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Thermo-economic Assessment of Three Different Gas Upgrading Technologies for the Production of Synthetic Natural Gas from Biomass

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Abstract:

Three different gas upgrade technologies for production of synthetic natural gas (SNG) from biomass gasification – amine-based absorption, membrane-based separation and pressure swing adsorption – are investigated for their thermo-economic performance against the background of different possible future energy market scenarios. Preparation of the CO_2-rich stream for carbon capture and storage is investigated for the amine-based absorption and the membrane-based separation technology alternatives. The resulting cold gas efficiency η_{cg} for the investigated process alternatives ranges between 0.65 and 0.695. The overall system efficiency η_{sys} ranges from 0.744 to 0.793, depending on both the gas upgrade technology and the background energy system. Amine-based absorption gives the highest cold gas efficiency whereas the potential for cogeneration of electricity from the process' excess heat is higher for membrane-based separation and pressure swing adsorption. The estimated specific production costs for SNG c_{SNG} vary between 103-127 €$_{2010}$/MWh$_{SNG}$. The corresponding production subsidy level $c_{subsidy}$ needed to achieve end-user purchase price-parity with fossil natural gas is in the range of 56-78 €$_{2010}$/MWh$_{SNG}$ depending on both the energy market scenario and the gas upgrade technology. Sensitivity analysis on the influence of changes in the total capital cost for the SNG plant on the production cost indicates a decrease of about 12% assuming a 30% reduction in total capital investment. Capture and storage of biogenic CO_2 – if included in the emission trading system – only becomes an option at higher CO_2 charges. This is due to increased investment costs, but in particular due to the rather high costs for CO_2 transport and storage that have been assumed in this study.

Keywords:

Synthetic Natural Gas, Thermo-economics, Gas Upgrade, Biomass Gasfication, Modelling, Energy Systems, Biofuels.

1. Introduction

Synthetic natural gas (SNG) from biomass is a promising alternative among gasification-based biofuels. It is close to being commercialized with several companies being on the way to start industrial scale demonstration plants [1-3]. The Gothenburg Biomass Gasification (GoBiGas) project with a 20 MW$_{SNG}$ production unit currently under construction is probably the most prominent example [1]. Several research studies have investigated the process of SNG production from an overall energy efficiency viewpoint aiming at pointing out the process bottlenecks and integration opportunities [4-11]. The focus of the studies has mainly been the influence of both gasification and methanation technology and operating conditions on the overall process performance. Gasification and methanation are identified as the major sources of exergy loss, directly followed by the CO_2 separation system [4] that is inherent for SNG production. Different technology options for CO_2 separation are available, but very limited data on the influence of the choice of CO_2 separation technology on overall SNG process performance is available, as also stated by Gassner [5]. A review of different candidate gas upgrade technologies for upgrade of biomass from fermentation to natural gas grid quality is given by Reppich et al. [12] but the conclusions cannot be transferred directly to SNG production via thermal gasification due to large differences in plant scale and thermal integration opportunities.

Within SNG production, a number of studies [6-8] assume physical absorption technology (e.g. Selexol) as CO_2 removal technology. Vitasari [4] compares physical absorption and cryogenic cooling in an exergy-based study of biomass feedstock influence on the SNG process performance, suggesting that the exergy losses in the CO_2 removal process can be reduced by 50% when using cryogenic cooling. Heyne et al. [9] assume chemical absorption for CO_2 separation in a study proposing integration of SNG production with existing energy service infrastructure. Pressurised water wash – that has been investigated for upgrading biogas from fermentation [13] – has not been investigated within SNG production, probably due to the high CH_4 slip associated. Gassner and Maréchal [10] and Gassner [11] incorporate three different technology options for CO_2 separation into a multi-objective optimisation methodology framework for process design and integration, namely physical absorption, membrane separation and pressure swing adsorption (PSA). The latter methodology is also applied to a detailed design example examining membrane separation for CO_2 separation in the framework of SNG production [5]. Using multi-objective optimisation different design layouts are investigated subject to varying objective functions. Gassner shows that optimising the separation system in isolation from the rest of the process can lead to sub-optimality both from energetic and economic viewpoints for the overall SNG process. In the latter study [5], CO_2 capture for storage is investigated since the CO_2 stream separated is of high purity. Capture costs are estimated to be in the range of 15 to 40 $€_{2006}$/ton, largely dependent on the electricity price. This latter concept of realizing negative CO_2 emissions in a biomass-based SNG production plant at low CO_2 avoidance costs is also investigated by Carbo et al. [8]. According to their study CO_2 separation costs of about 62 $€_{2010}$/ton can be expected for an n^{th} bio-SNG plant with 500 MW_{th} input, when biogenic CO_2 emissions are included in the CO_2 emission trading system.

In this paper, a detailed comparison of three alternative CO_2 separation technologies – amine-based chemical absorption (MEA), membrane-based separation, and pressure swing adsorption (PSA) – is presented, highlighting the advantages and drawbacks of the alternatives from a techno-economic viewpoint. Opportunities for CO_2 capture are investigated and the production costs for SNG from biomass gasification are assessed using future energy market scenarios. An analysis of the sensitivity or the production costs for SNG to changes in commodity prices (e.g. fuel and electricity prices) is presented.

