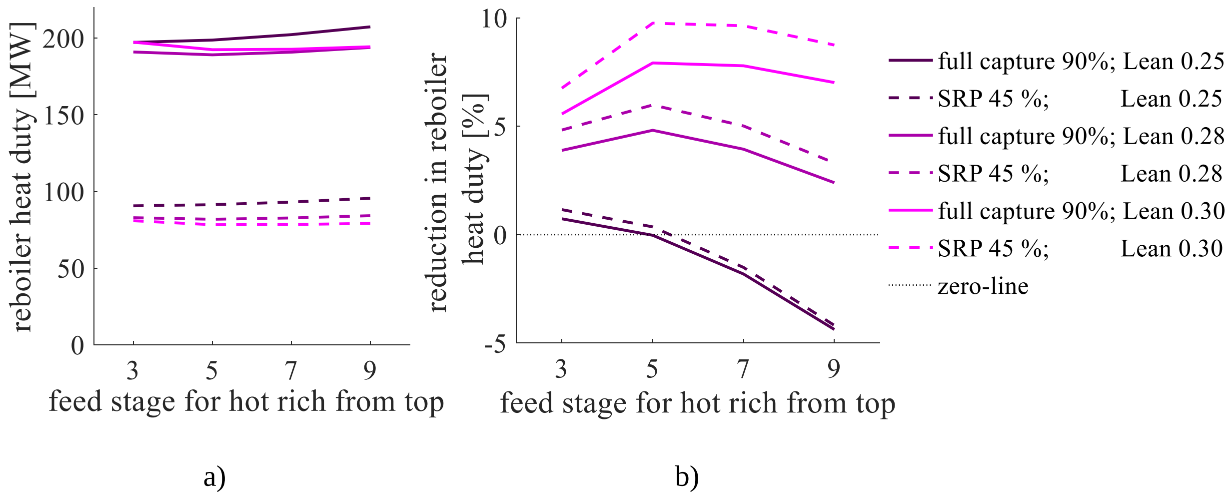


Figure 10. reboiler heat demand in rich split configuration (RSS) for full capture and partial capture (SRP) in dependence of feed stage for hot rich solvent in stripper. a) absolute values in MW; b) reduction potential in comparison to standard configuration.

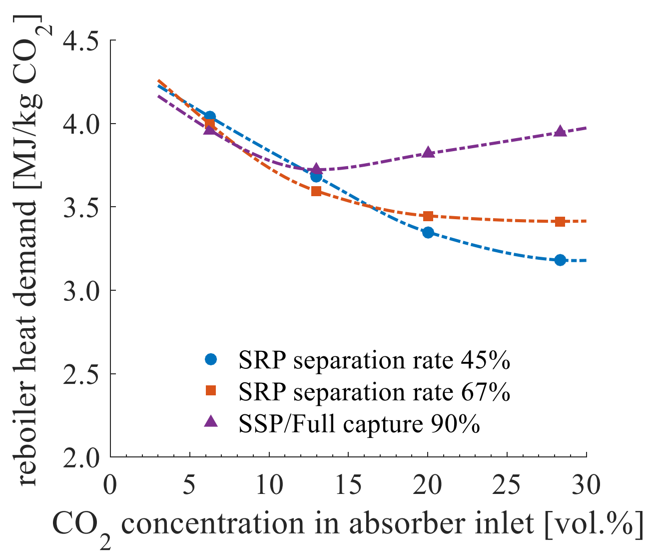


Figure 11. Influence of partial capture on reboiler heat demand in standard configuration for feed gas CO_2 concentrations between 6 – 28 vol.% at 200 kg/s. Note that the ordinate does not start from 0.

Cost of partial capture

We first present the analysis of the investment costs, then we discuss the total capture costs including OPEX, focusing on the impact of flue gas CO_2 concentration, flue gas flow (scale), and pricing of heat. It should be noted that these parameters depend on each other, i.e. changing one parameter in the model influences the others, and only one of these parameters was changed at a time.

