Challenges and uncertainties of ex ante techno-economic analysis of low TRL CO_2 capture technology

Lessons from a case study of an NGCC with exhaust gas recycle and electric swing adsorption

van der Spek, Mijndert; Ramirez, Andrea; Faaij, André

DOI
10.1016/j.apenergy.2017.09.058

Publication date
2017

Document Version
Final published version

Published in
Applied Energy

Citation (APA)
van der Spek, M., Ramirez, A., & Faaij, A. (2017). Challenges and uncertainties of ex ante techno-economic analysis of low TRL CO_2 capture technology: Lessons from a case study of an NGCC with exhaust gas recycle and electric swing adsorption. *Applied Energy*, *208*, 920-934. https://doi.org/10.1016/j.apenergy.2017.09.058

Applied Energy 208 (2017) 920–934

Contents lists available at ScienceDirect

Applied Energy

journal homepage: www.elsevier.com/locate/apenergy

Challenges and uncertainties of ex ante techno-economic analysis of low TRL CO_2 capture technology: Lessons from a case study of an NGCC with exhaust gas recycle and electric swing adsorption

Mijndert van der Spek[a,*], Andrea Ramirez[b], André Faaij[c]

[a] Copernicus Institute of Sustainable Development, Section Energy & Resources, Utrecht University, Heidelberglaan 2, 3584 CS Utrecht, The Netherlands
[b] Department of Energy Systems and Services, Section Energy & Industry, Delft University of Technology, Jaffalaan 5, 2628 BX Delft, The Netherlands
[c] Center for Energy and Environmental Sciences, IVEM, University of Groningen, Nijenborgh 6, 9747 AG Groningen, The Netherlands

HIGHLIGHTS

- The challenges of techno-economic analysis of very low TRL technologies are explored.
- Hybrid approaches to project future performance of low TRL technologies are proposed.
- Electric swing adsorption is infeasible to economically capture CO_2 from an NGCC.

ARTICLE INFO

Keywords:
Techno-economic analysis
Electric swing adsorption
Hybrid approach
Solid sorbents
System analysis
Technological learning

ABSTRACT

this work addresses the methodological challenges of undertaking techno-economic assessments of very early stage (technology readiness level 3–4) CO_2 capture technologies. It draws lessons from a case study on CO_2 capture from a natural gas combined cycle with exhaust gas recycle and electric swing adsorption technology. The paper shows that also for very early stage technologies it is possible to conduct techno-economic studies that give a sound first indication of feasibility, providing certain conditions are met. These conditions include the availability of initial estimates for the energy use of the capture technology, either from bench scale measurements, or from rigorous process models, and the possibility to draw up a generic (high level) equipment list. The paper shows that for meaningful comparison with incumbent technologies, the performance of very early stage technologies needs to be projected to a future, commercial state. To this end, the state of the art methods have to be adapted to control for the development and improvements that these technologies will undergo during the R & D cycle. We call this a hybrid approach. The paper also shows that CO_2 capture technologies always need to be assessed in sympathy with the CO_2 source (e.g. power plant) and compression plant, because otherwise unreliable conclusions could be drawn on their feasibility. For the case study, it is concluded that electric swing adsorption is unlikely to become economically competitive with current technologies, even in a highly optimised future state, where 50% of the regeneration duty is provided by LP steam and 50% by electricity: the net efficiency of an NGCC with EGR and optimised ESA (49.3%$_{LHV}$; min–max 45.8–50.4%$_{LHV}$) is lower than that of an NGCC with EGR and standard MEA (50.4%$_{LHV}$). Also, investment and operational costs are higher than MEA, which together with ESA's lower efficiency leads to an unfavourable levelised cost of electricity: 103 €/MWh (min–max 93.89–117.31 €/MWh) for NGCC with ESA, versus 91 €/MWh for NGCC with MEA.

1. Introduction

Carbon capture and storage (CCS) is a necessary technology towards deep decarbonisation of the energy and industrial sectors, thereby mitigating severe global warming [1]. To progress technical implementation of CCS, the past decade and a half have seen the discovery and development of a wide portfolio of carbon capture technologies [2–5], many of which are still in early stage of development, i.e. technology readiness level (TRL [6]) = < 4.

For CCS to succeed as a CO_2 mitigation strategy, it is necessary to especially advance the technologies that are most promising in terms of technical, economic and environmental performance. To reach a

---

* Corresponding author.
E-mail address: m.w.vanderspek@uu.nl (M. van der Spek).

http://dx.doi.org/10.1016/j.apenergy.2017.09.058
Received 12 June 2017; Received in revised form 31 August 2017; Accepted 10 September 2017

commercially ready portfolio of the most promising CCS technologies on time, and in an efficient way, targeted technology selection and investment are required [7]. This selection requires performance assessment of the different technology alternatives, by analysing the performance of the carbon capture technology in an integrated system (CO_2 source, carbon capture installation, transport, and storage) already during the early stages of development. These analyses will also point out key improvement options, thereby supporting RD & D.

The current best practice for techno-economic analysis of (near) commercial energy technologies involves well established methods. Typically, the technical performance is evaluated using process models [8–10], and the economic performance is estimated based on detailed equipment lists and mass and energy balances that are derived from the process model [11–14]. As we will show in this paper however, there are challenges to use these best practices for technologies down the TRL ladder, and shortcuts or simplified approaches may be needed to produce the desired performance metrics. Inherently, uncertainties in the analyses will be substantial and need specific attention, especially because they can point out hotspots for technology improvement.

In this context, the aim of this paper is threefold. First, to identify the key challenges when using state of the art methods for developing techno-economic assessments of very early stage CCS technologies (TRL 3–4). Second, to identify and develop shortcuts and/or other methods that can be used to deal with these challenges, and third, to extract lessons for performing meaningful techno-economic assessments for very early stage technologies. In this way the paper advances the state-of-the-art of very early stage ex-ante technology assessment. Its novelty is especially in presenting hybrid approaches to techno-economic performance assessment, approaches that have not been published before.

To this end, this work will consider the performance assessment of an emerging carbon capture technology that is at low TRL ($3-4$). In this paper, we will use electric swing adsorption (ESA) as case study. ESA is proposed as a postcombustion CO_2 capture technology analogous to temperature swing adsorption [15–18]. The differentiating aspect of ESA is that the heat required for the temperature swing is provided by electricity using the Joule effect, instead of by low pressure steam, which is used in TSA. The ESA process in this work is analysed in the context of CO_2 capture from a natural gas fired power plant, as suggested by the developers of the technology [16].

Although this work focuses on power plants with CO_2 capture technologies, its lessons are equally applicable to other emerging energy technologies, especially the ones down the TRL ladder. Examples include solar fuels, biofuels and bioproducts, and concentrating solar power.

2. General approach

This work systematically investigated the steps that were undertaken in the very early stage techno-economic assessment of NGCC with ESA. First, a generic literature scan was performed to select an emerging technology that could function as case study (ESA, see introduction). Second, an in depth literature study was conducted to investigate the technology characteristics and establish the case study scope (Sections 3.1–3.2). The literature study specifically investigated the availability of data for modelling purposes, and the development potential of the technology. The case study scoping also included selection of a reference technology for comparison of the techno-economic performance: also for low TRL technologies a benchmark is relevant to put their techno-economic performance into perspective. Third, available methods for techno-economic assessment of CO_2 capture technologies were reviewed, and suitable methods were introduced for this case study, notably hybrid approaches (Section 4). Fourth, the performance of the low TRL technology and the benchmark technology were modelled and the results, including sensitivities, were analysed (Section 5). Last, the identified challenges and lessons for this kind of (low TRL) technology analysis were summarized in the conclusions (Section 6).

3. Technology description and scope

3.1. Technology description

The aim of the literature review was to identify the technology characteristics and to set the scope for the case study. This amongst others meant investigating the availability and strength of data for modelling purposes, which was required for model selection (Section 4). In this respect, and given the low TRL, there were two options: the available knowledge base was sufficient to undertake a classic techno-economic analysis that included integration of the technology with the power plant (e.g. like [11,13,19,20]). This would require as a minimum A) the possibility to make an estimate of mass and energy balances and B) a rough understanding of equipment and size. Or, a classic techno-economic assessment was not possible due to data limitations, and a preliminary assessment method was required, for instance as described in [21,22]. Questions that needed to be answered included “are estimates available of CO_2 separation efficiency and purity or can these be produced?”, “Are estimates of separation energy requirement available or can these be produced?”, and “is a process design available or can this be produced?”

3.1.1. Electric swing adsorption

The investigated CO_2 capture technology (ESA) is a specific type of temperature swing adsorption (TSA) technology. It consists of trains of parallel columns with solid sorbents that are alternatively cooled and heated to respectively adsorb and desorb the CO_2 (Fig. 1). In a standard TSA process, the heat is provided by steam that is extracted from the power plant steam cycle [23]. However, this poses engineering issues for large scale application due to slow heat transfer of steam to adsorbent, resulting in long cycle times and large sorbent requirements [15,16,24,25]. ESA is meant to address this challenge, using electricity to heat the sorbents through the Joule effect [15–18].

The cycle of a basic ESA or TSA process consists of four steps: (1) feed, (2) heating, (3) desorption, and (4) purge [15,26]. But more advanced cycles have been proposed that may benefit from step and heat integration [18,25]. With respect to separation efficiency and energy requirement, Grande et al. [16] estimated a regeneration duty of $2.04\ GJ_e/t\ CO_2$ for an ESA process using zeolite 13X sorbent and a five step cycle. They reported approximately 80% CO_2 recovery and 80% CO_2 purity, a significant achievement, but not yet fully representative of the performance of a commercial CO_2 capture process, where 90% capture at $>$ 95% purity is required. For this 5-step cycle, also column sizes were reported which may form the basis of capital cost estimations.

The above concise technology description shows that the ESA technology is developed enough to simulate mass and energy balances, and that preliminary PFD’s and equipment sizes are available. This means that classic techno-economic analysis could be undertaken. However, because the technology is at TRL 3 (or proof of concept stage), performance results will change as it moves towards commercialisation. To provide a fair representation of potential, future, performance, it is necessary to assess the process taking into account potential performance advances over its development cycle. This requires knowledge on which advances are likely to take place and how these will influence the process.

The development potential of T/ESA will likely come from more efficient sorbents that rely on chemical bonding with CO_2 [2,25,27]. Chemical sorbents may have higher working capacities than typically used physical sorbents, such as zeolite 13X and zeolite 5 A [27–29]. As an additional benefit, chemical sorbents often perform better in humid environments such as flue gas streams, whereas physical sorbents quickly deactivate in the presence of water [23,25,30], and thus require an additional drying unit upstream the CO_2 capture step. An example of a chemical adsorbent that has been well characterised, and that has favourable properties in the presence of water is amine grafted pore

