Available online at www.sciencedirect.com

ScienceDirect

Energy Procedia 63 (2014) 217 – 224

Energy Procedia

GHGT-12

A systematic method for membrane CO_2 capture modeling and analysis

Simon Roussanaly*[,a], Karl Lindqvist[a], Rahul Anantharaman[a], Jana Jakobsen[a]

[a] *SINTEF Energy Research, Sem Sælandsvei 11, NO-7465 Trondheim, Norway*

Abstract

This work presents a new systematic methodology for the design and optimization of membrane systems for CO_2 capture integrating both technical and cost models. In this methodology, graphical solutions to the separation problem are generated to design a cost-optimal membrane system (process configuration, operating conditions) that satisfy CO_2 capture ratio and product purity requirements.

The methodology developed is illustrated through the design of a post-combustion CO_2 capture membrane system installed on an ASC power plant and its comparison with a MEA capture unit. This cost-optimal design of the membrane system leads to a levelized cost of electricity (LCOE) of 94 €/MWh which is 58% more expensive than the plant without capture and at the same price level as the reference plant with MEA CO_2 capture. The subsequent CO_2 avoided cost is evaluated to be 53 €/$tCO_{2,avoided}$ for both the membrane and MEA CO_2 capture system.

Finally a comparison between the cost model considered and models available in the literature is performed in order to demonstrate that the competitiveness of the membrane system designed in this paper is due to an improved design and not a possible underestimation of the membrane capture cost.

Peer-review under responsibility of the Organizing Committee of GHGT-12

Keywords: *Carbon Capture and Storage (CCS); CO_2 capture;Post-combustion membrane; Cost evaluation.*

* Corresponding author. Tel.: +47 474 41 763; fax: +47 735 97 250.
E-mail address: simon.roussanaly@sintef.no

Peer-review under responsibility of the Organizing Committee of GHGT-12
doi:10.1016/j.egypro.2014.11.023

1. Introduction

Carbon Capture and Storage (CCS) is regarded as one of the most promising technologies for reducing anthropogenic greenhouse gas emissions, and is projected to provide 14% of the lowest-cost solutions for reductions in man-made GHG emissions in 2050 [1]. However the economic viability of power plants and industrial processes with carbon dioxide capture is affected by the significant energy penalty associated with CO_2 capture. Reducing energy penalty of CO_2 capture has therefore been important topic of research in Carbon Capture and Storage.

Gas separation membranes are considered among one of the promising technologies for post-combustion capture and has been studied extensively. Membrane processes are conceptually very simple. However, with existing membrane properties (selectivity and permeability) and other limitations, a single stage membrane process is not feasible to ensure CO_2 purity of 95% and CO_2 capture rate of 90% in the case of post-combustion capture. This results in complicated membrane configurations where numerous design decisions (process configuration, operation conditions and membrane properties) are required to ensure suitable driving force while minimizing work requirement and membrane area.

The literature [2] on process design for post-combustion capture using membranes involves studies where membrane properties and the process configuration is fixed and sensitivity studies are performed on operating conditions to "optimize" the process and then evaluate the cost of this "optimized" system. However similarly to pipeline systems for transport of gases [3, 4], two competing effects take place in the design of CO_2 membrane capture systems: high membrane investment cost for large membrane areas and significant process energy consumption for low membrane areas. Therefore the optimal design of a membrane system should be based on a cost-based engineering optimisation as shown in Figure 1.

The main objective of this work is to present a systematic methodology for the design of membrane systems for CO_2 capture, integrating both technical and cost models. The design problem is solved by generating a set of graphical solutions in a novel *attainable region* approach developed at SINTEF Energy Research [5]. Selecting the best design results in a cost-optimal membrane system (process configuration, operating conditions) that satisfy requirements on CO_2 capture ratio, membrane properties of the system and product purity requirements. The methodology is here illustrated through the design of a post-combustion CO_2 capture membrane system from a coal fired power plant. Results are compared with a Monoethanolamine (MEA) capture unit and the impact of the assumed membrane cost model is discussed.