Investment costs. Figure 12 presents the total and specific CAPEX required relative to the amount of CO_2 captured. The reference is a capture plant for 1.4 Mt/a that requires an investment of around 120 M€. The total CAPEX for the capture unit, which to a large extent determines the risk of investment, are obviously reduced with lower amounts of CO_2 captured (Figure 12a). When capturing 50 % compared to full capture, the capture costs drop by 36 % and 39 % for SRP and SPP, respectively. Both design paths show similar reductions in CAPEX; although the cost of the SRP design is slightly higher when capturing small amounts. Due to the economy of scale, the specific CAPEX (€/tCO_2) increases from roughly 10 €/t CO_2 for full capture to 13 €/t CO_2 and 15 €/t CO_2 for SSP and SRP, respectively, when capturing only 30 % of the CO_2 emissions. The change in equipment cost for the SRP and SSP paths from full to 60 % capture is given in Figure 13. Both partial capture designs have similar costs for compression, pumps, and other vessels and filters. The SRP design has higher costs for the absorber and the DCC, which are similar to the full capture design, since the entire feed gas enters the column and is set to have the same residence time. The increased column costs for SRP are off-set by reduced reboiler costs, which is caused by the lower specific heat demand for SRP (see Figure 6). In addition, SRP has slightly lower cost for pumps and other vessels such as the reclaimer due to the reduced L/G ratio (see Figure 9). Overall, the largest savings for partial capture is the heat exchangers, foremost reboiler and cross-heat exchanger.

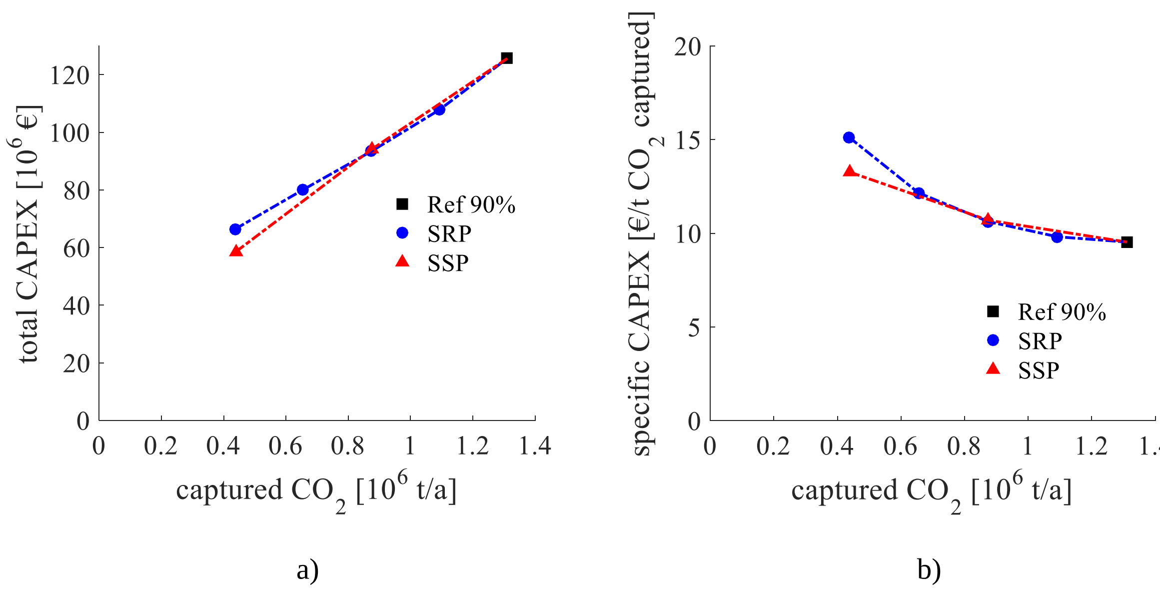


Figure 12. total (a) and specific(b) investment costs of partial capture with SSP or SRP design in comparison to full capture (Ref 90%); standard process configuration