2. SNG production process and quality requirements

The SNG production process layout considered is illustrated in Fig. 1. It is based on an indirect gasification step using a mixture of recycled product gas and superheated steam as fluidising agent. The woody biomass feed is pre-treated by low-temperature air drying reducing its moisture content from 50 to 20 wt-%. The gas cleaning chain consists of a tar reformer, a bag house filter and a water scrubber removing tar components, particles and gas impurities such as NH_3 and H_2S. The cleaned product gas is heated to the methanation temperature and steam is added for a simultaneous methanation and water-gas shift reaction in a fluidised bed catalytic methanation reactor. Details on the process modelling assumption are given in [9]. The raw SNG after methanation basically is a mixture of CH_4 and CO_2 containing impurities such as CO, H_2, N_2 and H_2O. In order to comply with grid specifications the two mandatory removal steps in the gas upgrade section are CO_2 and H_2O removal. Depending on the gas upgrade technology, a H_2-rich off-stream is available and recycled to the methanation step, or off-gases are available for heat generation in a furnace.

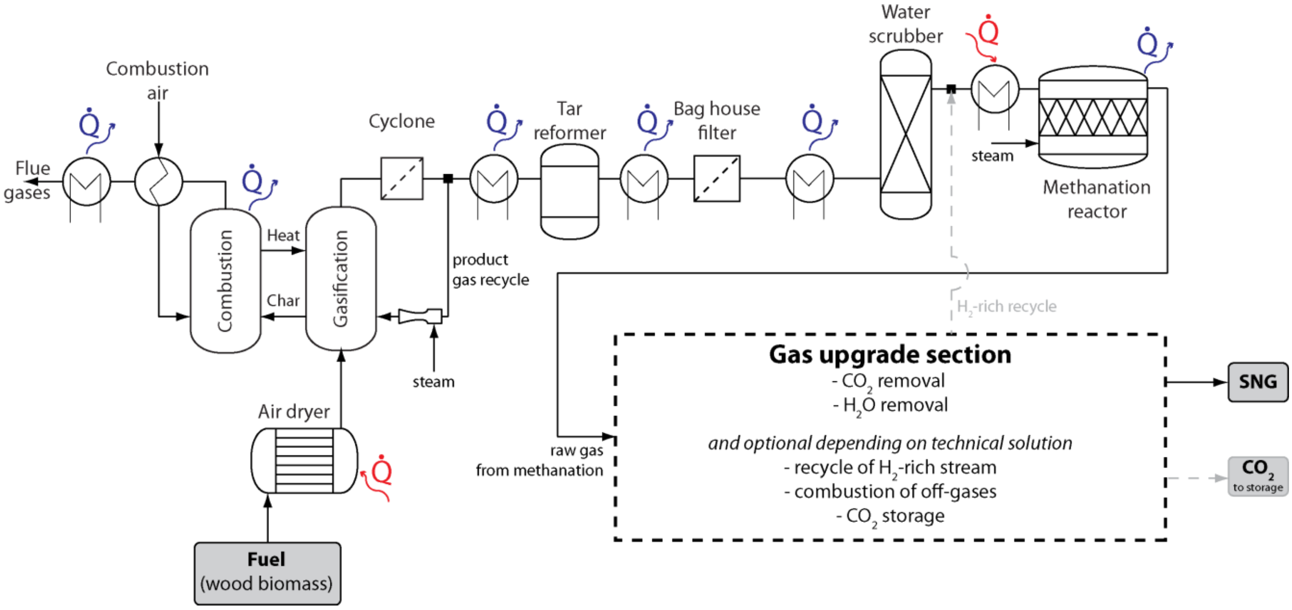

Figure 1: SNG production process flowsheet using indirect gasification with indication of heat sources and sinks ($\dot{Q}$) for process integration and steam generation.

Currently, there is no European standard on quality requirements for the injection of methane or natural gas from unconventional sources into the natural gas network. A mandate has been issued to the European Committee for Standardisation (CEN) [14], and the issue is being addressed by a new committee [15]. In current practice, national standards are used such as the Swedish standard for biogas as a motor fuel [16]. The latter standard is even used for specifying the quality of gas to be injected to the grid. In this study, the specifications that have to be met by the gas upgrade section are based on recommendations of the European Association for the Streamlining of Energy Exchange – gas (EASEE-gas) [17] given in Table 1. The final pressure and temperature of the SNG were set to 60 bar and 20 ºC assuming connection to a transmission pipeline.

For the CO_2 capture cases the specifications for the pressure and temperature of the CO_2-rich stream delivered for transport have been set to 75 bar and 20 ºC with a maximum water content of 500 ppm based on a project investigating carbon capture and storage in the Skagerrak/Kattegat region [18].

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Table 1: Synthetic natural gas (SNG) specifications according to EASEE-gas [17] used as basis for the design study.