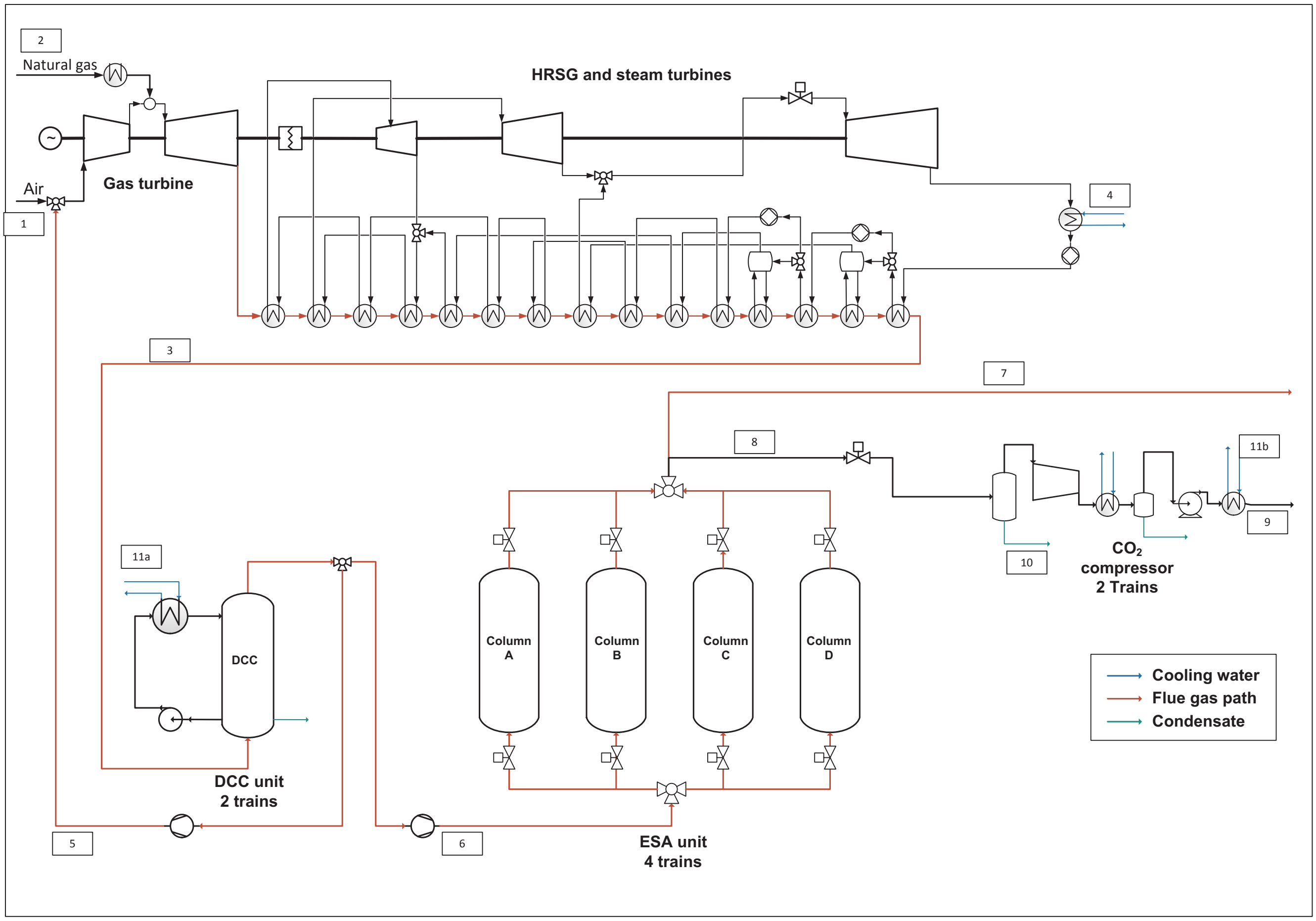

Fig. 1. Process flow diagram of an integrated NGCC with electric swing adsorption system. The numbers represent flows in the mass balance.

expanded silica (TRI-PE-MCM-41) [28,31,32], but many others have been developed (e.g. [27]). TRI-PE-MCM-41 has shown a high CO_2 capacity of up to 2.94 mol/kg at a CO_2 partial pressure of 5 kPa, and in the presence of water [28], which represent typical conditions of NGCC flue gas streams. Other process improvements may come from optimising the steps of the adsorption–desorption cycle.

3.1.2. *Natural gas combined cycle*

Natural gas combined cycles (NGCC) are commercial technologies. The selected configuration of the NGCC was based on the EBTF framework [14] and includes two generic F-class gas turbines (GT's), two heat recovery steam generation (HRSG) units, and one train of HP, IP, and LP steam turbines (Fig. 1). The power plant size is approximately 830 MW. To follow the current advancements in NGCC with CCS development, the NGCC was equipped with exhaust gas recycling (EGR), which reduces the flue gas volume and increases the flue gas CO_2 content, thereby decreasing capital costs of the CO_2 capture plant and simplifying CO_2 separation [33]. When standard gas turbines are used, their efficiency will decrease slightly because the EGR changes the composition of the air inlet. In this study, we however assumed the same efficiency values as suggested by EBTF, similar to [34], assuming GT designs can be modified to render the same efficiency with EGR.

3.1.3. *CO_2 compression*

CO_2 compression was considered a commercial technology because it is used in numerous US enhanced oil recovery projects. This study uses a 5-stage CO_2 compressor and a liquid CO_2 pump (Fig. 1), increasing the pressure to 110 bar. During compression the CO_2 stream is dried to an H_2O content of 200 ppm_v using knock out vessels and a drying unit [8].

3.1.4. *Reference technology*

The general purpose of a feasibility study is to investigate whether or not a novel technology performs better, or has advantageous characteristics, when compared to an incumbent technology. In the case of CO_2 capture, postcombustion capture with standard MEA technology is often seen as the incumbent CO_2 capture technology, and is typically used as reference [8,10,11,13,34]. Recently constructed CO_2 capture plants like Boundary Dam and Petra Nova use more advanced postcombustion solvent technologies than MEA, (Cansolv and MHI KS1,), that show improved energetic performance over MEA. Therefore, for a very novel technology like ESA to be techno-economically "feasible", it should at least outperform MEA technology in terms of parasitic impact on the NGCC and cost impact. Because Cansolv and KS1 are proprietary blends, and their exact formulation is unknown, this paper uses MEA as a reference technology, acknowledging that any emerging technology should show drastically improved performance with respect to MEA to be able to compete with commercially used solvent systems.

3.2. *System scope*

Because the ESA technology was developed far enough to undertake classic techno-economic assessment, the scoping was similar to that of advanced technologies. It included setting the location (NW-Europe), currency (€), and temporal (2014), boundary conditions such as ambient characteristics and cooling water specifications (Appendix A), and the units/plants that are included in the system. With respect to the latter, note that to assess the technical performance of any CO_2 capture technology, it is essential to analyse it in the context of the other units it is connected to, i.e. the power plant (or industrial CO_2 source) and the compression plant (further discussed in Section 5.1). For economic performance analysis, also CO_2 transport and storage should preferably be included [12].

4. Review and selection of techno-economic modelling methods

4.1. Technical performance assessment

The main purpose of the technical performance analysis of any CO_2 capture technology is to find the energy required for separation of the CO_2 from the power plant flue gas stream, given a specified CO_2 yield and purity. For solid adsorbents, this comes down to determining the energy requirement for adsorbent regeneration. The metric for this is the specific regeneration duty (SRD: energy used per tonne of CO_2 captured), analogous to the metric used for postcombustion solvent technology.

Selecting a suitable modelling method to estimate the technical performance is a problem that many process modellers face. Typically, a choice can be made between rigorous, shortcut, or simplified process models [9,37]. Rigorous methods have the potential to provide more detailed and/or accurate results. However, time, knowledge base, and sometimes skills may be limited, therefore limiting model selection to more simplified methods. To aid the selection process, an attribute complexity matrix and/or pedigree analysis of the knowledge base can be used [9,37]. These methods screen the (scientific) knowledge base of a technology on criteria such as empirical basis (which knowledge is available, from which kind of, and from how many sources was it derived); theoretical understanding (how strong is the theory on this technology, and the level of consensus between scientists); methodological rigour (how was the knowledge generated), and validation process (was the data validated and how sound was the validation exercise). The gained understanding of the technology's knowledge base can then be used to select modelling methods that fit the available knowledge and the purpose of the modelling study. For example, if the knowledge base strength is low to medium, one may need to select simplified modelling methods.

4.1.1. Electric swing adsorption unit

For solid sorbents, the existing best practice is to perform rigorous analysis of the adsorption–desorption cycle, using dynamic modelling methods. This was shown amongst others by Grande et al. [16,24,38] for ESA and is equivalent to cyclic modelling of PSA and TSA systems (e.g. [39,40]). This *rigorous method* considers the time aspect of adsorption cycles as well as mass transfer limitations into the sorbent pores. Models like these consist of a set of partial differential equations that are solved numerically. The construction of such a rigorous cyclic adsorption model is, however, a laborious process – model development times of over a year are not uncommon – and requires a great deal of skill from the modeller.

The second option is to use a *short-cut method* analogous to the one developed by Joss et al. for TSA [26]. The differential equations in this type of model can be solved analytically by treating the columns as isothermal rather than adiabatic, thereby neglecting temperature fronts. Also, it excludes mass transfer limitations, but rather calculates the CO_2 adsorption based on adsorption equilibrium. Although this kind of model is easier to construct, and *simulation* time can be greatly reduced compared to rigorous cyclic models, model *development* is still time and resource intensive, which can be an issue for rapid technology portfolio screening.

The third option is to use a *simplified method*. This method estimates the SRD of an adsorption process using simple (non-differential) equations to calculate the adsorption heat, sensible heat, gas (adsorbate) heating, and water vaporization, and sums these values to a total SRD estimate [41]. This method can be suited to estimate *initial performance* of *continuous adsorption processes* (such as found in sorbent systems using two fluidized beds), but tends to overestimate the SRD in a *cyclic process,* because it fails to include interactions between the steps in a cycle (e.g. sensible heat generated during the adsorption step that already partially heats the column for the desorption step).

In this specific case, the NGCC with ESA assessment was part of a larger European research project where ten CCUS technologies were screened in four years, something not uncommon in energy system analysis. This, however, limited the available modelling/assessment time. Therefore, in this paper we used a hybrid option to project the performance of the ESA technology: we based the SRD estimate of the ESA process on existing, preliminary, *rigorous* modelling results (the aforementioned zeolite 13X modelling results by Grande et al. [16]). Then, we constructed a performance estimate of a *future* advanced/commercial ESA process, extrapolating Grande's results to a state where 90% CO_2 at 96% purity is captured using an advanced amine grafted sorbent (TRI-PE-MCM-41, Section 3.1.1). This means that preliminary results of a *rigorous modelling method* were used, and were adapted using the analogy of the *simplified method*.

The chosen technical assessment strategy required adjusting Grande's results in two ways. First, an adjustment was needed to increase the CO_2 recovery and purity to respectively 90% and 96%. To this end we assumed that these levels could be reached by two system improvements: 1) increasing the flue gas CO_2 concentration to $6.4\%_{vol}$ by means of the EGR (in the original study by Grande et al. this was $3.5\%_{vol}$ leading to a higher required enrichment factor), and 2) using the aforementioned TRI-PE-MCM-41 sorbent, which is much more selective towards CO_2 than the original zeolite sorbent. This sorbent has the additional advantage that it is able to operate effectively in humid flue gas, contrary to zeolite 13X [28].

The second adjustment involved the value of the SRD. We projected the SRD of the future ESA process on the value found by Grande et al. [16], accounting for differences in sorbent properties (Table 1) and process design (Table 2), and including water vaporization. This was done in three steps (Fig. 2), using the analogy of the SRD of a continuous (non-cyclic) adsorbent process [41]:

$$Q_{des} = \Delta H_a + m\cdot C_p\cdot(T_{des}-T_{ads}) + Q_{H_2O} \tag{1}$$

Where Q_{des} is the total heat requirement (kJ/mol), ΔH_a is the enthalpy of adsorption (kJ/mol), m is the sorbent mass (kg), C_p is the sorbent heat capacity (kJ/kgK), T_{ads} and T_{des} are the respective adsorption and desorption temperature, and Q_{H2O} is the heat of water vaporization.

Step 1 encompassed approximating the division between regeneration energy used for CO_2 desorption and for sensible heat. This division was required for step 3. The results from Grande et al., that

Table 1
Sorbent properties of Zeolite 13X and TRI-PE-MCM41.

Adsorbent properties	Unit	Zeolite 13X	TRI-PE-MCM-41
Adsorbent type		30% AC binder + 70% 13X[1]	30% AC binder + 70% TRI-PE-MCM-41[2]
Max adsorbent loading	mol/kg	3[1]	2.94[3]
Working capacity	mol/kg	0.8[1]	25% of max loading[4]
Heat of Adsorption	kJ/mol	45[1]	70[5]
Heat capacity	J/kg/K	900[1]	1000[6]

[1] [16].

[2] same amount of binder assumed as in the zeolite monolith.

[3] [28]. The highest measured loading capacity was used to represent an advanced ESA process.

[4] [25]: working capacities of 0.18 to 0.26 (% max capacity) were reported for an amine-impregnated polymeric resin. A high value was selected to represent an advanced ESA process. This working capacity was validated with TRI-PE-MCM-41 adsorption–desorption cycles presented in Serna-Guerrero et al. [42].

[5] An average value of 70 kJ/mol was used as reported by [32]. This value falls with within the range of 43–92 kJ/mol that was measured/calculated for similar amine impregnated sorbents [2,41,43].

[6] [43]. This value is relatively conservative compared to heat capacity of activated carbon and SiO_2 (both ~ 700 kJ/kg K), which constitute about $65\%_{wt}$ of the solid sorbent.

Table 2
Advanced ESA process design parameters.