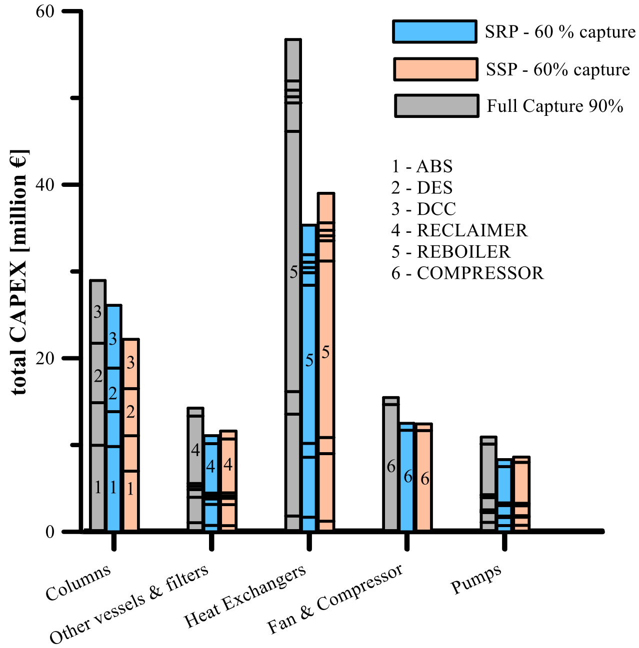


Figure 13. Detailed equipment costs for partial capture design paths SRP & SSP at 60% capture rate in comparison with full capture. Standard process configuration. Notation: 1- Absorber; 2 – Desorber; 3- Direct Contact Cooler; [4-6 as shown]

Total capture cost. The total capture cost comprise both CAPEX and OPEX. In this section, we highlight aspects that influence the total capture cost per captured tonne CO_2 for partial capture.

Process modifications. Figure 14 presents the specific total capture costs for the process configurations of this work. The lower specific heat demand of the SRP design (Figure 7) at reduced capture rate favors partial capture while the lower fixed costs favors full capture and a minimum, highly dependent on the assumptions, in total specific cost is seen. With a steam price of 17 €/t, capturing 60% with SRP design in standard configuration reduces the capture costs by 4 €/t CO_2 captured (~6.4 % reduction) compared to reference design. At separation rates below 60%, the increase in CAPEX dominates over the reduction in OPEX (decreased heat demand) and the total capture costs increases for partial capture with SRP design. The RSS configuration is cost effective throughout the studied span of separation rate with roughly 2.9 % reduction of the specific cost compared to standard configuration at the same separation rate. Compared to the full capture reference design, RSS at 60% separation can save 9.3 % in specific capture cost. The ICA configuration is only cost-effective for a separation rate >75 %. ICA can deliver a reduction in specific cost of up to 2.7 % at full capture compared to standard configuration. The combination of ICA + RSS yields reduction in specific cost by 5.4 % at 90% separation, and is a cost-effective modification for SSP design or full capture design.

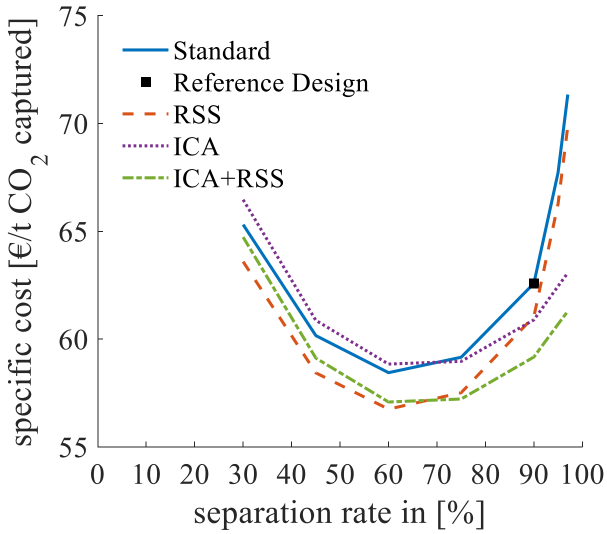


Figure 14. Specific cost (OPEX + CAPEX) of partial capture (SRP) from REF gas in dependence of separation rate for different process modifications. Note that the ordinate does not start from 0.