Parameter	Unit	Min	Max
WI[a]	kWh/m^3	13.76	15.81
ρ_{rel}	[-]	0.555	0.700
CO_2	mol-%	-	2.5
H_2O	ppm	-	≈48[b]

[a] Reference state: 0ºC, 1,01325 bar, HHV-based (HHV-reference temperature of 25ºC used) [17]

[b] Corresponding level for dew point temperature of H_2O at 70 bar of -8ºC [17] based on partial pressure of water in saturated humid air at -8ºC ($3,35 \cdot 10^{-3}$ bar [19])

3. Methodology

The evaluation and comparison of the different gas upgrading technologies is done based on an existing model for SNG production [9]. Modelling in Aspen Plus [20] is used to obtain the mass and energy balances of the process. The available excess heat from the process streams is used for cogeneration of electricity using a steam Rankine cycle. Investment costs for the overall process are estimated and production costs of SNG are estimated using future energy market scenarios. Details on the different steps are given in the following sections.

3.1. Process modelling and estimation of cogenerated electricity

Heat streams are available from both the basic raw SNG production process as well as from the upgrading section. This heat can be used to cover heat demand internally. Additional excess heat can be used for co-generation of electricity using a steam cycle. To estimate the amount of mechanical work that can be co-generated from the available excess heat the approach proposed by Dhole and Linnhoff [21, 22] based on Pinch analysis is adopted. Based on an exergy-based representation of all heat streams – the so-called Exergy or Carnot Grand Composite Curves (GCC) – the theoretical potential for mechanical work extraction for a given process can be determined by integration of the area below the Carnot GCC. Given the exergetic efficiency η_{ex} of the turbine system that indicates the ability of a real system to harness the theoretical potential for shaft power output, it is possible to obtain the actual shaft work generated. Linnhoff and Dhole [21] demonstrate that this approach gives good results with an error below 2% compared to detailed simulations for the example of shaftwork calculations for a refrigeration system. Based on a detailed simulation of the integration of a steam cycle for combined heat and power (CHP) production to the SNG production process in former work [9], values for η_{ex} close to 0.7 can be expected for cases with a high level of thermal integration making use of internal heat pockets of the SNG process. Therefore η_{ex} = 0.7 is used for the evaluation of the electricity production from the recoverable excess heat for the different cases investigated here. For the generation of the composite curves, individual temperature differences necessary for the heat exchange have been assumed for the streams depending on their nature.

3.2 Thermodynamic performance indicators

The comparison of the three alternative gas upgrade technologies is based on a number of performance indicators defined for the overall process, as well as specifically adopted for the gas upgrade sections. To judge the performance from an overall process perspective the efficiency from biomass to SNG η_{cg} – also referred to as cold gas efficiency – is defined on a lower heating value basis according to:

$$\eta_{cg} = \frac{\dot{m}_{SNG} \cdot LHV_{SNG}}{\dot{m}_{wood\ fuel} \cdot LHV_{wood\ fuel}} \qquad (1)$$

with $\dot{m}_{SNG}$ and $\dot{m}_{wood\,fuel}$ being the mass flow of the produced SNG and the wet biomass wood fuel input, respectively. $LHV_{wood\ fuel}$ is the lower heating value of the wet fuel (at 50 wt-% moisture). Eq. (2) defines the overall thermal energy efficiency η_{sys} from a system perspective taking into account the exchange of electricity output between the SNG production process and the electricity grid in the energy system background:

$$\eta_{sys} = \frac{\dot{m}_{SNG} \cdot LHV_{SNG} + \dfrac{P_{el}^{-}}{\eta_{el,ref}}}{\dot{m}_{wood\ fuel} \cdot LHV_{wood\ fuel} + \dfrac{P_{el}^{+}}{\eta_{el,ref}}} \qquad (2)$$

P_{el} represents the net electricity production (-) or consumption (+) and $\eta_{el,ref}$ the reference electricity production efficiency for the corresponding energy market scenario. Eq. (2) thereby represents a comparison of the processes’ energy output compared to the input on a primary energy level.

In addition to these two overall efficiency definitions, performance indicators specific to the upgrade section defined in eqs. (3) to (5) are used for comparison of the different technologies.

$$\phi_{CH_4} = \frac{\dot{m}_{CH_4,SNG}}{\dot{m}_{CH_4,rawgas}} \qquad (3)$$

$$\eta_{upgrade} = \frac{\dot{m}_{SNG} LHV_{SNG}}{\dot{m}_{rawgas} LHV_{rawgas} + P_{el,upgrade}} \qquad (4)$$

$$p_{upgrade} = \frac{P_{el,upgrade}}{\dot{m}_{SNG} LHV_{SNG}} \qquad (5)$$

The methane recovery factor ϕ_{CH_4} compares the amount of methane in the final product $\dot{m}_{CH_4,SNG}$ to the methane content in the raw gas entering the gas upgrade section $\dot{m}_{CH_4,rawgas}$. The upgrade efficiency $\eta_{upgrade}$ is a comparison of the thermal power output in form of SNG to the input to the upgrade section in form of thermal ($\dot{m}_{rawgas} LHV_{rawgas}$) and electric power $P_{el,upgrade}$. The specific power consumption for upgrade $p_{upgrade}$ is the ratio between electric power consumption within the upgrade section $P_{el,upgrade}$ and the thermal power output in form of SNG ($\dot{m}_{SNG} LHV_{SNG}$). As the thermal output from the SNG process varies with the choice and operating conditions of the upgrading section, it might be argued that $p_{upgrade}$ should rather be based on the thermal power input to the upgrading section for consistency reasons as for example done by Gassner et al. [5]. But due to the existence of a recycle stream with differing composition and absolute flow for the amine-based and PSA upgrade technologies, the thermal energy of the raw gas stream into the gas upgrade section will differ between the three alternatives. It was therefore deemed more appropriate to define $p_{upgrade}$ based on the thermal output that is subject to the quality specifications according to section 3.1.