Parameter	Unit	Value
Flue gas temperature	°C	37.5[1]
Regeneration temperature	°C	110[2]
Condenser temperature	°C	40[3]
Flue gas inlet pressure	bar	1,1[4]
CO_2 outlet pressure	bar	1[4]
ESA cycle time	min	40[4]
Column height	m	11.6[4]
Train size	Columns/train	4[4]
Allowed sorbent deactivation	% of max activity	70[5]

[1] NGCC model output.
[2] Max amine modified sorbent regeneration temperature is 120 °C. Higher temperatures cause increased sorbent degradation [31].
[3] Meant to cool the CO_2 stream before entering the compression unit.
[4] Same cycle/process design as in Grande et al. [16].
[5] Educated guess.

were used as a basis specified a total SRD of 2.04 GJ_e/t CO_2 captured, of which 0.13 GJ_e/t for heating of the adsorbate (gas), and the remaining 1.91 GJ_e/t for heating of the adsorbent. This means that the 1.91 GJ_e/t includes adsorption enthalpy and sensible heat. A division between the two – expressed as their contributions (%) to total heat - was estimated using the analogy of the continuous process (Eq. (1)), assuming the division of sorbent heating in the cyclic process was similar to the continuous process (Fig. 2b).

Step 2. Second, the heat of water vaporization was added. Zhang et al. [41], estimated this was 0.38 GJ/t for a continuous adsorption process. Again, this value was expressed as the contribution of the total heat of adsorption and analogously added to the cyclic SRD (Fig. 2c).

Step 3. Third, the SRD value of the future ESA process with the TRI-PE-MCM-41 sorbent was projected based on the SRD terms of the current, zeolite sorbent process (Fig. 2d), incorporating the differences in sorbent properties and process design following Eqs. (2)-(5):

$$Q_{des,TRI} = Q_{des,ZEO} \cdot F_1 \tag{2}$$

$$Q_{sens,TRI} = Q_{sens,ZEO} \cdot F_2 \cdot F_3 \cdot F_4 \tag{3}$$

$$Q_{gas,TRI} = Q_{gas,ZEO} \cdot F_5 \tag{4}$$

$$Q_{H2O,TRI} = Q_{H2O,ZEO} \tag{5}$$

Where $Q_{des,TRI}$ and $Q_{des,ZEO}$ are the enthalpies of ad/desorption of the advanced sorbent and the zeolite sorbent, $Q_{sens,TRI}$ and $Q_{sens,ZEO}$ are the sensible heat requirements of the advanced sorbent and the zeolite sorbent, $Q_{gas,TRI}$ and $Q_{gas,ZEO}$ are the adsorbate heating requirements of the advanced sorbent and the zeolite sorbent, $Q_{H2O,TRI}$ and $Q_{H2O,ZEO}$ are the water vaporization heat requirements of the advanced sorbent and the zeolite sorbent, and F_1, F_2, F_3, F_4, and F_5, are factors representing the differences in sorbent properties and process design.

Finally, the total SRD (GJ electricity per tonne of CO_2 captured) was calculated by summing the individual heat components of the ESA process:

$$SRD = Q_{des,TRI} + Q_{sens,TRI} + Q_{gas,TRI} + Q_{H2O,TRI} \tag{6}$$

4.1.2. NGCC with exhaust gas recycling

The NGCC model was specified in Aspen Plus V8.4., following the EBTF design specifications [14]. Because NGCC technology is commercial, and many such models exist, this was a straightforward activity, contrary to ESA modelling. An exhaust gas recycle (EGR) ratio of 35% was assumed. Flame stability in the GT combustion chamber may allow higher recycle ratios, but tests suggest that above 35% recycle excessive NO_x and CO start to form which is undesired from an environmental perspective [44]. The EGR includes a direct contact cooler (DCC) to cool the flue gas to 25 °C before re-entering the combustion chamber.

4.1.3. CO_2 compression unit

Like the NGCC, the compression unit was modelled with Aspen plus V8.4., using the same design as in [8,9].

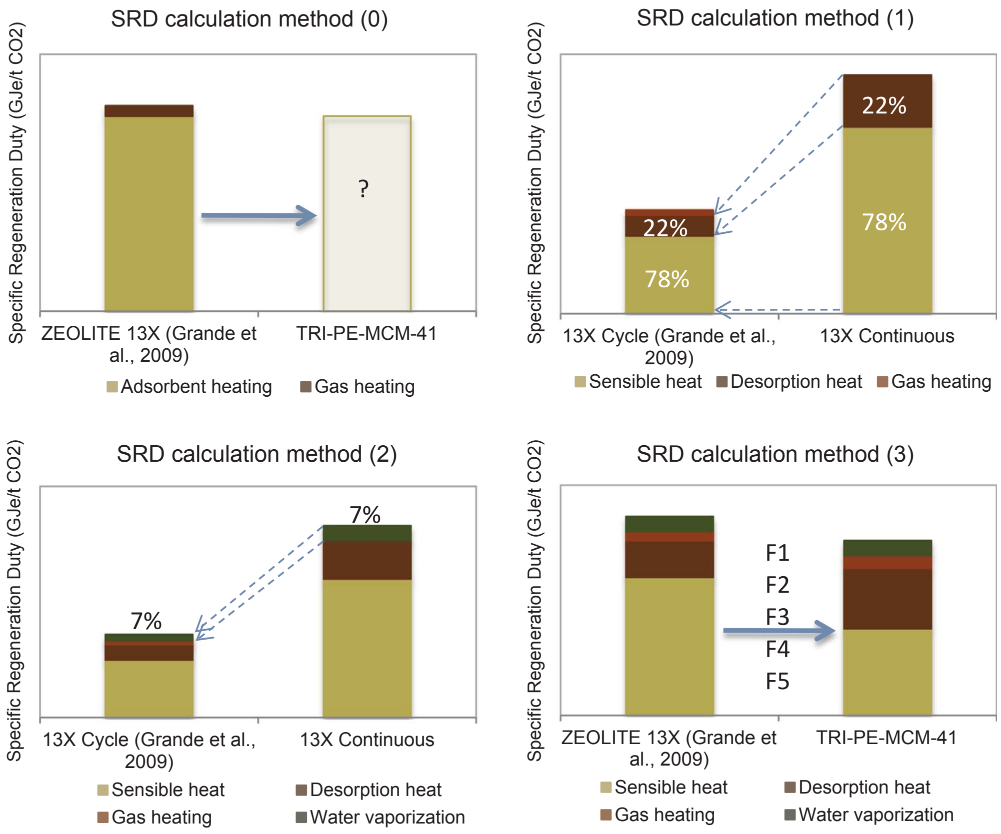

Fig. 2. Graphic representation of SRD calculation method using the hybrid approach, based on the modelling work by Grande et al. [16].

4.1.4. Benchmark MEA absorption unit

The benchmark MEA absorption unit was modelled using a rate-based model that was specified in Aspen Plus V8.4. The model used kinetic reactions for the bicarbonate (Eq. (7)) and MEA carbamate (Eq. (8)) formation following Kvamsdal and Rochelle [45], applying their reported kinetic parameters for the Arrhenius equation. The reaction equilibria of other reactions were calculated from the Gibbs free energy. The flowsheet included a standard absorber-stripper configuration with water washes on top of absorber and stripper [9]. Advanced process configurations like lean vapour compression and split flows were excluded from this study. The MEA process was earlier reported in [9].

$$OH^- + CO_2 \rightarrow HCO_3^- \tag{7}$$

$$MEA + CO_2 + H_2O \rightarrow MEACOO^- + H_3O^+ \tag{8}$$

4.2. Economic performance assessment

Economic performance assessment of CCS technologies typically aims to find estimates of capital and operational costs, and uses these to calculate the Levelised Cost of Electricity (LCOE) and the Cost of CO_2 Avoided (CCA) ([12,13], see Appendix B for equations).

The main challenge for economic analysis of early stage technologies is to find the investment costs of a potential Nth of a kind (NOAK) plant, and to estimate the technology specific operational costs. Estimating the generic operational costs and selecting financial parameters is more straightforward because often standard values are used. Therefore, the description below focusses mainly on methodological aspects of capital cost estimation and technology specific operational costs.

4.2.1. Capital cost estimation

4.2.1.1. Exponent methods versus bottom up methods. Different capital costing methods exist for technology screening and feasibility studies, of which the most well-known are the exponent method and the bottom up (or factoring) method [12,46–48].

Exponent methods can be used when a (reliable) cost estimate of the same (or similar) technology already exists. The exponent method uses a cost estimate of a reference study and scales this to the size of the plant in a new study:

$$C = C_{ref}\left(\frac{Q}{Q_{ref}}\right)^n \tag{9}$$

Where C and C_{ref} are the capital costs of the equipment/plant in respectively the new study and the reference study; Q is the equipment capacity in the new study, Q_{ref} is the equipment capacity in the reference study, and n is the scaling exponent.

A typical characteristic of very low TRL technologies is that reference cost estimates are unavailable. In that case, (only) a bottom up method can be used, which calculates the capital costs based on an equipment list of the studied process.

4.2.1.2. Direct methods versus indirect methods for Nth of a kind cost estimations. Bottom up capital costing methods rely on a detailed equipment list and purchased equipment cost estimates, which are then multiplied with factors for installation, engineering, procurement and contracting, and contingencies [12,49]. This *direct method* of estimating the cost of an NOAK plant is suitable for technologies that are close(r) to commercialisation, and of which the design is well-defined. Early stage technologies (TRL 3, 4, sometimes 5) often lack this level of design detail, and are usually described by simplified process flow diagrams and basic equipment lists. The estimated costs of these technologies typically escalate in the period between the lab stage and the first large demonstration plants (first of a kind (FOAK); TRL 7, 8) (Fig. 3) [50,51]: during upscaling from lab to demonstration the design is further detailed and unforeseen technological issues are uncovered which need additional design solutions, typically increasing the costs of the technology. After this point, the costs start to fall as more commercial plants are built (TRL 9) and technological learning commences (Fig. 3). The Nth of a kind plant is reached when cost decline starts to level out, i.e. when a significant part of the learning has taken place.

The *direct* bottom up method directly calculates the costs of the NOAK plant but disregards the cost curve that very early stage technologies follow and the effect this may have on NOAK costs. Instead, an *indirect method* or *hybrid method* may be more suitable. Rubin [50,51] proposed such an indirect method including the following three steps [51]: (1) estimate the preliminary engineering, procurement and contracting (EPC) costs bottom up, based on a (simplified) equipment list of the novel technology, (2) find the FOAK total plant cost (TPC) costs by adding appropriate process and project contingencies, and (3) find the NOAK total plant cost using learning curves. This novel hybrid method provides an elegant solution to low TRL technology cost estimation, because it includes the large cost escalations to the first built plants, but also because it allows more detailed evaluation of uncertainties during the different stages of the development curve.

4.2.1.3. Capture unit capital costs. The capital costs of the ESA (and to a lesser extent of the EGR) equipment could not be calculated using an exponent method because a reference cost for this – or a similar - technology was lacking in available literature. Therefore, we chose to estimate the NOAK capital costs bottom up with the indirect method, using a preliminary equipment list of the ESA process. As a reference, the NOAK TPC costs of the CO_2 capture equipment were also estimated using the *direct* bottom up method, to highlight the differences and similarities in outputs between the two approaches.

The three steps of the indirect method were taken as follows:

Step 1. Bottom up estimation of the EPC costs based on a simplified process flow diagram and equipment list. At the heart of the ESA process are the adsorption columns, flue gas and CO_2 fans, and some gas storage tanks (Fig. 1). The ESA column sizes were estimated based on the process design presented by Grande et al. [16]. The height of the adsorption columns was kept equal to Grande's study, while the column diameters were varied to match the flue gas flow in this work, taking into account the working capacity and density of TRI-PE-MCM-41 (see Table 3). This ensured the gas speed in the columns to remain the same as in Grande's work. The size of the other ESA equipment (fans, storage vessels, and valves) was defined based on the ESA mass balance and engineering rules of thumb.

Based on the simplified equipment list, the purchased equipment costs were calculated using the Aspen capital cost estimator (V8.4). The cost of each equipment was then multiplied with an individual installation and EPC factor that were retrieved from an in-house database [13]. This led to an engineering, procurement, and contracting (EPC) cost estimate. The same approach was applied to the EGR equipment.