Influence of CO_2 concentrations. Figure 15 presents the impact of CO_2 concentration on total specific cost. As expected, specific capture cost for full capture decrease with higher CO_2 concentrations. At low CO_2 concentrations, the specific cost of partial capture (SRP design) increases more than for full capture. In the example shown, the SRP design gives a lower specific cost for concentrations above 17 vol.% and 13 vol.% for separation rates of 45 and 67 %, respectively. The SSP design (not shown) follows the trend of the full capture curve, although with higher cost, since it has the same specific heat demand as full capture (Figure 11) but higher specific CAPEX. Note that the example illustrates capture from a large source (200 kg/s) and the result is sensitive to scale.

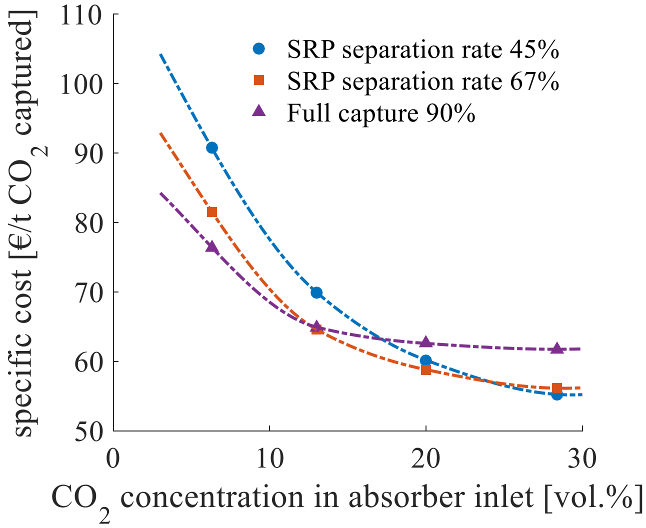


Figure 15: Specific cost (CAPEX + OPEX) of partial capture (SRP) in comparison with full capture; standard configuration for feed gas CO_2 concentrations between 6 – 28 vol.% at 200 kg/s. Steam price 17 €/t. Note that the ordinate does not start from 0.

Influence of scale. In Figure 16, the total specific capture cost for partial capture is shown in dependence of the flue gas flow and the annually amount of CO_2 captured. The cost for SSP/Full capture (▲) is shown for three flue gas flows (200 kg/s, 133 kg/s and 67 kg/s). A SRP curve (blue) is included for each of this flue gas flow with a blue marker indicating the minimum specific cost. Above the annual captured amount of approx. 0.47 Mt/a (36 % capture rate), there is at least one of the SRP curves that shows lower capture cost than the full capture reference. The merit of SRP over SSP (distance to the SSP curve) decreases for low amounts of CO_2 captured - for a gas flow of 200 kg/s the capture cost for SRP is 9% (@0.7 Mt/a) less than the SSP design, whereas for a third of the gas flow (67 kg/s) the merit is only 6% (@0.3 Mt/a) - as the relation between OPEX and CAPEX changes with the amount of CO_2 captured, and thus reduces the benefit of SRP over SSP because of its lower specific heat demand. Besides illustrating the impact of scale, Figure 16 reveals the economic potential in combining SSP and SRP design paths. Figure 17 demonstrates the ratio of OPEX/CAPEX declining from 5.6 for REF to 3.3 and 4.2 for SRP and SSP, respectively. Compared to the findings here, early works in literature on partial capture from coal power generation find a lower cost for the SSP design path compared to SRP[43,44]. However, literature gives no results which can be used for a direct comparison between the design paths for the same capture rates in a similar manner as our study. The likely reason for the SSP design path yielding lower cost than the SRP path, is that these studies[43,44] show smaller ratios of OPEX/CAPEX, possibly caused by a first-of-a-kind approach and differing assumptions such as, for example the cost year and CO_2 pressure after compression. Low ratios of OPEX/CAPEX may prevent OPEX savings for SRP from being significant enough to outperform SSP in cost. Also, in the above works, CO_2 concentrations in flue gases from coal power were not varied (maintained within 13-14 %). Thus, the advantage of SRP in saving reboiler heat (confer Figure 11 and

Figure 15) for process industry (e.g., cement, steel, steam methane reforming, Kraft pulping) with higher CO_2 concentrations of ca. 20% or more, has not been considered in the above works.