3.3 Economic evaluation using future energy market scenarios

For the economic evaluation of the different technologies the production costs for SNG are estimated based on the necessary investment, operating costs, and revenues. The price levels for different energy services and products have been set up in four scenarios representing cornerstones of possible price projections for the year 2030. Applying consistent sets of energy market scenarios, a packaged sensitivity analysis of the dependence of SNG production costs on energy and commodity price levels can be performed.

The equipment costs for the different process steps are calculated based on handbooks and scientific literature (see Appendix A). Using the factorial approach proposed by Smith [23] the total fixed capital costs (TFCI) and the total capital costs C_F – including the working capital necessary for start-up of the plant – are determined based on the cost of the delivered equipment items C_E.

$$\begin{aligned} C_F = & \sum_i \left[f_M f_P f_T (1 + f_{PIP}) \right]_i C_{E,i} + \\ & + \left(f_{ER} + f_{INST} + f_{ELEC} + f_{UTIL} + f_{OS} + f_{BUILD} + f_{SP} + f_{DEC} + f_{CONT} + f_{WC} \right) \sum_i C_{E,i} \end{aligned} \tag{6}$$

Values used for the factors in eq. (6) are provided in Table 2. Depending on whether the process step evaluated is handling solids (e.g. drying) or fluids (e.g. gas upgrade), different factors are used. Material factors f_M, as well as pressure and temperature factors f_P and f_T, respectively, have been determined based on the operating conditions using the ranges given by Smith [23]. As the capital cost estimation is based on delivered equipment costs C_E, costs given as purchased equipment costs are increased by 10% accounting for freight and transport costs. In a similar manner, cost data obtained from literature that partly incorporates installation costs was back-calculated to delivered equipment costs C_E using the factors given in Table 2 in order to consistently apply eq. (6) for all cost estimation. All costs are evaluated on a €$_{2010}$ basis using the Chemical Engineering Plant Cost Index (CEPCI) [24] for updating the obtained cost data from the different sources. Details on the cost estimation for the different types of equipment are given in Appendix A.

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Table 2: Factors used for capital cost estimation based on delivered equipment costs (taken from Smith [23]).

Item	Symbol	Type of process	
		Fluid processing	Solid processing
Direct costs [a]			
Piping	f_{PIP}	0.7	0.2
Equipment erection	f_{ER}	0.4	0.5
Instrumentation & controls	f_{INST}	0.2	0.1
Electrical	f_{ELEC}	0.1	0.1
Utilities	f_{UTIL}	0.5	0.5
Off-sites	f_{OS}	0.2	0.2
Buildings (incl. services)	f_{BUILD}	0.2	0.3
Site preparation	f_{SP}	0.1	0.1
Indirect costs [a]			
Design, engineering, and construction	f_{DEC}	1.0	0.8
Contingency	f_{CONT}	0.4	0.3
Working capital			
Working capital	f_{WC}	0.7	0.6

[a] Direct and indirect cost factors are adding up to the total fixed capital investment *TFCI* (e.g. for equipment at base conditions (f_M,f_P and f_T equal to 1): $TFCI = 4.8 \cdot C_E$ for fluid processing and $TFCI = 3.8 \cdot C_E$ for solid processing)

Using the annualized total capital investment it is possible to calculate the specific production costs for SNG c_{SNG} in €$_{2010}$/MWh$_{\mathrm{LHV}}$ for the different cases:

$$c_{SNG} = \frac{aC_F + c_{OM} + c_{lb}t_{lb} + \left(\dot{m}_{fuel}LHV_{fuel}c_{fuel} + \dot{m}_{O_2}c_{O_2} - P_{el}c_{el} - \dot{m}_{CO_2}\left(c_{CO_2} - c_{CO_2,ts}\right)\right)t_{op}}{\dot{m}_{SNG}LHV_{SNG}t_{op}} \qquad (7)$$

In Eq (7), a represents the annuity factor, c_{OM} the operating and maintenance costs per year, and the term $c_{lb}t_{lb}$ the yearly labour costs. $\dot{m}_i$ is the flow and c_i the cost for stream or service i, and t_{op} the yearly operating hours. The operating and maintenance costs are assumed as a fraction of the investment cost for all three gas upgrade technologies. Actually, there might be differences between the operation costs for each technology, but it is hard to put reliable numbers on them. And as the operating and maintenance cost contribution to the specific production costs is only about half of the cost contribution of e.g. fuel or annualized investment cost contribution, the error made by this simplification is considered acceptable.

The estimated production costs c_{SNG} can then be compared to the market price of natural gas c_{NG}, yielding an indication of the minimum necessary subsidy level $c_{subsidy}$ for rendering SNG production economically viable.

$$c_{support} = c_{SNG} - c_{NG} \qquad (8)$$

The economic cost parameters valid for all energy market scenarios investigated are given in Table 3. Table 4 represents the prices levels for energy services and products for the four scenario cases set up, as well as the necessary input to the scenarios.