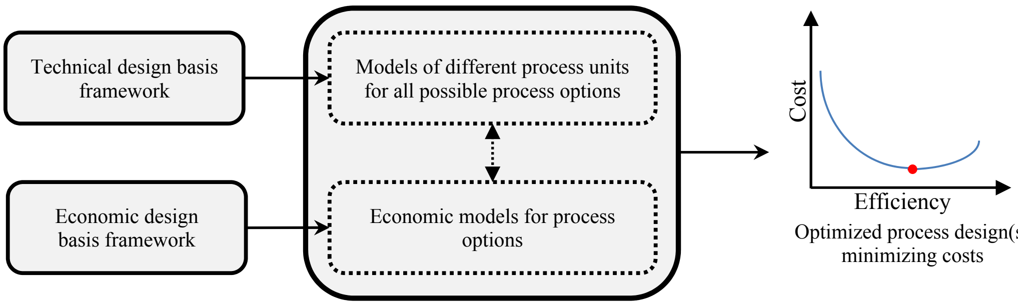

Figure 1: Suggested methodology for design of CO_2 membrane processes based on techno-economic optimization

2. Methodology

This section describes the boundary conditions and the methodology used to assess the characteristics of the CO_2 membrane capture unit after the coal fired power plant while the technical and economic data of the power plant and the reference capture technology, MEA based-CO_2 capture, are directly extracted from the EBTF report [6].

2.1. Technical modelling

2.1.1. ASC power plant and reference capture technology [6]

The power plant considered is based on an Advanced SuperCritical (ASC) boiler and turbine delivering a gross power of 819 MWe and 754.3 MWe (net) without carbon capture. This power plant emits a wet flue gas of 781.77

kg/s containing 13.73 $\%_{wet,vol}$ of CO_2 (equivalent to 15.21 $\%_{dry,vol}$). The yearly average utilization rate of the plant and capture units is 85% in order to account for planned and unplanned maintenance requirements.

The reference capture concept considered a conventional amine scrubbing post-combustion CO_2-capture process based on a basic absorption-desorption process using a 30%wt MEA solvent. The MEA capture stage was designed for a CO_2 capture rate of 90% of the CO_2 contained in the flue gas. After CO_2 capture, the CO_2 is conditioned to reach the pipeline requirements of 110 bar and 25°C. The characteristics of the power plant without and with MEA CO_2 capture are presented in Table 1.

Table 1: Electricity power and emissions of the ASC power plant without and with MEA CO_2 capture [6]

Parameter	Without capture	With MEA capture
Gross electricity power output (MWe)	819	684.2
Auxiliary power consumption (MWe)	65	135
Net electicity (MWe)	754	549.2
CO_2 emitted (kg/MWh)	763	104.7

2.1.2. Membrane CO_2 capture [5, 7]

2.1.2.1. Design methodology

A graphical methodology for systematic and consistent design of membrane processes for post-combustion capture has previously been developed at SINTEF Energy Research [5]. This methodology is now applied to the coal power plant case to design a simple, cost-optimal membrane process with a high CO_2 capture ratio. The membrane separation task is divided into several stages, where one stage includes a membrane unit as well as its own rotating equipment and intercoolers (cf. Figure 2). A membrane module, a rotating equipment module and a cost model (presented in section 2.2) is used to calculate the technical and economic performance of each step. An *attainable region* diagram is used to visualize the possible operating window of each membrane stage in addition to its optimal operating region. The number of stages and operating points are then easily identified using a step-wise approach similar to the McCabe Thiele diagram. Complex process features, such as retentate recycles or retentate heating before expansion, are not included in the graphical solutions generated.

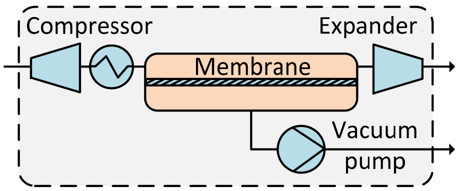

Figure 2: Membrane unit used to create the attainable region diagram

The attainable region diagram is drawn for a certain *stage* carbon capture ratio. This stage capture ratio is determined considering the overall capture ratio to be attained and the approximate number of stages. The overall capture ratio can be subject to economic optimization, but is set to 90% in this work in line with the EBTF report [6]. The minimum allowable permeate side vacuum is set to 0.2 bar.

The total specific capture cost (in €/t CO_2) for *n* separation stages can be calculated from the stage capture costs and the stage capture ratios. Once the design is set, the actual operating conditions (feed pressure, permeate pressure and area) are back-calculated from the targeted stage purity using the membrane model.