Step 2. The second step included estimation of the total plant cost of the FOAK plant, based on the EPC cost estimated in step 1 (Fig. 3). To this end, the EPC costs were escalated with process and project contingencies. Guidelines exist for the amount of process and project contingencies to apply, based on technology readiness and type of cost estimate, respectively ([12], following EPRI and AACE) as displayed in Table 4Table 5. Note that both tables give ranges for the process and project contingencies, rather than single values, representing the uncertainty in cost estimate escalations. This work used these ranges as uncertainty margins for the FOAK cost estimates. For the ESA (+EGR) technology, process contingencies from 30 to 70% were applied with a nominal value of 50%, representing its status as a "concept with bench scale data" (TRL 3/4). For the MEA (+EGR) technology, process contingencies from 5 to 20 were applied with a nominal value of 12,5%, representing its "full-sized modules have been operated" status. We considered both cost estimates AACE class 4 estimates [46] and hence 30–50% project contingencies were applied with a nominal value of

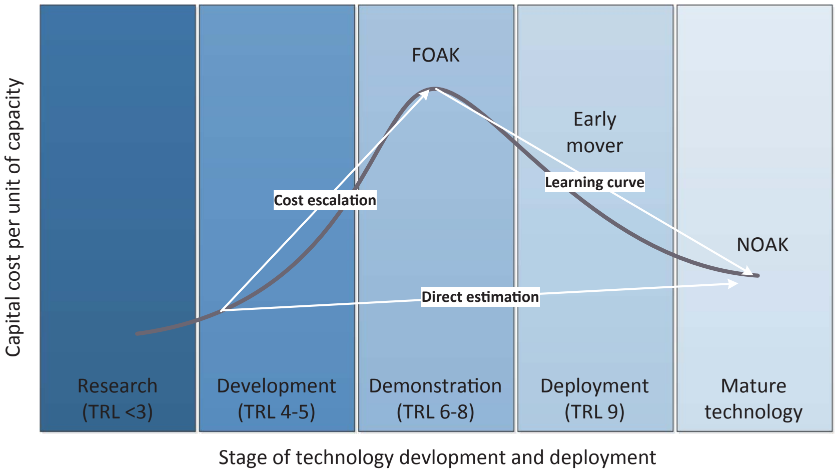

Fig. 3. Typical capital cost trend of a new technology. White arrows show the direct and indirect/hybrid approaches to estimate the NOAK capital costs of a new technology. .
Adapted from [51]

Table 3
Column sizing input parameters.

Parameter	Unit	Value
Feed step duration	min	10[a]
CO_2 flow to ESA	kg/s	79.88[b]
Parallel columns per train	–	4[a]
Column void fraction	–	0.4[a]
Column length	m	11.6[a]
Column diameter	m	To be calculated
Sorbent density	kg/m^3	910[c]

[a] [16].
[b] Aspen Plus NGCC process model.
[c] The value of 910 kg/m^3 was assumed based on measured densities of TRI-PE-MCM-41 [52] and activated carbon [15]. Activated carbon (989 kg/m^3) and TRI-PE-MCM-41 (880 kg/m^3) were assumed present in the monolith in a 30–70 ratio.

Table 4
Guidelines for process contingency costs ([12] based on EPRI 1993).

Technology status	Process contingency (% of associated process capital)
New concept with limited data	40 +
Concept with bench scale data	30–70
Small pilot plant data	20–35
Full-sized modules have been operated	5–20
Process is used commercially	0–10

Table 5
Guidelines for project contingency costs ([12] based on EPRI 1993).

Cost classification	Design effort	Project contingency (%)[1]
Class I (similar to AACE class 5/4	Simplified	30–50
Class II (similar to AACE class 3	Preliminary	15–30
Class III (similar to AACE class 3/2)	Detailed	10–20
Class IV (similar to AACE class 1)	Finalised	5–10

[1] Percentage of the total of process capital, engineering and home office fees, and process contingency.

40% (see also Section 4.3 sensitivity analyses).

Step 3. The NOAK costs were calculated using a single factor learning curve [53]:

$$C_{NOAK} = C_{FOAK}\cdot\left(\frac{N_{NOAK}}{N_{FOAK}}\right)^{b} \tag{10}$$

Where C represents the capital costs (total plant cost, TPC), N is the number of installed plants, and b is the learning rate coefficient, calculated from:

$$LR = 1{-}2^{b} \tag{11}$$

Where LR is the learning rate. This single factor learning curve[1] combines learning by doing, learning by searching (continued RD & D in the commercial stage of technology deployment), and scale factors [7].

The three determining parameters in Eqs. (10) and (11), other than the costs of the FOAK plant, are the values of N_{FOAK}, N_{NOAK}, and LR. It is common practice in literature on learning curves to use value for the installed capacity rather than installed number of plants, N [53]. However, Greig et al. [54] provided a rather appealing and useful definition for FOAK, early mover, and NOAK CCS plants, based on the number of installed plants, rather than installed capacity. Their definition of N_{FOAK} and N_{NOAK} includes the number of installed plants worldwide and defines FOAK as < 10 demonstration and/or commercial scale plants wold-wide, early movers as > 10 – < 20 commercial scale plants world-wide, and NOAK as > 20 commercial scale plants world-wide. We adopted Greig's definition in this study: for N_{FOAK} the range of 1–10 built plants was used with a nominal value of 5 (see also Section 4.3 sensitivity analyses). For N_{NOAK} the value of 20 built plants was used.

Because learning rates are based on the costs of built and operating plants, they are still unknown for CCS technologies. Instead, learning rates from analogous technologies can be used as a proxy. Rubin [51] and Van den Broek et al. [53] proposed to use the learning rate from flue gas desulphurisation (LR = 0.11) as a proxy for wet CO_2 capture systems, because of process similarities.

Table 6 presents a selection of learning rates found for the process industry, including a range of gas processing technologies. The table shows that most gas processing technologies have learning rates between 0.10 and 0.14, but that also outliers and inconsistencies exist: for example, Rubin et al. [55] report a learning rate of 0.27 for steam methane reformers, while Schoots et al. [56] report a value of 0.11 ± 0.06 for the same technology.

For MEA technology, we adopted a learning rate of 0.11 as proposed by Rubin [55] and Van den Broek [53]. For ESA technology, it was more challenging to decide on an appropriate learning rate. As said, an ESA process most resembles a pressure swing adsorption process as used in e.g. hydrogen production. Learning rates for hydrogen production have been reported (Table 6) but cover the whole plant: steam

---

[1] Also so-called multi factor learning curves exist where the single factor is disaggregated to separately treat learning by doing, learning by searching, and cost reductions by economies of scale [7]. However, data scarcity often inhibits this division in separate cost reduction drivers, and even when data is available, there may be significant overlap between learning by doing, searching, and scale, leading to question the validity of such a division [7].

Table 6
Selection of learning rates reported for analogous technologies in the process industry.

Technology	Capital cost learning rate	Reference
Flue gas desulphurisation (FGD)	0.11	[55]
Selective catalytic reduction (SCR)	0.12	[55]
LNG production	0.14	[55]
Oxygen production (ASU)	0.10	[55]
Hydrogen production (SMR)	0.27	[55]
Hydrogen production (SMR)	0.11	[56]
Ammonia production	0.29	[57]
Urea production	0.11	[57]
Bulk polymers (PE/PP/PS/PVC)	0.18–0.37	[57]

methane reforming, water gas shift, and PSA, rather than the PSA alone. A second complication was that the reported hydrogen production learning rates vary significantly (0.11 and 0.27). A third consideration in the learning rate decision was that all gas processing technologies in Table 6 (with exception of the SMR value reported by [55]) are in the order of 0.10 to 0.14, while only the liquids processing technologies show higher learning rates (ammonia: 0.29; bulk polymers: 0.18–0.37). Given these three considerations, for the ESA technology we also adopted a learning rate with a nominal value of 0.11, and applied a minimum of 0.10 and a maximum of 0.14 in the sensitivity analyses (see Section 4.3).

4.2.1.4. Power plant capital costs. The NGCC capital costs could be calculated using the exponent method because a number of reliable capital cost estimates exist for this technology. The NGCC capital costs in this work were based on those estimated in the EBTF study [58] since we also applied the technical NGCC design of the EBTF. To calculate the TPC for an NOAK reference NGCC without CCS, the reported EBTF bottom-up EPC costs were multiplied with 20% project contingencies. Process contingencies were excluded because this type of power plant was commercially available at the time of the EBTF study.

For the NGCC with CCS cases (ESA and MEA), the different equipment of the NGCC plant - GT, HRSG, ST, heat rejection - were scaled to their respective required sizes using Eq. (9). Based on DOE/NETL guidelines [59], an exponent of 0.8 was used for the gas turbines and steam turbines, and exponent of 0.7 was used for the HRSG and cooling water sections. After scaling, the costs were escalated from 2008€ to 2014€ using the Eurozone Harmonized Index of Consumer Prices (HICP).

4.2.1.5. CO_2 compression capital costs. Finally, the TPC of the CO_2 compression plants was calculated bottom-up using the direct estimation method, because this technology is considered mature and has already gone through the learning curve shown in Fig. 3 [53].

4.2.2. Operations and maintenance costs

Operational costs of the CO_2 capture, EGR, and CO_2 compression equipment were estimated including labour (1 extra shift and 1 extra process technologist in comparison to NGCC w/o CCS), maintenance costs, and variable costs (e.g. process water) (Table 7). Calculation of sorbent costs and replacement frequency are provided in Appendix C.

For the power plant, labour costs were taken from the EBTF study [14] and escalated to 2014€ using the HICP. Maintenance and insurance costs were calculated as a percentage of TPC (Table 7). The costs of water make-up and fuel were calculated based on their respective feed flows and multiplied with unit costs (Table 7). To complete the O & M cost estimate, transport and storage costs were added based on the ZEP reports [67,68] (Table 7), similar to [13].

Table 7
Operating cost assumptions used in this study.

Cost item	Unit	Value
Power plant		
Naturel gas	€/tonne	8,15€/GJ[a]
Labour	2008 M€/a	6[b]
Fixed maintenance costs	% TPC/a	3[b]
Insurance costs	% TPC/a	2[b]
Process water	€/m^3	1[c]
EGR, capture unit, compression unit		
Maintenance	% TPC/a	4[d]
Operators & supervision	k€/a	421,5[e]
Plant technologist	k€/a	100[f]
Process water	€/m^3	1[c]
MEA	€/tonne	2100[g]
Active carbon	€/tonne	1100[h]
Adsorbent monolith	€/tonne	7650[i]
NaOH	€/tonne	2100[c]
Solvent/sorbent disposal	€/tonne	375[c]
Transport & Storage		
Transport (180 km offshore)	€/tonne	6[j]
Storage (offshore depleted oil/gas field)	€/tonne	10[k]

[a] 2014 average industrial consumer price of natural gas in The Netherlands [60].
[b] [14].
[c] [61].
[d] [62].
[e] Wage information retrieved from the Norwegian Confederation of Trade Unions [63]. 1 additional operator assumed in 6 shift rotation.
[f] Wage information retrieved from the Confederation of Norwegian Enterprises [64]. One additional plant technologist assumed.
[g] Based on [65].
[h] [66].
[i] Estimated using Lichtenberg's method.
[j] [67].
[k] [68].

4.3. Sensitivity analyses

The previous sections highlighted some of the methodological choices, trade-offs, and simplifications that need to be made when analysing low TRL CO_2 capture technologies. A low technology development stage inherently leads to many uncertainties in technology performance, and because the use of simplifications is unavoidable, uncertainties in performance results will increase. Study of the available literature, general engineering knowledge, and attention to the particular characteristics of the technology can provide insights into the range of many uncertainties. I.e., an informed, educated value for input uncertainties can be derived, and hence their effect on performance results can be analysed in a sensitivity analysis. This may lead to a preliminary, but substantiated understanding, of potential performance ranges.

The analysis in this case study focused especially on ESA specific parameters but included sensitivity analysis of some generic parameters too, to provide a reference for the ESA specific uncertainties. The inputs to the technical sensitivity analyses are provided in Table 8.

To investigate the uncertainties in ESA capital cost results, the values of EPC cost, N_{FOAK}, LR, process contingencies, and project contingencies where varied. Also, a scenario with maximum EPC costs and contingencies and minimum learning, as well as a scenario with minimum EPC costs and contingencies and maximum learning were calculated (see Section 4.2.1, EPC, FOAK, LR, and contingencies varied simultaneously). This led to a capital cost range as an addition to the point estimate. The LCOE sensitivity analysis also included variation of net system efficiency – a result of the technical sensitivity analysis - and sorbent replacement frequency, as well as more generic economic parameters (Table 9).