Table 3: Economic cost parameters.

Economic parameter	**Symbol**	**Value**	**Unit**
annuity factor	a	0.1	1/y
operating hours	t_{op}	8000	h/y
Operating and maintenance cost	c_{OM}	$0.05 y^{-1} \cdot C_F$	€$_{2010}$/y
labour (man hours)[a]	t_{lb}	77000	h/y
labour costs [b]	c_{lb}	32	€$_{2010}$/h
CO_2 transport and storage cost[c]	$c_{CO_2,ts}$	22	€$_{2010}$/(10^3 kg CO_2)
Oxygen supply cost[d]	c_{O_2}	51.7	€$_{2010}$/(10^3 kg O_2)

[a] based on a plant capacity of about 100 000 kg$_{SNG}$/day in a highly automated plant and 7 processing steps [25].

[b] based on hourly labour costs for the Netherlands (assumed to be close to Central European average) in 2010 according to Eurostat [26].

[c] 13 €$_{2010}$/ton CO_2 for transport, 9 €$_{2010}$/ton CO_2 for storage [18].

[d] based on O_2-demand for membrane case with CO_2 storage (826 Nm^3/h), On-site plant, cryogenic, 7 \$$_{2000}$/100 Nm^3 O_2 [27].

For setting up the four energy market scenarios the ENPAC tool (version 1.8) developed at Chalmers University of Technology [28] has been used. The inputs to the tool are fossil fuel price and CO_2 emission charge levels, as well as the level of production subsidy for renewable electricity generation. Scenarios 1 to 3 are based on price projections for fossil fuels and CO_2 charge for 2030 taken from IEA's World Energy Outlook 2011 [29], and Scenario 4 uses the average of the extreme price level predictions from World Energy Outlook in order to serve as a centre point. The level of production subsidy for renewable electricity production was set to 20 €$_{2010}$/MWh$_{el}$ for all scenarios, representing an average value for the European Union [28] based on historic data. For consistency reasons it actually would be preferable to use a predicted value of the renewable electricity production subsidy level for the year 2030 instead, but no data is available for this kind of data to the authors' knowledge. Based on this input the ENPAC tool determines the build margin technology based on the technology that achieves minimum total costs for electricity generation, as well as other fuel price levels. For wood fuel the price is determined based on the willingness-to-pay of an alternative biomass user. In the current study – having a medium term perspective with year 2030 – coal condensing power plants using biomass for co-firing were set as reference biomass user.

Table 4: Future energy market scenario input (to the ENPAC tool [28]) and end used price output (used for economic analysis).

Parameter	**Unit**	**Scenario 1**	**Scenario 2**	**Scenario 3**	**Scenario 4**
Fossil fuel price level (input)					
Crude oil	$€_{2010}/MWh_{LHV}$	63.2	55.2	45.6	54.4
Natural gas	$€_{2010}/MWh_{LHV}$	36.1	33.5	27.8	31.9
Coal	$€_{2010}/MWh_{LHV}$	11.6	10.9	7.4	9.5
CO_2 *charge (input)*	$€_{2010}/t\ CO_2$	30.2	30.2	71.7	51
Renewable electricity production subsidy (input)	$€_{2010}/MWh_{el}$	20	20	20	20
End user prices and policy instruments					
Wood fuel (forest residue) c_{fuel}	$€_{2010}/MWh_{LHV}$	30.7	30.0	41.3	36.0
Electricity (incl. CO_2 charge) c_{el}	$€_{2010}/MWh_{el}$	68.2	66.9	86.3	78.9
Reference electricity production technology		Coal	Coal	NGCC	Coal
Natural gas (incl. CO_2 charge) c_{NG}	$€_{2010}/MWh_{LHV}$	48.1	45.5	49.1	48.6

4. Gas upgrade technologies for SNG production

The three different gas upgrade technologies – amine-based chemical absorption, membrane separation, and pressure swing adsorption – investigated in this study have been previously analysed isolatedly in a screening study using an average composition of the raw gas and neglecting effects of recycle streams [30]. Part of the work presented here is based on the modelling efforts in that work. All three technology alternatives have been adapted to fulfil the quality requirements for SNG and CO_2 (when relevant). Using the process model, each upgrade technology has been tuned for thermodynamically favourable process parameters according to eqs. (3) to (5) using sensitivity analysis. No mathematical optimisation routine has been applied but the adjustment of the operating conditions has been done within the technically feasible range for each specific technology based on engineering knowledge.

4.1. Amine-based chemical absorption

The main difference between amine-based absorption and the two other separation technologies investigated in the current study is the fact that the energy necessary for the separation process is mainly supplied as thermal energy. Amine-based absorption for CO_2 separation has mainly been investigated for the removal of CO_2 from flue gases from power plants with a considerable number of techno-economical investigations available (e.g. [31-35]). In the framework of SNG production via thermal gasification, amine-based absorption has received less attention. A recent modelling-based study [13, 36] analyzes different CO_2 separation technologies for gas mixtures originating from biological fermentation. The gas composition of these biogenic gases is similar to the raw gas after methanation during SNG production, mainly consisting of CH_4 and CO_2. Amine-based chemical absorption is shown to be superior to physical absorption techniques giving a low methane slip, high methane recovery and high CO_2 separation rate [36]. For the amine-based CO_2 separation model used in this study, basic data from Götz et al. [36] has been used together with estimations based on a detailed model for flue gas CO_2 separation [37]. From the data given in [36] a thermal energy demand for the regeneration of the amine-solution of 3.3 MJ/kg CO_2 separated can be calculated. A range of 3-5 MJ/kg CO_2 is proposed in a recent review paper [38] for the thermal demand for regeneration, but undisclosed vendor information even indicates lower heat demand of about 2.45 MJ/kg CO_2 [36]. The differences are mainly due to the choice of absorption medium (monoethanolamine (MEA), diethanolamine (DEA), or various patented amine-based aqueous solutions), the complexity of the installation, and the operating conditions.