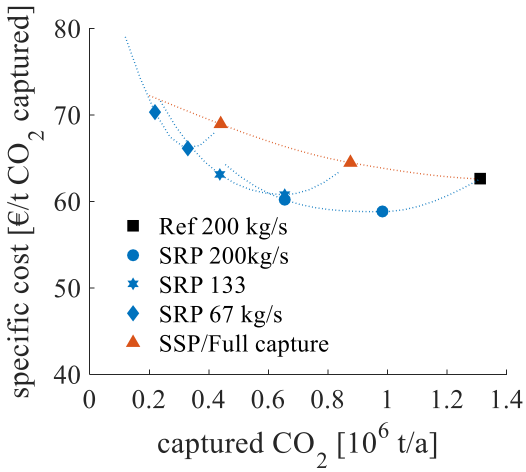


Figure 16. Specific cost depending on annually captured amount of CO_2; standard configuration 20 vol.% CO_2; steam price 17 €/t. Note that the ordinate does not start from 0.

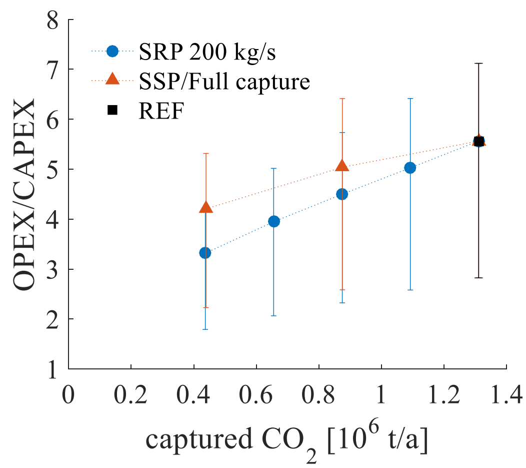


Figure 17. Relation between OPEX and CAPEX depending on annually captured amount of CO_2; standard configuration, 20 vol.% CO_2; the error bars represent a steam price of 2 €/t and 25 €/t for the lower and upper bound, respectively.

The price of steam. The bars in Figure 17 demonstrate the impact of steam price on the OPEX/CAPEX ratio. The steam price has a large impact on the capture cost and also on the relation between OPEX and CAPEX. The same span in prices, namely 2 €/t for steam from recovered waste heat and 25 €/t for steam from a natural gas fired boiler, is applied in Figure 18, where the detailed capture cost are shown for SRP and SSP (both 60 % capture rate) and the full capture 90 % reference. Therein, the span for steam cost contributes to capture cost from less than 5 €/t CO_2 up to well above 40 €/t CO_2. This illustrates how important it is to consider industry and site-specific conditions for heat supply when estimating costs. The diagram also shows, that it is the lower price for steam due to lower heat demand that makes the SRP design the more economic option for partial capture. In addition, the lower L/G ratios (see Figure 9) reduce the cost for chemicals & waste and the cooling cost. Worth mentioning is that in this diagram, the same steam prices for full capture and partial capture were applied. This however is most likely not realistic – full capture naturally requires more steam which likely has to be produced at a higher cost despite economy of scale (e.g. extra boiler, extra fossil fuel/power supply) than waste heat. This means that at a lower capture rate the share of low-value steam derived from waste heat will increase and thus reduce the cost (this can be seen as a step function with increased steam cost per tonne versus amount supplied steam). In addition to the results shown here, partial capture may thus perform even more economic than the results presented in this work.