5. Results and discussion

This section presents the results of this work. Section 5.1 presents the technical performance of the ESA case study, after which Section 5.2 deals with the methodological insights on technical performance

Table 8
Nominal, low, and high values used for technical sensitivity analysis of the NGCC with ESA system.

Input parameter	Unit	Nominal value	Low value	High value
ESA specific input parameters				
13X adiabatic working capacity	mol/kg	0.8	0.4[a]	1.1[k]
SRD 13X ESA process	GJe/t CO_2	2.04	1.9[b]	3[l]
TRI max adsorbent loading	mol/kg	2.94	2[c]	3[c]
TRI heat capacity	J/kg/K	1000	700[d]	1500[m]
TRI adiabatic working capacity	%	25	18[e]	28[e]
TRI heat of adsorption	kg/mol	70	35[f]	90[f]
13X heat capacity	J/kg/K	900	700[d]	1000[a]
TRI regeneration temperature	C	110	90[g]	120[n]
Additional water vaporization heat	GJ/t CO_2	0.38	0[h]	0.76[o]
13X heat of adsorption	kg/mol	45	35[i]	50[a]
13X gas heating duty	GJe/t CO_2	0.13	0065[j]	0195[j]
NGCC input parameters				
Gas turbine net efficiency	%	38.15	37.15[p]	39.15[p]
LP turbine efficiency	%	88	88[q]	94[u]
Generator efficiency	%	98.5	97.5[r]	99.5[r]
Ambient temperature	C	15	5[s]	20[s]
HP steam pressure	Bar	121	100[t]	150[u]
HP turbine efficiency	%	92	86[u]	92[q]

[a] Educated guess, half of the value used that was derived from [16].
[b] The value of 1.9 GJe/t was calculated using a more advanced adsorption cycle [24].
[c] TRI max adsorpbent loadings were based on the experimental range reported in [28].
[d] Heat capacity of pure activated carbon assumed as minimum value.
[e] The minimum and maximum values of TRI working capacity were based on [25]. This source reports working capacities for amine modified polymeric resins, which were used as a proxy for amine grafted pore expanded silica.
[f] Minimum and maximum values taken from the range reported in [32]. Values depend amongst others on sorbent loading.
[g] Temperature assumed well below sorbent degradation temperature.
[h] Assuming complete heat recovery of the water vaporization energy.
[i] Value derived from [30].
[j] Plus and minus 50% of the nominal value assumed.
[k] Adiabatic working capacity reported in [69].
[l] Fifty percent higher than nominal value assumed.
[m] Heat capacity of amine impregnated solid sorbent used in [41].
[n] Max regeneration temperature due to sorbent degradation restrictions.
[o] No heat recovery of water vaporization energy.
[p] Minimum and maximum reported values for GT efficiency in [14].
[q] Same value as nominal value, other sources often report higher values.
[r] Educated guess, generator efficiency seems well established figure.
[s] Average ambient temperatures as they are estimated to be found within Europe.
[t] Educated guess.
[u] Steam turbine efficiencies and steam pressures as reported in [49].

projection of very early stage technologies. Section 5.3 then shows the economic results of the ESA case and draws conclusions on its economic feasibility, and on very early stage cost estimation in general.

5.1. Technical performance results

Following the approach outlined in Section 4.1.1 and using the factors presented in Table 10, the different contributions to the ESA regeneration duty were calculated. The SRD estimate of the advanced ESA process amounts to 1.9 gigajoule electricity per tonne of CO_2 captured (Table 11). In first instance, this seems low compared to the regeneration duty of the benchmark MEA process (3.6 GJ steam/t CO_2, Table 12). However, because ESA uses electricity instead of steam as energy input for regeneration, the ESA process has a considerable impact on the net output of the NGCC (see Table 12 and Fig. 4). In fact, the net output and net efficiency of the NGCC + ESA system are substantially lower than those of the NGCC + MEA system. This is

Table 9
Nominal, low, and high values used for the economic sensitivity analysis of the NGCC with ESA system.

Parameter	Unit	Nominal value	Low value	High value
Capture plant TPC[a]	M€	508	Mininum	Maximum
Net efficiency[b]	%	49.16	Minimum	Maximum
Fixed labour cost capture plant[c]	M€/a	0.22	0.22	0.44
Fixed labour cost power plant[d]	M€/a	6.55	4.9	8.2
Maintenance cost power plant[e]	%	2.5	1.5	3.5
Maintenance cost capture plant[f]	%	4	2	6
Sorbent replacement frequency[g]	p/a	2	0.5	4
Transport & storage costs[h]	€/t CO2	16	6	34
Fuel price[i]	€/GJ	8.15	6	10
Discount rate[j]	%	7.5	5	10
Life time[h]	y	25	12.5	25
Process contingencies[k]	%	50	30	70
Project contingencies[l]	%	40	30	50
FOAK value[m]	(–)	5	1	10
Learning rate[m]	%	11	10	14

[a] The minimum TPC is a scenario with minimum contingencies and maximum learning: process contingencies equal 30% for ESA (5% for MEA), project contingencies equal 30% for ESA (and MEA), learning starts after the first plant (N_{FOAK} = 1), learning rate is 14%. The maximum TPC is a scenario with maximum contingencies and minimum learning: process contingencies equal 70% for ESA (20% for MEA), project contingencies equal 50% for ESA (and MEA), leaning starts after the tenth plant (N_{FOAK} = 10), learning rate is 10%.
[b] Minimum and maximum net efficiency used from technical sensitivity analysis (Section 5.1).
[c] Nominal and low value equal 1 shift of operators plus 1 process technologist as addition to normal NGCC crew. High value equals 2 operator shifts plus 2 process technologists as addition to normal NGCC crew.
[d] ± 25% (educated guess).
[e] ± 40% (educated guess).
[f] Two to four percent are typical values for maintenance costs [47,61], a maximum value of 6 percent was used to illustrate the high uncertainty of a novel process.
[g] A maximum value of 4 replacements per annum was used assuming a case where max deactivation is 15% instead of 30% (Section 4.2.2), a minimum of 1 replacement per 2 years was used to represent a case where sorbent is further developed to deactivate slower, or where sorbent can be regenerated. This value is more in line with industrial practice where sorbent beds are replaced less frequently. Based on ZEP [67,68].
[h] In techno-economic studies the economic life time is often assumed the same as the technical life time, 25 years in case of an NGCC plant. In reality, operators depreciate their assets faster, and an economic life time of less than 25 years is more realistic. Hence the nominal and maximum value are equal (25 years), and the minimum value is taken as half of that.
[i] Based on Dutch bureau of statistics [70].
[j] Based on [14,49].
[k] See Table 4.
[l] See Table 5.
[m] See Section 4.2.1

Table 10
Factors used in SRD projection of advanced ESA process (using Equation (2) - Equation (5)).

Effect	Unit	Zeolite 13X	TRI-PE-MCM-41	Factor	Factor value
Desorption heat	kJ/mol	45	70	F_1	1.56
Working capacity	mol/kg	0.8	0.75	F_2	1.09
Heat capacity	J/kg/K	900	1000	F_3	1.11
Regeneration T	°C	180	110	F_4	0.52
Gas heating[a]	mol/kg	0.8	0.75	F_5	1.09
Water vaporization	GJ/tCO_2	–	–	F_6	1

[a] The factor is based on the volume, which is a function of working capacity. Therefore the values for F_2 and F_5 are the same.

Table 11
Breakdown of specific regeneration duty calculated using Eqs. (2)–(5) and the factors in Table 10.

SRD item	Unit	Value
Desorption heat	GJ_e/tCO_2	0.66
Sensible heat	GJ_e/tCO_2	0.93
Gas heating	GJ_e/tCO_2	0.14
Water vaporization	GJ_e/tCO_2	0.15
Total SRD	GJ_e/tCO_2	1.90

Table 12
Technical performance indicators of NGCC without capture, with MEA capture, and with ESA capture.

Performance indicator	Unit	NGCC w/o capture	NGCC MEA	NGCC ESA	NGCC ESA 75/25	NGCC ESA 50/50
SRD (steam)	GJ/tCO_2	–	3.66	–	0.48	0.95
SRD (electricity)	GJ_e/tCO_2	–		1.90	1.42	0.95
Gross power output	MW	835	759	835	824	814
Parasitic load	MW	7	43	182	148	113
Net power output	MW	829	716	653	676	701
Gross efficiency	$\%_{LHV}$	58.7	53.3	58.7	57.9	57.2
Net efficiency	$\%_{LHV}$	58.2	50.4	45.9	47.5	49.3
SPECCA	GJ/tCO_2	–	3.15	5.47	4.57	3.67
CO_2 intensity	kg/MWh	348	40	44	43	41

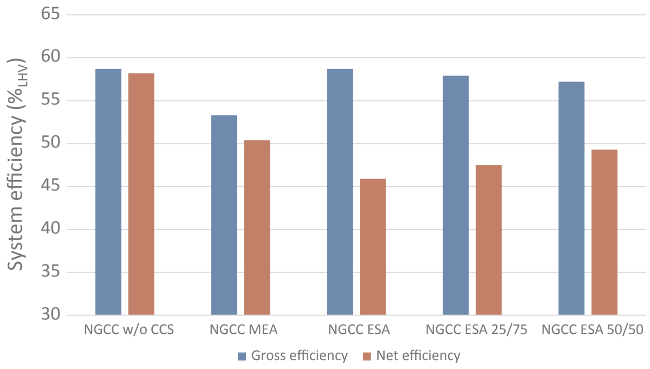

Fig. 4. Graphic representation of gross and net system efficiency of the 5 analysed NGCC systems.

explained by the use of high quality electricity instead of low quality LP steam for regeneration (LP steam is typically transformed to electricity in steam turbines with an efficiency of 20%–30%).

The picture slightly improves if it is assumed that part of the regeneration duty of the ESA system could be supplied by steam instead of electricity. In how far this would be possible is subject to further investigation. The NGCC + ESA net efficiency increases from 45.9% to 47.5% under an assumption of 25% regeneration with steam (NGCC ESA 25/75), and to 49.3% under an assumption of 50% regeneration with steam (NGCC ESA 50/50). This is however still 1 %-point lower than the net efficiency of and NGCC with MEA configuration. From a technology perspective, these results thus show that NGCC with ESA capture does not have an energetic advantage over NGCC with MEA.

Fig. 5 presents the sensitivity analysis of SRD and net system efficiency of the NGCC ESA 50/50 system. The SRD sensitivity analysis shows that the base SRD value is on the low end and that it is likely to be higher than lower. The mainly higher SRD values lead to mainly lower values of net system efficiency: this indicator could be reduced with over 3 %-point compared to the base case. Also, the ESA input parameters generally have a higher impact than the NGCC parameters, stressing the importance of the SRD estimate.

5.2. Methodological insights on future technical performance projection

The results show that it is possible to make a rough but sound technical performance projection of a future advanced ESA process, despite its low TRL. This is mostly due to the availability of results from rigorous modelling (Section 4.1.1) of a preliminary ESA process, and the availability of lab data for advanced solid adsorbents. Two general methodological insights can be extracted from this. First, at least some level of rigorous modelling work - or laboratory SRD measurements - are required to produce meaningful technical performance estimates of CCS technologies. And second, basic lab data of advanced sorbents need to be available to project the future performance. These requirements are unlikely to be met at TRL < 3. It is therefore unlikely that performance estimates of lower TRL technologies will lead to reliable results.

Furthermore, the results showed the importance of system analysis for understanding the performance of capture technologies. In the case of ESA, the capture technology has been reported as very promising due to its low regeneration duty [16]. This conclusion was based on the performance of the capture unit alone. However, results change when the full system is analysed, indicating that this system cannot compete with MEA technology. CCS technology screening thus requires the connection between power plant and capture plant, and is otherwise inconclusive.

Last, the SRD sensitivity analysis showed that the use of simplified models may lead to physically impossible outputs. In this sensitivity analysis the SRD sometimes went below the ideal CO_2 separation energy. This stresses the care that must be taken when projecting technical performance with simple methods.

5.3. Economic performance

At the interface of technical and economic evaluation is the equipment sizing and costing. Table 13 shows that for the studied ESA design a total of 16 columns were required, divided over 4 trains. The total amount of adsorbent required to fill these columns equalled 5936 tonnes, 371 tonnes per column. Table 13 also gives the costs of the columns and of the other equipment.