In Fig. 2 the flowsheet for the CO_2 upgrade section with chemical absorption is illustrated. The basic steps are the CO_2 separation, an intermediate compression followed by the H_2 separation by membrane, and the final gas drying before compression to the pipeline pressure. The optional compression of CO_2 for storage is indicated in Fig. 2 as well. No further treatment of the CO_2 rich stream is to be expected. Details on the modelling assumptions can be found in Appendix B.

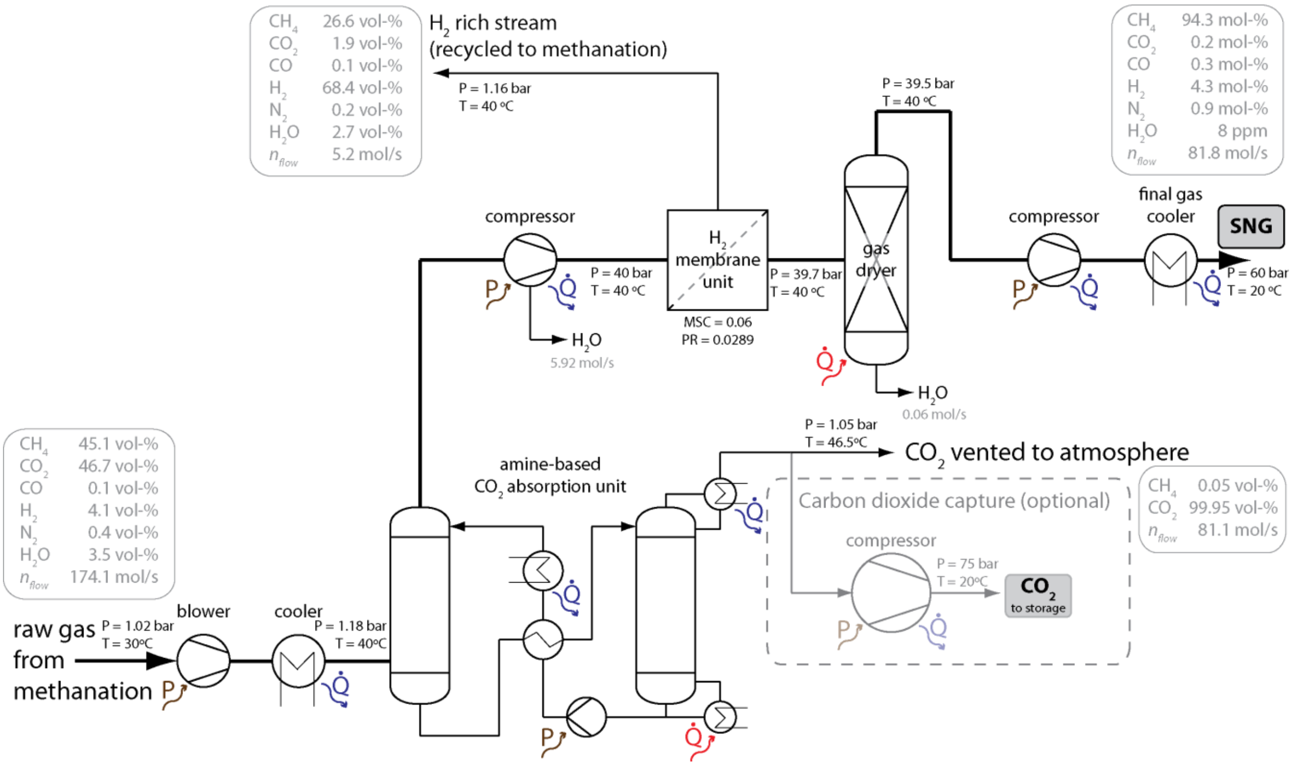

Figure 2: Flowsheet section for the gas upgrade with amine-based CO_2 absorption with indication of major process parameters and the major streams' composition and flow as obtained from the process model (grey boxes).

4.2. Membrane separation

The most prominent application area for membrane-based gas separation is hydrogen production from e.g. methane steam reforming or hydrogen recovery from off-gases [39]. Membrane separation is also used for removing CO_2 from crude natural gas [40, 41], resembling the gas upgrade process in the current study. A comparison of the economic performance of membrane separation and amine-based separation processes for natural gas upgrade [42] shows results in favour of membrane applications, even pointing out hybrid solutions involving both techniques as economically interesting for certain cases. The membrane material assumed in the studies on natural gas upgrade [40-42] is polymeric, namely cellulose acetate. As this is also a membrane material commonly used in commercial units for hydrogen separation [39, 43] it has been assumed for all membrane separation units in this study.