2.1.2.2. Membrane and rotating equipment models

The design methodology is dependent on robust models for the membrane separator and the rotating equipment. A membrane model for two gas components, after Saltonstall [8], is adopted for the present work. The model assumes a membrane unit in cross-flow configuration with plug flow on the feed side and no mixing with the bulk stream on the permeate side. These assumptions allow analytical solution of all model equations, which is favorable in terms of robustness and computational speed. A disadvantage of this approach is that water vapor permeation cannot be modeled. As a result, the cost and/or power consumption of drying units before or after the membrane module are included in the present analysis.

Rotating equipment are modeled as isentropic expansions/compressions of an ideal gas. The heat capacity ratio of the binary mixture is calculated from a linear regression as a function of CO_2 concentration. An isentropic efficiency is applied to account for irreversibilities. The approach is accurate to approx. ±1% for compressors, expanders and heat exchanger duties and approx. ±5% for the vacuum pump in the rage of pressures considered.

2.2. Cost modelling

This study assumes costs of a "NOAK" (N^{th} Of A Kind) plant to be built at some time in the future, when the technology is mature. Such estimates reflect the expected benefits of technological learning, but they may not adequately take into account the increased costs that typically occur in the early stages of commercialization [9].
Investment and operating costs are given in 2008 prices which correspond to the reference year for costs in the EBTF report [6] used as reference for the ASC power plant and MEA capture costs. While investment and operating costs of the ASC power plant, as well as the MEA CO_2 capture unit are extracted directly from the EBTF report, the following sections detail the cost methodology used to design and evaluate the CO_2 membrane capture units

2.2.1. Investment costs

A factor estimation method is used in order to estimate investment costs of the process equipment, where the estimated equipment costs are multiplied by direct[†] and indirect[‡] cost factors to obtain the investment costs.

European-based equipment costs function ($€_{2009}$) of carbon steel equipment has been estimated using Aspen Process Economic Analyzer® v7.2, based on simulations performed in the in-house membrane system design code. The investment cost of a given equipment is then calculated by multiplying the component's specific equipment cost by the direct and indirect cost factors (see Table 3).

Table 2: Membrane direct cost, rotating equipment and heat exchanger equipment costs

Type of equipment	Unitary cost	Unit
Membrane module [10]	36	€/m²
Compressor (First stage)	682	€/kW
Compressor (Second stage)	417	€/kW
Compressor (Third stage)	89	€/kW
Expander	414	€/kW
Vacuum pump	77	€/kW
Cooler	293	€/m²

Table 3: Direct and indirect cost factors [6]

Cost factor	Value
Direct Cost Factor	1.77
Indirect Cost Factor	1.31

The total investment cost in $€_{2008}$ is then determined by summarizing the estimated investment cost for all components within defined system boundaries (Equation 1).

$$\text{Total investment cost} = \sum(\text{Equipment cost} \cdot \text{Direct cost factor} \cdot \text{Indirect cost factor}) \tag{1}$$

However due to their specificity, CO_2 membrane framework are estimated differently. Van Der Sluijs et al. [11] suggested a cost function for the membrane framework based on the extrapolation of a membrane separation system in an ammonia plant of DSM . As the membrane separation system in an ammonia plant of DSM operates at 55 bar, simulations has been carried out in Aspen Process Economic Analyzer® to account for the impact of the operating pressure leading to the following equation.

† Which includes the costs of erection, secondary equipment, piping, insulation, and civil work.
‡ Which includes the costs associated with engineering, commissioning, administration, and contingencies.

$$I_{mf} = \left(\frac{A}{2000}\right)^{0.7} \cdot K_{mf} \cdot \left(\frac{Pressure}{55}\right)^{0.875} \quad (2)$$

Where:
A is the overall area of the membrane module in (m^2). It is worth noting that a limitation of 50,000 m^2 of membrane area per module is considered in order to avoid having unrealistically large modules.
I_{mf} is the direct cost of the membrane framework (in €).
K_{mf} is the cost of the reference cost of a membrane framework given for a reference membrane area of 2,000 m^2, and equal to 259 $k€_{2009}$ [11][§].
$Pressure$ is the operating pressure of the membrane module (in bar).

The technical characteristics and costs associated with CO_2 conditioning from 1 to 150 bar are modelled using the BIGCCS transport modules previously presented and illustrated [3, 4].

Finally, the investment costs are reported as an overnight cost occurring at the end of the construction assuming shared investment over the construction time. For instance, power plants and capture facilities are assumed to be built over four years with the annual allocation of project finance over the construction time presented in Table 4.