Based on the estimated equipment costs, Fig. 6 shows the progression of capital costs from currently estimated EPC, to FOAK TPC, and finally NOAK TPC. For the ESA plant, the addition of process and project contingencies escalates the EPC from 305 M€ to first-of-a-kind TPC of 641 M€. Technological learning then reduces the FOAK TPC to an NOAK TPC of 508 M€. This is 19% higher than when the TPC was estimated with the direct costing method. For the MEA plant, the indirect TPC estimate is actually 11% *lower*.

Fig. 6 also shows the uncertainty ranges of the direct and indirect capital cost estimates. The ranges applied to the direct TPC estimates are simply the -30%/+50% accuracy which is typical for an AACE class 4 estimate. Because the AACE does not specify a typical accuracy for EPC costs (before the addition of contingencies), the same range was used to display the margin on EPC. The ranges of the indirect NOAK TPC estimates doubled in size compared to the EPC margins, because the uncertainty in contingencies and learning was added (outside uncertainty ranges in Fig. 6). If no uncertainty on the EPC was assumed, the NOAK TPC margins were around ± 40% (inside uncertainty ranges in Fig. 6).

Finally, Fig. 6 shows that the ESA capital costs are likely to be higher than the MEA capital costs. This means that also from an investment cost perspective ESA is likely the lesser option when compared to postcombustion solvent technology.

From a methodological point of view, the capital cost analysis shows

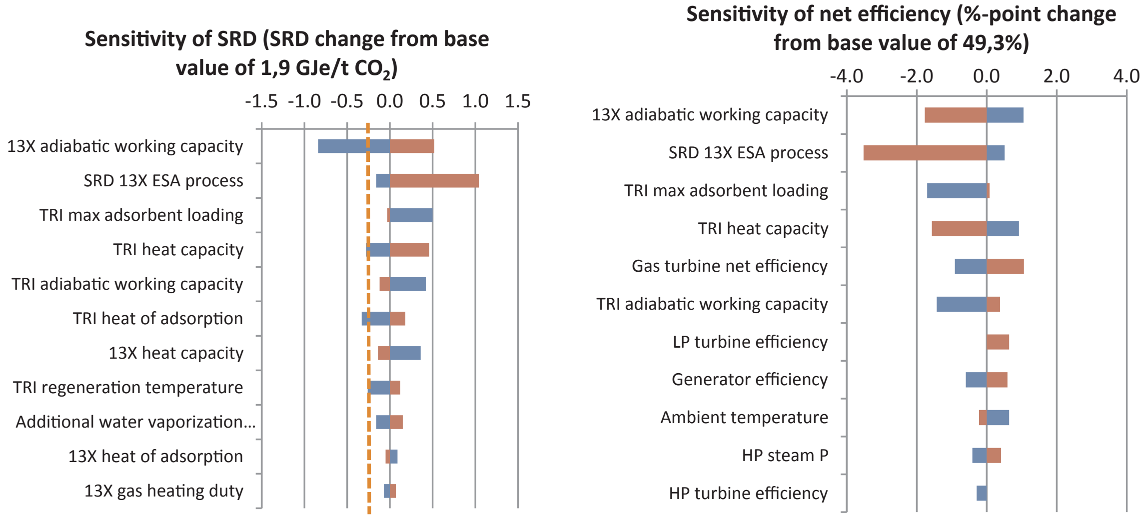

Fig. 5. Sensitivity analysis of ESA specific regeneration duty (l) and NGCC + ESA net LHV efficiency (r). The orange dotted line in the left graph depicts the minimum ideal separation energy of CO_2 from flue gas at a partial pressure of 64 mbar (based on [71]).

Table 13
Main ESA equipment amounts and costs. Other equipment includes EGR equipment, fans, FG/CO_2 storage tanks and heaters/coolers.

Equipment	Amount	Purchased equipment costs (M€$_{2014}$)	EPC costs (M €$_{2014}$)
Adsorption column	16	4,1	150
Adsorption monolith	16	514	73
Valves (incl. instrumentation)	96	0036	10
Other equipment			73

that NOAK TPC estimates using the direct estimation method are close to the nominal value calculated with the indirect method. The methods thus produce similar results, but their validity can only be proven with cost data from real plants. When looking in more detail, we observe that for the low TRL technology (ESA), the indirect estimate is higher, thus more conservative, than the direct estimate, but for the higher TRL technology (MEA), the indirect estimate is lower. This may indicate that the indirect method is particularly suited for low TRL technologies (TRL $<$ 6), but too optimistic for higher TRL technologies. To corroborate this tentative insight, more examples and case studies at different TRL would be required.

Looking into detail of the sensitivity of TPC to the individual input parameters, Fig. 7 highlights that especially the values of the EPC estimate are relevant to the final value of TPC. This means that the indirect method, like the direct method, relies heavily on accurate bottom up estimation of equipment costs and installation, and the other costs included in the EPC estimate. The values used for N_{FOAK} and contingencies have less impact on the value of NOAK TPC. These parameters have an impact of 7–20%. The learning rate only has a minor impact on the TPC result.

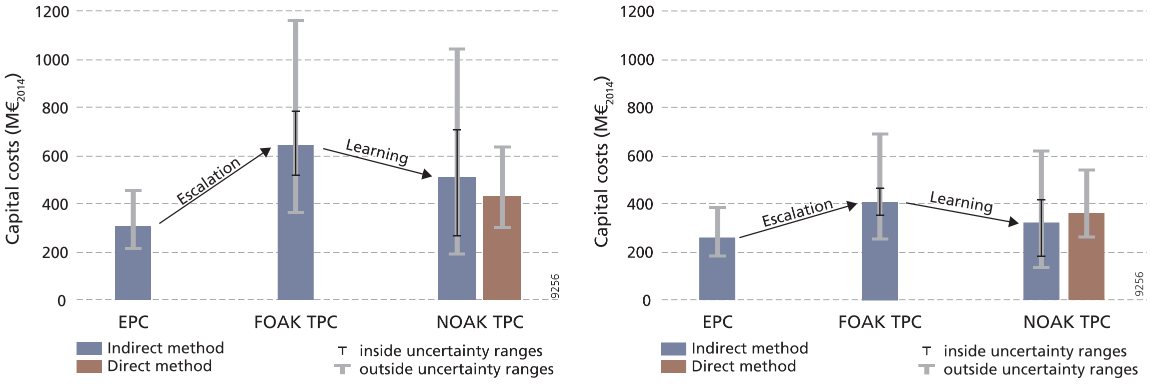

Fig. 6. Capital costs for the postcombustion capture units. ESA (left) and MEA (right). The figure shows the costs results of the indirect (blue) and of the direct (red) capital cost estimation methods. The lower outside uncertainty bounds represent the case of minimum EPC and contingencies and maximum learning. The higher outside uncertainty bounds represent the case of maximum EPC and contingencies and minimum learning. The inner uncertainty bounds represent the same, but without an initial uncertainty for EPC included.

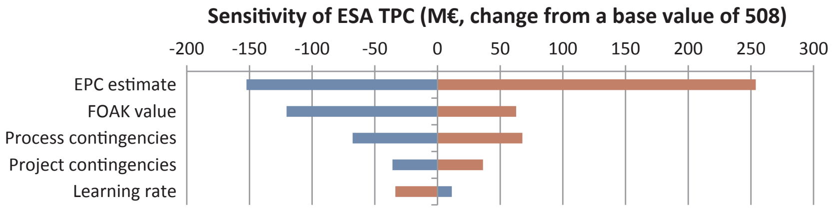

Fig. 7. Sensitivity of Total Plant Cost (calculated using the indirect method).

Table 14
Costs of the NGCC without CCS, with MEA, and with ESA capture. Note that the costs of EGR equipment are included in the costs of the capture units and that the capture plant TPC values are calculated using the indirect method. All cost values are in €$_{2014}$.

Costs item	Unit	NGCC w/o CCS	NGCC ESA 50/50	NGCC MEA
Total Plant Costs				
NGCC	M€	603	574	524
Capture unit	M€		508	326
Compression unit	M€		52	44
O & M costs				
Fuel costs	M€/a	311	311	311
NGCC	M€/a	34	33	31
Capture unit	M€/a		41	19
Compression unit	M€/a		2	2
Transport & storage	M€/a		32	31
Performance indicators high utilisation (85% CF)				
LCOE	€/MWh	65.9	102.6	90.7
CCA	€/t CO_2		119.4	80.7
Performance indicators low utilisation (45% CF)				
LCOE	€/MWh	79.1	134.1	112.7
CCA	€/t CO_2		178.82	109.07

Table 14 presents the economic performance metrics of the NGCC ESA 50/50[2] configuration and a comparison to the NGCC w/o CCS and the NGCC MEA systems. The table shows higher capital costs for NGCC ESA than for NGCC MEA, for all three components of the systems. For the NGCC ESA system, the NGCC costs are higher than in the MEA case because larger steam turbines and heat rejection systems are required. The compression unit is also more expensive because the CO_2 stream is more diluted (4% inerts), and because the CO_2 leaves the capture unit at 1 bar instead of 1.8 bar. The O & M costs of the ESA plant are also higher, amongst others due to high sorbent replacement costs. The higher capital and O & M costs of the NGCC ESA system lead to a 12 point higher LCOE compared to NGCC MEA. The effect of the high capital costs becomes more apparent at low plant utilisation, increasing the LCOE difference between ESA and MEA to 21 points.

Last, Fig. 8 shows that the uncertainty ranges of total plant cost (extracted from Fig. 6) and net efficiency have a significant impact on the levelised cost of electricity. They rank as second and fifth most influential parameters in the LCOE sensitivity analysis, but are outranked by fuel price (and net efficiency also by two general financial parameters). Note however, that the total plant cost ranges are a combination of maximum/minimum EPC, maximum/minimum contingencies and minimum/maximum learning (i.e. EPC, FOAK value, process and project contingencies, and learning rate were varied together). The individual influence on LCOE of these input parameters is much smaller (Fig. 8); they all fall in the lower half of the tornado diagram. So where these separate indicators have a high influence on the capital cost estimate, they have less impact on the LCOE estimate. Operational cost uncertainties like maintenance and labour costs are among the less influential LCOE input parameters.

6. Lessons and conclusions

This work aimed to explore the key challenges when using state-of-the-art methods to analyse the techno-economic feasibility of a very early stage (TRL 3, 4) CO_2 capture technology; to develop other methods that could help overcome these challenges, notably novel hybrid approaches; and to draw lessons on very early stage techno-economic performance assessment, lessons that can also be used for the assessment of other emerging energy technologies. To this end, a case study of CO_2 capture with electric swing adsorption was used and its current (lab stage/model) performance was projected to a future, advanced, commercial ESA process.

As expected, the key challenges lie in constructing performance estimates that represent a commercial process, while the process itself is in very early stage of development. At low TRL, it is challenging to construct a conclusive rigorous process model due to data, time, or resource limitations. This makes it hard to estimate conclusive mass and energy balances, however, it often is possible to produce preliminary mass and energy balances. A challenge following from this is to project these preliminary results to a future state of commercial technology. Also, the use of state-of-the-art direct bottom up capital cost estimation may prove challenging, and may lead to too optimistic results, mainly because process design is still preliminary, prohibiting to identify (the size of) every equipment. Using an indirect bottom up costing method was found to provide a solution, but this came with its own difficulties, for instance identifying technology specific learning rates. A last key challenge identified was to estimate technology specific operational cost estimates, such as sorbent replacement costs, although this work and our previous work [13] found that sorbent/solvent replacement costs are among the less important cost items for feasibility studies.

To address these challenges, it proved useful to apply novel hybrid methods to estimate the future performance of ESA technology, based on the currently available knowledge. For instance, we projected the future technical ESA performance based on earlier reported preliminary (rigorous) modelling results, accounting for improvements in sorbent and process design. In a similar fashion, we projected the capital costs of a commercial ESA plant, based on a preliminary equipment list, and using cost escalations and technological learning. This may also be a solution for other low TRL CCS technologies.