Bhide and Stern [40] showed that two or even three-stage setups are necessary to obtain satisfactory gas upgrade. Gassner et al. [5] also set up different configurations involving two to three membrane separation stages for the gas upgrade within SNG production. In the current study a two-stage setup is chosen as the increase in overall efficiency as well as reduction in production costs is rather small when going from two to three stages even though methane recovery increases substantially [5]. In Fig. 3 the gas upgrade flowsheet using membrane technology for CO_2 separation is given, illustrating the major process parameters and modelling assumptions. The membrane unit modelling is done based on a short-cut design model in analogy with counter-current heat exchangers [44]. The off-gases from the first membrane unit are burnt in a catalytic combustions unit as for the PSA process. For cases including CO_2 compression for storage, pure oxygen is used in order to avoid dilution of the CO_2 stream. As the water vapour present in the raw product gas will mainly end up in the CO_2 rich stream, no final drying stage is necessary for reaching the imposed gas specifications for SNG. More details on the model parameters are given in Appendix B.

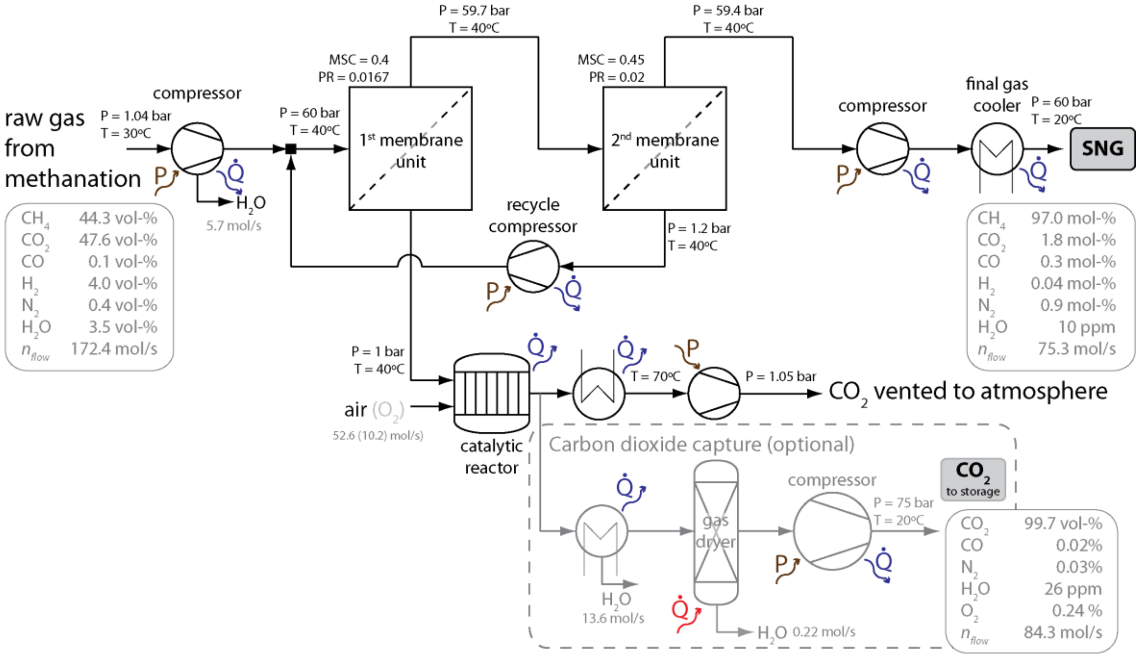

Figure 3: Process flowsheet for the gas upgrade process based on membrane separation units for CO_2 separation with indication of major process parameters and the major streams' composition and flow as obtained from the process model (grey boxes).

4.3. Pressure swing adsorption (PSA)

PSA is a commonly used industrial process for air (N_2/O_2) separation, gas drying or hydrogen purification [45]. The technology is also used for the upgrading of landfill gases or biogases from fermentation that resemble the composition of raw SNG being basically a mixture of CH_4 and CO_2 [46-52]. The active material for adsorption may be silica gel, zeolites, or carbon molecular sieves for the separation of CO_2 from CH_4. In a recent paper, Santos et al [47] estimate an upgrade efficiency for biogas using PSA reaching methane recovery of 85% with a purity higher than 99% for a mixture of 67% CH_4 and 33% CO_2 using zeolite 13X. Cavenati et al. [46] experimentally investigated the upgrade of a 55% CH_4/45% CO_2 mixture representative of landfill gas using a vacuum pressure swing adsorption (VPSA) technology with carbon molecular sieves. They estimate that purity of methane of 96% can be obtained with recovery above 75% for a VPSA process operating between 3.2 bar (adsorption pressure) and 0.1 bar (regeneration pressure). The superior performance of molecular sieves for landfill gas upgrade has also been demonstrated by Pilarczyk et al. [51, 52] with zeolites being sensitive to humidity and silica gel showing inferior adsorption characteristics compared to a carbon molecular sieve. During test runs with simulated biogas, a methane recovery of 97% was obtained at a very high gas purity of 99.5% [51]. Care must however be taken when scaling up PSA from pilot to industrial scale as performance losses are to be expected [53]. The methane recovery in this study was therefore set to 95% with a purity of methane of 97% assuming a binary mixture of CH_4 and CO_2. For the modelling details, see Appendix B.