Table 4: Annual allocation of costs for plan construction [6]

Year	1	2	3	4
Cost share per year (%)	20	30	30	20

2.2.2. Maintenance and operating costs

The operating costs are split into fixed and variable operating costs.

The fixed operating cost depends on the investment cost and covers maintenance, insurance, and labour costs. The annual fixed operating cost is set to 6% of total direct costs for the membrane system and the process units [6]. In addition, fixed operating costs include the cost of replacing the membrane modules every 5 years [12].

The variable operating cost of the CO_2 capture plant are a function of the amount of the amount of CO_2 captured, and covers consumption of utilities: electricity consumption and sea water cooling. The annual variable operating costs are estimated using the utilities consumptions given by the technical modelling of the process and utility costs given in Table 5. It is important to note that an initial electricity cost is required to optimize the membrane system while the actual electricity cost is calculated based on the system costs and electricity consumption.

Table 5: Utility costs

Utilities	Reference costs	Cost Units	Reference year
Electricity [6]	94.5	€/MWh	2008
Sea water cooling [13]	0.02	€/m^3	2008

2.2.3. Project valuation

In order to benchmark the new CO_2 capture alternative to the reference capture technology for CO_2 capture from an ASC and an ASC without capture evaluated in the EBTF report [6], two key performance indicators are employed: the electricity production cost and the cost of CO_2 avoided.

The levelized cost of electricity (€/MWh) is here used as a key performance indicator to measure the unitary cost of the electricity production of a plant with and without CO_2 capture. The electricity production cost approximates the average discounted electricity price over the project duration that would be required as income to match the net

[§] Equal to 238 $k\$_{1991}$.

present value of capital and operating costs for the project. It is equal to the annual costs divided by the annual net electricity production, as shown in equation (3).

$$\text{Levelized cost of electricity} = \frac{\text{Annualized investment + Annual OPEX}}{\text{Annual gross electricity production - Plant auto-consumption}} \quad (3)$$

The CO_2 avoided cost (€/tCO_2), obtained by comparing the levelized cost of the plant with and without the CO_2 capture as shown in equation (4), is also used to compare the two CO_2 capture options. The CO_2 avoided cost approximates the average discounted CO_2 tax or quota over the project duration that would be required as income to match the net present value of additional capital and operating costs due to the CO_2 capture infrastructure. It is worth nothing that at this stage neither transport nor storage costs are considered

$$CO_2 \text{ avoided cost} = \frac{(\mathrm{LCOE})_{\mathrm{CCS}} - (\mathrm{LCOE})_{\mathrm{ref}}}{(t_{CO2}/\mathrm{MWh})_{\mathrm{ref}} - (t_{CO2}/\mathrm{MWh})_{\mathrm{CCS}}} \quad (4)$$

Where $(\mathrm{LCOE})_{\mathrm{CCS}}$ and $(\mathrm{LCOE})_{\mathrm{ref}}$ are respectively the levelized cost of electricity of the plant with and without CCS (€/MWh), while $(t_{CO2}/\mathrm{MWh})_{\mathrm{CCS}}$ and $(t_{CO2}/\mathrm{MWh})_{\mathrm{ref}}$ are respecctively the CO_2 emission rate to the atmosphere of the plant with and without CCS $({}_{CO2}/\mathrm{MWh})$.

3. Results and discussions

3.1. Base case

The design of the membrane CO_2 capture unit lead to a cost-optimal system composed of three membrane stages with the characteristics given in Table 6.

Table 6. Characteristics of the cost-optimal membrane system

Parameter	Membrane stage 1	Membrane stage 2	Membrane stage 3
Stage feed flow (kg/s)	781	294	184
Stage feed CO_2 content ($\%_{vol}$)	15.2	45	80
Stage feed pressure (bar)	2.4	1.5	1
Stage permeate CO_2 content ($\%_{vol}$)	45	80	95
Stage permeate pressure (bar)	0.2	0.2	0.28

Based on the electricity output and the power plant and capture costs, the cost of electricity (LCOE) is used to compare the three power plants cases without or with CO_2 capture as shown in Figure 3. The evaluation shows that the cost of electricity with membrane CO_2 capture is 58% more expensive than the plant without capture and is at the same price level than the reference plant with MEA CO_2 capture. When looking directly at the cost of capturing CO_2, the membrane concept, with a cost of 53 €/t$CO_{2,avoided}$, is also at the same cost level than the MEA capture process.