Key lessons on very early stage performance assessments and the use of hybrid assessment methods include the following: first, every CCS technology assessment, also at very early stage, is inconclusive unless it is performed in connection with the CO_2 source (power or industrial plant) and the CO_2 compression plant. Failing to include this may lead to invalid conclusions on feasibility. Second, it is unlikely that classic techno-economic studies can be undertaken for CCS technologies at TRL < 3, because sound SRD estimates based on rigorous modelling or lab measurements are lacking, as well as the required basic data to make development projections. Third, the results of very early stage technology screening come – as expected – with higher uncertainty margins than more advanced technologies. In the case of ESA we found ranges of minus 15% to plus 50% for SRD; and minus 60% to plus 100% for total plant cost. These TPC ranges are double that of commercial technology. Fourth, the direct and indirect capital costing methods employed in this paper provided similar results (max 19% difference). This could be an indication that both methods are valid and reliable, or that they have the same bias. The indirect method provided a more conservative estimate than the direct method for the low TRL technology (ESA), but a more optimistic estimate for the high TRL technology (MEA). From this we tentatively draw that the indirect method may be more suitable for very early stage assessments, but this remains subject to further study. Finally, for CO_2 capture technologies, real plant based learning rates are still lacking, and definitions for “first of a kind” and “Nth of a kind plants” are preliminary. However, in the last two years, four CO_2 capture plants have come online [35,36,72,73], with others currently under construction, providing the opportunity to start tracking cost development, and thus filling knowledge gaps on learning rate and definitions of N_{FOAK} and N_{NOAK}.

Based on the analyses presented in this paper, it can be concluded that electric swing adsorption is infeasible of economically capturing CO_2 from power plants. An advanced, future ESA process was projected to have a specific regeneration duty of 1.9 GJ electricity per tonne of CO_2 (min–max: 1.6–2.9 $GJ_e/t\ CO_2$). This may seem low compared to the standard MEA postcombustion SRD of 3.6 GJ/t CO_2, but because the ESA process uses electricity instead of steam for regeneration, the impact on net power plant efficiency is much larger than that of MEA: 45.9%$_{LHV}$ versus 50.4%$_{LHV}$ for NGCC with ESA and NGCC with MEA

---

[2] The ESA 50/50 configuration was used because this presented the best technical performance.

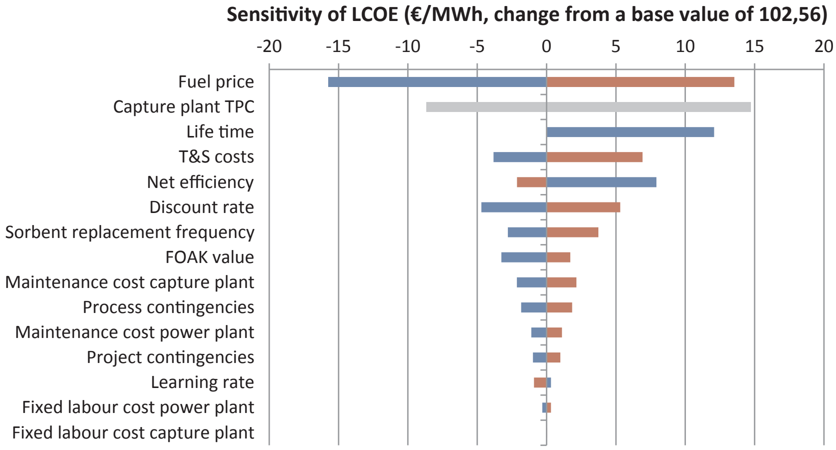

Fig. 8. Sensitivity of LCOE to economic input parameters. Capture plant TPC ranges (grey colour) include simultaneous variation of EPC, process and project contingencies, FOAK value, and learning rate.

respectively. If the ESA regeneration duty can be supplied fifty percent by steam, the NGCC with ESA system efficiency can be cranked up to 49.3%$_{LHV}$, which is still lower than the NGCC with standard MEA system. As a result, it is unlikely that ESA will be able to compete with commercial, second generation solvent systems.

The total plant costs of the Nth of a kind ESA plant have a range from 190 to 1050 million €$_{2014}$, with an expected value around 500 million €$_{2014}$. This is higher than the expected total plant costs of the MEA plant: 323 M€$_{2014}$ (min–max 129–629 M€$_{2014}$). If a configuration is assumed where the ESA regeneration duty is supplied 50/50 by steam and electricity, and the plant is running at base load, the system has an LCOE of 103 €/MWh, versus 91 €/MWh for the NGCC MEA system. At part load the difference is bigger: 134 €/MWh versus 113 €/MWh for NGCC ESA and NGCC MEA respectively.

Concluding, this paper has shown that also for very early stage technologies (TRL 3–4) it is possible to conduct techno-economic studies that give a sound, first indication of feasibility, for instance by using hybrid analysis methods that combine preliminary rigorous estimates with projections of further development. This will help in the process of selecting and progressing the most feasible (CO_2 capture) technologies given constraints on R & D resources, and will aid to the development of a portfolio of the most promising CO_2 capture technologies, as well as other emerging energy technologies.

Acknowledgements

This work is part of the European project EDDiCCUT, which is supported by the Norwegian Research Council under project number 218952 and by an industrial consortium (Bharat Petroleum, Uniper, and Norske Shell). We would like to thank Nils Henrik Eldrup and Ragnhild Skagestad for their help in preparing the cost estimates.

Appendix A. Boundary conditions

See Table A1.

Table A1
Boundary conditions used in this case study, based on EBTF [14] and IEAGHG [49].

Parameter	Unit	Value
Ambient conditions		
Ambient temperature	°C	15
Ambient pressure	bar	1013
Ambient humidity	%	60
Cooling water temperature	°C	12
Cooling water max T increase	°C	7
Cooling water pressure	bar	2.5
CO_2 stream conditions		
CO_2 capture rate	%	90
CO_2 pipeline pressure	bar	110
CO_2 purity	%	96
CO_2 water content	ppm$_v$	< 200
Emissions settings		
SOx (at 6% O_2, dry basis)	mg/m^3	85
NOx (at 6% O_2, dry basis)	mg/m^3	120
Particles (at 6% O_2, dry basis)	mg/m^3	8

Appendix B. Economic performance indicators

Levelised cost of electricity:

$$LCOE\,(€/MWh) = \frac{\sum_{i=1}^{n} \frac{I_i + O\,\&\,M_i}{(1+r)^i}}{\sum_{i=1}^{n} \frac{E_i}{(1+r)^i}} \tag{12}$$

Where I_i is the investment cost in year i, $O\,\&\,M_i$ are the operations and maintenance costs in year i, r is the real discount rate (%), and E_i is the electricity production (MWh) in year i.

Cost of CO_2 avoided:

$$CCA\,(€tCO_2) = \frac{LCOE_{CC} - LCOE_{ref}}{C_{ref} - C_{cc}} \tag{13}$$

Where $LCOE_{cc}$ is the levelised cost of electricity in the plant with CCS, $LCOE_{ref}$ is the levelised cost of electricity in the plant without CCS, C_{ref} is the CO_2 intensity (t CO_2/MWh) in the plant without CCS, and C_{cc} is the CO_2 intensity (t CO_2/MWh) in the plant with CCS.

The economic indicators were calculated for a 25 year lifetime, including 3 years of construction. Because gas-fired combined cycles are used as base loaders as well as cycling plants [74–76], the economics were calculated for a high power plant utilisation scenario (85% capacity factor) and for a low power plant utilisation scenario (45% capacity factor) [13]. All costs are reported as constant (real) costs, using a discount rate of 7,5%, similar to [13].

Appendix C. Calculation of sorbent cost and replacement frequency

Sorbent costs: the ESA sorbent is a mixture of activated carbon (30%wt), SiO_2 (35%wt) and triamine silane (35%wt) formed into a monolith. The costs of activated carbon are well-known, but triamine silane costs were not available. Therefore, a proxy price of the monolith was estimated using Lichtenberg's method [77,78]:

$$P_{monolith}\left(\frac{€}{tonne}\right) = \frac{P_{low} + P_{high} + 3\cdot P_{average}}{5} \tag{14}$$

Where, P_{low} is the price of the cheapest component, P_{high} is the price of the most expensive component, and $P_{average}$ is the average price of the components included in the analysis. The unit costs of four materials were used as basis for the estimate: activated carbon, zeolite, silica gel, and polymeric resin. These were retrieved from the Aspen capital cost estimator (V8.4) and used to estimate the monolith material costs for this study. Monolith manufacturing costs from the raw materials were excluded. Sensitivity analysis (Sections 4.3,5.3) showed that this simplification was justified within the scope of this study.

Replacement frequency: because TRI-PE-MCM − 41 deactivation studies are scarce and/or inconclusive, the deactivation of solid sorbent impregnated with another amine - polyethylene imine (PEI) - was used as a proxy [31]. From reference [31], monolith replacement was estimated to take place twice a year (PEI deactivation rates measured in a lab environment were found to be 1.79% per 300 cycles. Given the cycle duration of 40 min and assuming a max allowable deactivation of 30%, the bed lifetime equals 3360 h, or half a year).

References

[1] IPCC, CLIMATE CHANGE 2014 Synthesis Report Summary for Policymakers, Geneva, 2014.

[2] Abanades JC, Arias B, Lyngfelt A, Mattisson T, Wiley DE, Li H, Ho MT, Mangano E, Brandani S. Emerging CO_2 capture systems. Int J Greenh Gas Control 2015.

[3] (Henry) Liang Z, Rongwong W, Liu H, Fu K, Gao H, Cao F, Zhang R, Sema T, Henni A, Sumon K, Nath D, Gelowitz D, Srisang W, Saiwan C, Benamor A, Al-Marri M, Shi H, Supap T, Chan C, Zhou Q, Abu-Zahra M, Wilson M, Olson W, Idem R, (Pt) Tontiwachwuthikul P. Recent progress and new developments in post-combustion carbon-capture technology with amine based solvents. Int J Greenh Gas Control 2015;40:26–54.

[4] Idem R, Supap T, Shi H, Gelowitz D, Ball M, Campbell C, Tontiwachwuthikul P. Practical experience in post-combustion CO2 capture using reactive solvents in large pilot and demonstration plants. Int J Greenh Gas Control 2015;40:6–25.

[5] Stanger R, Wall T, Spörl R, Paneru M, Grathwohl S, Weidmann M, Scheffknecht G, McDonald D, Myöhänen K, Ritvanen J, Rahiala S, Hyppänen T, Mletzko J, Kather A, Santos S. Oxyfuel combustion for CO_2 capture in power plants. Int J Greenh Gas Control 2015;40:55–125.

[6] US Department of Energy, U.S. DEPARTMENT OF ENERGY Technology Readiness Assessment Guide, 2009.

[7] Junginger M, Van Sark W, Faaij A, Eds., Technological learning in the energy sector. Lessons for policy, industry, and science, 1st ed. Cheltenham, UK: Edward Elgar Publishing Limited, 2010.

[8] Van der Spek M, Arendsen R, Ramirez A, Faaij A. Model development and process simulation of postcombustion carbon capture technology with aqueous AMP/PZ solvent. Int J Greenh Gas Control 2016;47:176–99.

[9] Van der Spek M, Ramirez A, Faaij A. Improving model uncertainty evaluation by using pedigree matrices. a case study on CO_2 capture with Monoethanolamine. Comput Chem Eng 2016;85:1–15.

[10] Sanchez Fernandez E, Goetheer ELV, Manzolini G, Macchi E, Rezvani S, Vlugt TJH. Thermodynamic assessment of amine based CO_2 capture technologies in power plants based on European Benchmarking Task Force methodology. Fuel Aug. 2014;129:318–29.

[11] Zhai H, Rubin ES. Techno-economic assessment of polymer membrane systems for postcombustion carbon capture at coal-fired power plants. Environ Sci Technol Mar. 2013;47(6):3006–14.

[12] Rubin ES, Short C, Booras G, Davison J, Ekstrom C, Matuszewski M, McCoy S. A proposed methodology for CO_2 capture and storage cost estimates. Int J Greenh Gas Control 2013;17:488–503.

[13] van der Spek M, Sanchez Fernandez E, Eldrup NH, Skagestad R, Ramirez A, Faaij A. Unravelling uncertainty and variability in early stage techno-economic assessments of carbon capture technologies. Int J Greenh Gas Control 2017;56:221–36.

[14] Cesar, D2.4.3. European Best Practice Guidelines for Assessment of CO2 capture technologies, Delft, 2011.

[15] Grande C, Rodrigues A. Electric Swing Adsorption for CO_2 removal from flue gases. Int. J. Greenh. Gas Control 2007;2:194–202.

[16] Grande CA, Ribeiro RPL, Rodrigues AE. CO_2 capture from NGCC power stations using Electric Swing Adsorption (ESA). Energy Fuels 2009;23(11):2797–803.