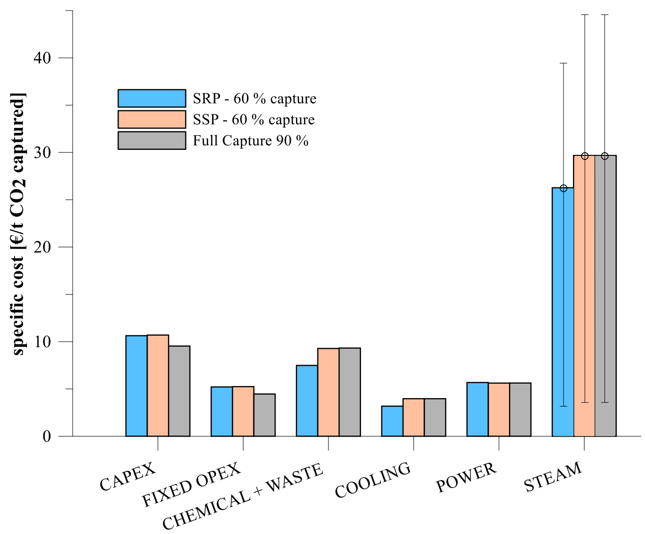


Figure 18. Specific capture cost for partial capture design paths SRP & SSP at 60% capture rate in comparison with full capture. Standard process configuration; the bar height for steam represents a steam price of 17 €/t; the error bars represent a steam price of 2 €/t and 25 €/t for the lower and upper bound, respectively.

CONCLUSIONS

An extensive modelling study on partial capture designs with 30 wt.% MEA as solvent has been carried out to determine the effect on reducing the capture rate on the design and cost of the process. Capture from a fairly large (200 kg/s) and CO_2-rich (20 vol.%) stream, representing a generic point source from process industry, is examined. Two partial capture design paths are investigated: 1) *split-stream path (SSP)*, where 90% of the CO_2 is removed from a slip stream; 2) *separation rate path (SRP)*, where the separation rate in the absorber is reduced, covering a range from 30 % to 97 % capture of the CO_2 in the flue gas. Both paths, especially SSP at low capture rates, reduce the absolute investment required and, thus, the risk for the investor. Depending on site specific conditions, the steam cost for partial capture with *either* design path may be lower than for full capture – enabling lower specific capture cost than for full capture despite working against economy of scale. The study concludes, that the SRP path requires less heat for solvent regeneration per tonne CO_2 and, thus, possesses the potential to outperform full capture in specific costs. Under the condition of this study, the SRP is favorable for CO_2 concentrations above 20 vol.%, a minimum of 0.2 Mt CO_2 captured per year, and given steam costs at 17 €/t and about 10 % cost reduction can be achieved compared to full capture (standard configuration). The study shows that, similar to full capture, intercooling (ICA) and rich split (RSS) process modifications are a cost effective process configuration for partial capture. The RSS configuration is favorable for all separation rates between 45 – 90 %, while intercooling is not viable for separation rates below 75 %. The techno-economic evaluation illustrates that operating costs, especially the steam cost, dominates the cost for CCS. This is also true for smaller scale; i.e. for partial capture the OPEX/CAPEX ratio remains above 1 for capture above 0.4 Mt CO_2 per year. With OPEX dominating the total capture cost, future work could study the gradual deployment of CCS at specific sites through SRP partial capture to full capture. By proper dimensioning of critical components, the capacity to expand the CO_2 capture towards full capture can be planned for.

In summary, this study has shown that partial capture may be a low-cost and cost-effective alternative to full capture CCS and can thereby incentivize investments into CCS by stakeholders – both industry and governments. Full capture in the example here is estimated to 63 €/t CO_2 capture. Applying partial capture to the same CO_2 source, costs between 35 – 70 €/t CO_2 captured (excluding transport and storage) are still required to initiate the investment and trigger the implementation of CCS in both industry and the fossil energy sector.

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Notes

The authors declare no competing financial interest.

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ABBREVIATIONS

CAPEX, capital expenditures; ICA, intercooled absorber; L/G, liquid-to-gas; MEA, monoethanolamine; OPEX, operational expenditures (including maintenance and labor); REF, reference CO_2-rich gas; RSS, rich solvent splitting; SRP, separation rate path; SSP, split stream path

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