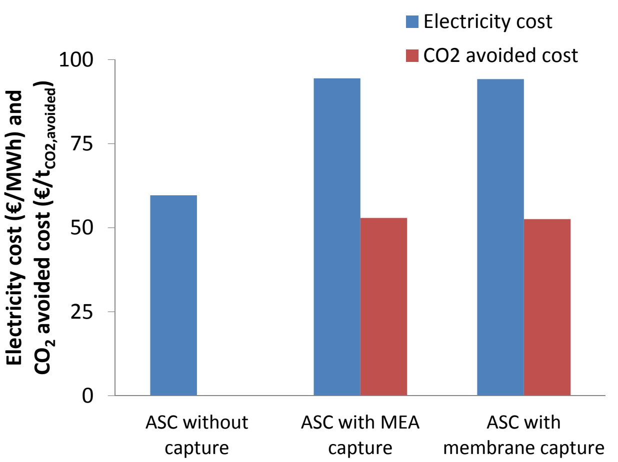

Figure 3: Membrane unit used to create the attainable region diagram

Under the hypotheses considered in this paper, the systematic method for membrane CO_2 capture modeling and analysis seems therefore to lead to a membrane system design which could compete with the MEA technology for the capture CO_2 from the exhaust flue gas of an ASC power plant. It is also worth noting that the membrane systems do not require direct integration with the power plant (no steam integration) hence, start-up, show-down and transient operation can expected to be better for membrane systems.

However, as in the literature [14, 15], cost estimates for membrane CO_2 capture from an ASC power plant often lead to higher costs, it appears necessary to evaluate if this difference in costs is due to an improved design based on the methodology presented in this paper or to a possible underestimation of the input cost data in the methodology.

Figure 4 illustrates the costs of the membrane system presented in Table 6 both with the cost methodology considered in this paper and the cost methodology extracted from Zhai and Rubin [14] . The evaluation show that the cost methodology of the present work leads to costs 9% higher for the same membrane design especially due to higher costs associated with the membrane module (which are due to higher membrane framework and replacements costs in the model considered), as well as higher turbomachinery cost[**]. Therefore the competitive cost of the membrane system design with the methodology presented in this paper is indeed linked to an improved design of the process and cannot be imputed to an underestimation of the cost in assessment.

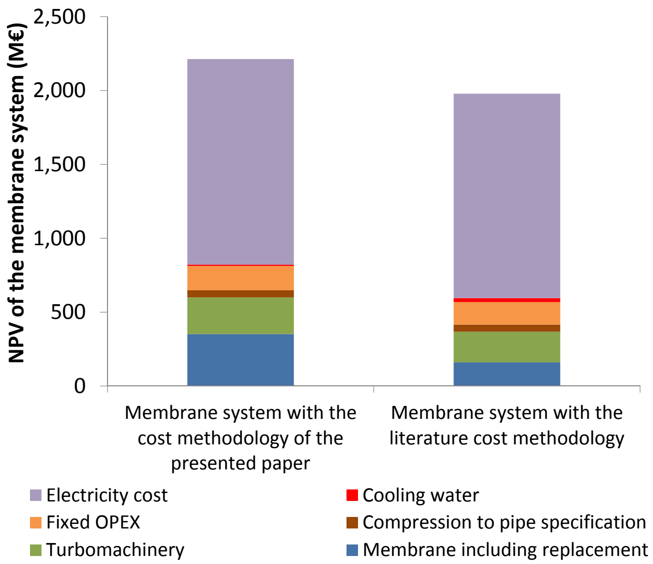

Figure 4: Comparison of the considered methodology and literature cost methodology on the evaluation of the cost-optimal membrane system

4. Conclusions

This work presents a new systematic methodology for the design and optimization of membrane systems for CO_2 capture integrating both technical and cost models. In this methodology, graphical solutions to the separation problem are generated to design a cost-optimal membrane system (process configuration, operating conditions) which satisfy requirements on CO_2 capture ratio and product purity.

The developed tool is here illustrated through the design of a post-combustion CO_2 capture membrane system from an ASC power plant and its comparison with a MEA capture unit. The design of the membrane CO_2 capture unit lead to a cost-optimal system composed of three membrane stages with permeate purities of respectively 45, 80, and 95%. This cost-optimal design of the membrane system leads to a levelized cost of electricity (LCOE) of 94 €/MWh which is therefore 58% more expensive than the plant without capture and at the same price level than the

[**] It is worth noting that even the turbomachinery costs are similar with both models, considered cost model lead to higher compressor costs while vaccum pump costs are lower.

reference plant with MEA CO_2 capture. The subsequent CO_2 avoided cost is evaluated to 53 €/t$CO_{2,avoided}$ for both the membrane and MEA CO_2 capture system.