[17] Tlili N, Grévillot G, Latifi A, Vallières C. Electrical swing adsorption using new mixed matrix adsorbents for CO_2 capture and recovery: experiments and modeling. Ind Eng Chem Res 2012;51(48):15729–37.

[18] Bonalumi D, Lillia S, Manzolini G, Grande C. Innovative process cycle with Zeolite (MS13X) for post combustion adsorption. Energy Procedia 2017;114(November 2016):2211–8.

[19] Meerman JC, Hamborg ES, van Keulen T, Ramírez A, Turkenburg WC, Faaij APC. Techno-economic assessment of CO_2 capture at steam methane reforming facilities using commercially available technology. Int J Greenh Gas Control Jul. 2012;9:160–71.

[20] Versteeg P, Rubin ES. A technical and economic assessment of ammonia-based post-combustion CO_2 capture at coal-fired power plants. Int J Greenh Gas Control Nov. 2011;5(6):1596–605.

[21] Patel AD, Meesters K, den Uil H, de Jong E, Blok K, Patel MK. Sustainability assessment of novel chemical processes at early stage: application to biobased processes. Energy Environ Sci 2012;5(9):8430.

[22] Patel AD, Meesters K, Den Uil H, De Jong E, Worrell E, Patel MK. Early-stage comparative sustainability assessment of new bio-based processes. ChemSusChem 2013;6(9):1724–36.

[23] Joss L, Gazzani M, Mazzotti M. Rational design of temperature swing adsorption cycles for post-combustion CO2 capture. Chem Eng Sci 2017;158(June 2016):381–94.

[24] Grande CA, Ribeiro RPL, Oliveira ELG, Rodrigues AE. Electric swing adsorption as emerging CO_2 capture technique. Energy Procedia 2009;1(1):1219–25.

[25] Grande CA, Ribeiro RPPL, Rodrigues AE. Challenges of electric swing adsorption for CO_2 capture. ChemSusChem 2010;3(8):892–8.

[26] Joss L, Gazzani M, Hefti M, Marx D, Mazzotti M. Temperature swing adsorption for the recovery of the heavy component: an equilibrium-based shortcut model. Ind Eng Chem Res 2015;54(11):3027–38.

[27] Hedin N, Andersson L, Bergström L, Yan J. Adsorbents for the post-combustion capture of CO_2 using rapid temperature swing or vacuum swing adsorption. Appl Energy 2013;104:418–33.

[28] Harlick PJE, Sayari A. Applications of pore-expanded mesoporous silica. 5. triamine grafted material with exceptional CO_2 dynamic and equilibrium adsorption performance. Ind Eng Chem Res 2007;46(2):446–58.

[29] Sjostrom S, Krutka H. Evaluation of solid sorbents as a retrofit technology for CO_2 capture. Fuel Jun. 2010;89(6):1298–306.

[30] Bahamon D, Vega LF. Systematic evaluation of materials for post-combustion CO_2 capture in a Temperature Swing Adsorption process. Chem Eng J 2016;284:438–47.

[31] Heydari-Gorji A, Sayari A. Thermal, oxidative, and CO_2-induced degradation of supported polyethylenimine adsorbents. Ind Eng Chem Res 2012;51(19):6887–94.

[32] Belmabkhout Y, Sayari A. Effect of pore expansion and amine functionalization of mesoporous silica on CO_2 adsorption over a wide range of conditions. Adsorption 2009;15(3):318–28.

[33] Sipöcz N, Tobiesen FA. Natural gas combined cycle power plants with CO_2 capture – Opportunities to reduce cost. Int J Greenh Gas Control 2012;7:98–106.

[34] Turi DM, Ho M, Ferrari MC, Chiesa P, Wiley DE, Romano MC. CO2 capture from natural gas combined cycles by CO2 selective membranes. Int. J. Greenh. Gas Control 2017.

[35] Preston CK. INTEGRATED CARBON CAPTURE AND STORAGE PROJECT AT SASKPOWER'S BOUNDARY DAM POWER STATION," Cheltenham, 2015.

[36] A. Borisoff and D. Lin, "Petra Nova – Capturing the future," Project Finance International. Thomson Reuters, no. PFI yearbook 2015, 2015.

[37] van der Spek M, Ramirez A. A structured approach for selecting carbon capture process models. a case study on monoethanolamine. Energy Procedia 2014;63:1287–95.

[38] Ribeiro RPPL, Grande CA, Rodrigues AE. Activated carbon honeycomb monolith – Zeolite 13X hybrid system to capture CO_2 from flue gases employing Electric Swing Adsorption. Chem Eng Sci 2013;104:304–18.

[39] Lei M, Vallieres C, Grevillot G, Lati MA. Thermal swing adsorption process for carbon dioxide capture and recovery : modeling, simulation, parameters estimability, and identification. Ind Eng Chem Res 2013;52:7526–33.

[40] Luberti M, Friedrich D, Brandani S, Ahn H. Design of a H_2 PSA for cogeneration of ultrapure hydrogen and power at an advanced integrated gasification combined cycle with pre-combustion capture. Adsorption 2014;20(2–3):511–24.

[41] Zhang W, Liu H, Sun Y, Cakstins J, Sun C, Snape CE. Parametric study on the regeneration heat requirement of an amine-based solid adsorbent process for post-combustion carbon capture. Appl Energy 2016;168:394–405.

[42] Serna-Guerrero R, Belmabkhout Y, Sayari A. Further investigations of CO_2 capture using triamine-grafted pore-expanded mesoporous silica. Chem Eng J 2010;158(3):513–9.

[43] Krutka H, Sjostrom S, Starns T, Dillon M, Silverman R. Post-Combustion CO_2 Capture Using Solid Sorbents: 1 MWe Pilot Evaluation. Energy Procedia Jan. 2013;37:73–88.

[44] Evulet AT, El Kady AM, Branda AR, Chinn D. On the performance and operability of GE's dry low NO_x combustors utilizing exhaust gas recirculation for postcombustion carbon capture. Energy Procedia 2009;1(1):3809–16.

[45] Kvamsdal HM, Rochelle GT. Effects of the temperature bulge in CO_2 absorption from flue gas by aqueous monoethanolamine. Ind Eng Chem Res 2008;47:867–75.

[46] Christensen P, Dysert LR. AACE International Recommended Practice 18R–97 - Cost Estimate Classification System – As Applied in Engineering, Procurement, and Construction for the Process Industries. West Virginia: Morgantown; 2005.

[47] Towler G, Sinnot R. Capital Cost Estimating. Chemical Engineering Design: principles, practice and economics of plant and process design. Second edi. Oxford: Butterworth-Heinemann; 2013.

[48] Gerrard AM. Guide to Capital Cost Estimating. Fourth Edi. Rugby: IChemE; 2000.

[49] IEAGHG, "CO2 Capture at Gas Fired Power Plants, 2012/8," Cheltenham, 2012.

[50] Rubin ES. Evaluating the Cost of Emerging Technologies. Presentation to the Climit workshop on emerging CO_2 capture technologies. Oslo: Climit; 2016.

[51] Rubin ES. Seven Simple Steps to Improve Cost Estimates for Advanced Carbon Capture Technologies. Presentation to the DOE transformational carbon capture technology workshop. Pittsburgh, Pa: DOE NETL; 2014.

[52] Serna-Guerrero R, Sayari A. Modeling adsorption of CO_2 on amine-functionalized mesoporous silica. 2: Kinetics and breakthrough curves. Chem Eng J 2010;161(1–2):182–90.

[53] van den Broek M, Hoefnagels R, Rubin E, Turkenburg W, Faaij A. Effects of technological learning on future cost and performance of power plants with CO_2 capture. Prog Energy Combust Sci 2009;35(6):457–80.

[54] Greig C, Garnett A, Oesch J, Smart S. Guidelines for Scoping and Estimating Early Mover CCS Projects, Brisbane, 2014.

[55] Rubin ES, Yeh S, Antes M, Berkenpas M, Davison J. Use of experience curves to estimate the future cost of power plants with CO_2 capture. Int J Greenh Gas Control 2007;1(2):188–97.

[56] Schoots K, Ferioli F, Kramer GJ, van der Zwaan BCC. Learning curves for hydrogen production technology: an assessment of observed cost reductions. Int J Hydrogen Energy 2008;33(11):2630–45.

[57] Patel M, Weiss M, Simon T, Ramirez A. The chemical sector. In: Junginger M, Van Sark W, Faaij A, editors. Technological learning in the energy sector. Lessons for policy, industry, and science, 1st ed. Cheltenham, UK: Edward Elgar Publishing Limited, 2010.

[58] Manzolini G, Sanchez Fernandez E, Rezvani S, Macchi E, Goetheer ELV, Vlugt TJH. Economic assessment of novel amine based CO_2 capture technologies integrated in power plants based on European Benchmarking Task Force methodology. Appl Energy Jul. 2015;138:546–58.

[59] DOE/NETL, "Quality guidelines for energy systems studies. Capital Cost Scaling Methodology," Pittsburgh, PA, 2013.

[60] CBS, Steam coal; import price from non-EU countries, Centraal Bureau voor de Statistiek, 2015. [Online]. Available: http://statline.cbs.nl/StatWeb/publication/?DM=SLEN&PA=37215ENG. [Accessed: 02-Sep-2015].

[61] Sinnot RK. Coulson & Richardson's Chemical Engineering Volume 6. Fourth Edi. Oxford: Butterworth-Heinemann; 2004.

[62] Melien T, Brown-Roijen S, Economics. In: Carbon dioxide capture for storage in deep geologic formations – results from the CO2 capture project. Volume three: advances in CO2 capture and storage technology. Results (2004-2009), 1st Ed., L. I. Eide, Ed. Newbury, UK: CPL Press, 2009, p. 470.

[63] LO, Wage information Norway, Norwegian confedaration of trade unions, 2014. [Online]. Available: www.lo.no/lonn–tariff. [Accessed: 04-Sep-2015].

[64] NHO, Information about wages in Norway, Organisation of Norwegian Employers, 2014. [Online]. Available: https://www.nho.no/en. [Accessed: 04-Sep-2015].

[65] NETL, Cost and Performance Baseline for Fossil Energy Plants Volume 1a: Bituminous Coal (PC) and Natural Gas to Electricity Revision 3, Pittsburgh, 2015.

[66] Shao R, Stangeland A. Amines Used in CO2 Capture - Health and Environmental Impacts, Oslo, 2009.

[67] Zep, The Costs of CO2 Transport. Post-demonstration CCS in the EU, Brussels, 2011.

[68] Zep, The Costs of CO2 Storage: Post-demonstration CCS in the EU, Brussels, 2011.

[69] Clausse M, Merel J, Meunier F. Numerical parametric study on CO_2 capture by indirect thermal swing adsorption. Int J Greenh Gas Control 2011;5(5):1206–13.

[70] CBS, Natural gas and electricity, average prices for industrial end consumers, Centraal Bureau voor de Statistiek, 2017. [Online]. Available: http://statline.cbs.nl/Statweb/publication/?DM=SLNL&PA=81309NED&D1=5&D2=0&D3=0&D4=4,9,14,19,24,29,34,39,44&VW=T. [Accessed: 24-Jan-2017].

[71] Svendsen HF, State of the art on post combustion capture technologies, no. June. Amsterdam, 2014.

[72] IEAGHG, CCS Cost Network, 2016 WORKSHOP, 2016/09, Cheltenham, 2016.

[73] Reiner DM. Learning through a portfolio of carbon capture and storage demonstration projects. Nat Energy 2016;1(1):15011.

[74] Van der Spek M, Manzolini G, Ramirez A. New approach to techno-economic assessment of power plants with CCS. The inclusion of realistic dispatch profiles to calculate techno-economics of part load operations. Energy Fuels 2016;31(1):1047–9.

[75] ENTSOE, Information on National Generating Capcities and national/European energy balances, ENTSOE Transparancy Platform, 2015. [Online]. Available: https://transparency.entsoe.eu/. [Accessed: 05-Aug-2016].

[76] Brouwer AS, Van Den Broek M, Seebregts A, Faaij A. Impacts of large-scale Intermittent Renewable Energy Sources on electricity systems, and how these can be modeled. Renew Sustain Energy Rev 2014;33:443–66.

[77] Lichtenberg S. The Successive Principle – a scientific crystal ball for management. Int Roundup 2006;18(4):1–6.

[78] Lichtenberg S. Projekt planlaegning i en foranderlig verden. 3rd ed. Kongens Lyngby: Polyteknisk Forlag, DTU; 1990.