Finally, a comparison between the cost model considered in this paper and models available in the literature is performed in order to demonstrate that the competiveness of the membrane system designed in this paper is due to an improved design and not a possible underestimation of the membrane capture cost. As a consequence, the systematic method for membrane CO_2 capture modeling and analysis presented in this paper lead to a membrane system design which could compete with the MEA technology to capture CO_2 from the exhaust flue gas of an ASC power plant.

The systematic method for membrane CO_2 capture modeling and analysis seems to have the potential to design improved membrane capture systems for post-combustion emissions. This methodology is expected to be further developed in order to model more complex membrane systems (inclusion of multi-components model, recirculation configuration, etc.), membrane system adapted to pre-combustion cases, and other industrial cases.

Acknowledgements

This publication has been produced with support from the BIGCCS Centre, performed under the Norwegian research program Centres for Environment-friendly Energy Research (FME). The authors acknowledge the following partners for their contributions: ConocoPhillips, Gassco, Shell, Statoil, TOTAL, GDF SUEZ and the Research Council of Norway (193816/S60).

References

[1] International Energy Agency. Technology Roadmap: Carbon capture and storage. 2013.
[2] Zhang X, He X, Gundersen T. Post-combustion Carbon Capture with a Gas Separation Membrane: Parametric Study, Capture Cost, and Exergy Analysis. Energy & Fuels. 2013;27:4137-49.
[3] Roussanaly S, Jakobsen JP, Hognes EH, Brunsvold AL. Benchmarking of CO_2 transport technologies: Part I—Onshore pipeline and shipping between two onshore areas. International Journal of Greenhouse Gas Control. 2013;19C.
[4] Roussanaly S, Brunsvold AL, Hognes ES. Benchmarking of CO_2 transport technologies: Part II – Offshore pipeline and shipping to an offshore site. International Journal of Greenhouse Gas Control. 2014;28:283-99.
[5] Lindqvist K, Anantharaman R. A graphical method for the synthesis of membrane systems for CO_2 capture applications. Chemical Engineering Transaction. 2014;39.
[6] Anantharaman R, Bolland O, Booth N, Dorst EV, Ekstrom C, Franco F, et al. D1.4.3 European best prectice guidelines for assessment of CO_2 capture technologies. DECARBit Project; 2011.
[7] Lindqvist K, Roussanaly S, Anantharaman R. Multi-stage membrane processes for CO_2 capture from cement industry. Submitted to Energy Procedia GHGT-12 Special Edition. 2014.
[8] Saltonstall CW. Calculation of the membrane area required for gas separations. Journal of Membrane Science. 1987;32:185-93.
[9] Metz B, Davidson O, Coninck HD, Loos M, Meyer L. Carbon Dioxide Capture And Storage: IPCC Special Report: Cambridge University Press; 2005.
[10] Merkel TC, Lin H, Wei X, Baker R. Power plant post-combustion carbon dioxide capture: An opportunity for membranes. Journal of Membrane Science. 2010;359:126-39.
[11] Van Der Sluijs JP, Hendriks CA, Blok K. Feasibility of polymer membranes for carbon dioxide recovery from flue gases. Energy Conversion and Management. 1992;33:429-36.
[12] Bhide BD, Stern SA. A new evaluation of membrane processes for the oxygen-enrichment of air. II. Effects of economic parameters and membrane properties. Journal of Membrane Science. 1991;62:37-58.
[13] Haugen H, Anundskås A, Edrup N, Schaatun A, Glittum G, Skagestad R, et al. CO_2 fangst av utslipp fra industrianlegg. Porsgrunn2009.
[14] Zhai H, Rubin ES. Techno-Economic Assessment of Polymer Membrane Systems for Postcombustion Carbon Capture at Coal-Fired Power Plants. Environmental Science & Technology. 2013;47:3006-14.
[15] Zhao L, Menzer R, Riensche E, Blum L, Stolten D. Concepts and investment cost analyses of multi-stage membrane systems used in post-combustion processes. Energy Procedia. 2009;1:269-78.