The gas is usually compressed prior to the adsorption unit to a pressure between 3-9 bar while the low pressure of the PSA cycle is below atmospheric pressure, typically around 0.05 bar [51]. A recycle of the purge gas substantially increases the recovery and product gas purity [51]. The setup using a PSA unit for the gas upgrade section is shown in Fig. 4. A catalytic reactor with preheat is used for burning the off-gases from the PSA unit as the methane concentration in the off-gas/air mixture is below the lower flammability limit for methane of 5 vol-% [54]. The combustion of the off-gases allows for considerable heat recovery increasing the steam generation and in consequence the power generation. For the PSA gas upgrade the option of CO_2 compression for storage was not investigated as the CO_2-rich off-gas is released at sub-atmospheric pressure rendering the concept unfavourable for CO_2 storage.

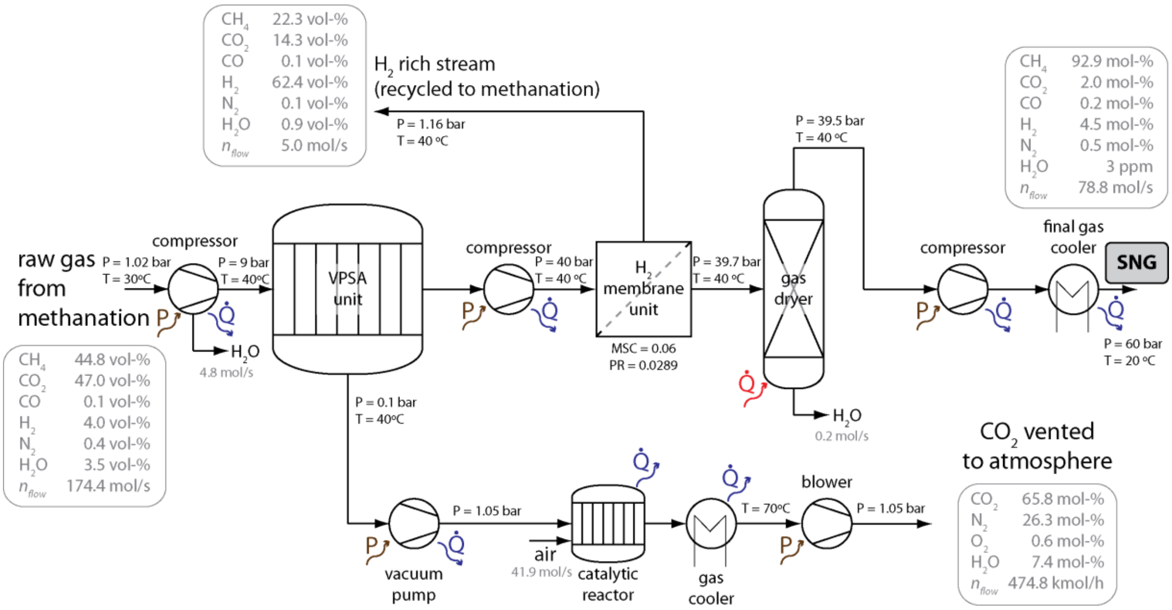

Figure 4: Process flowsheet for the gas upgrade process based on pressure swing adsorption for CO_2 separation with indication of major process parameters and the major streams’ composition and flow as obtained from the process model (grey boxes)..

5. Results & Discussion

5.1 Energy performance

Based on the energetic yield of SNG, amine-based absorption performs best among the three technology alternatives. Table 5 gives the overall energy figures for the three technologies, as well as for the alternatives with CO_2 separation (MEA and membrane case). The net power production is highest for the PSA case and lowest for the amine-based absorption. As the MEA case mainly uses excess heat for the separation process, this was to be expected. For the cases where CO_2 is separated, the exergy-based analysis for estimating power generation indicates a higher electric power production that is partly due to the increased cooling demand for compressor cooling of the CO_2 compressor train. It is questionable if this additional excess heat really will translate to increased power production. The increase has been considered anyway for consistency reasons and will not affect the results significantly as it also is partly compensated by higher investment cost for e.g. the steam cycle equipment.

Table 5: Overall energy balance for the different gas upgrade technology alternatives.

		MEA (with CO_2 storage)	**Membrane (with CO_2 storage)**	**PSA**
Fuel input	MW_{LHV}[a]	90.3	90.3	90.3
SNG production	MW_{LHV}	62.8	58.7	59.6
Gas upgrade section power consumption	MW	1.7 (3.1)	4.0 (5.4)	3.1
Overall power consumption[b]	MW	6.3 (7.7)	8.6 (10.1)	7.7
Power generation	MW	10.7 (11.0)	14.3 (14.8)	13.8
Net power export	MW	4.3 (3.3)	5.6 (4.7)	6.1

[a] 50 weight-% moisture

[b] including gasification section, gas upgrade section and auxiliaries (taken as 2% of thermal fuel input)

The energetic performance evaluation based on the previously defined indicators is given in Table 6. The cold gas efficiency η_{cg} for the process ranges between 0.65 and 0.695 depending on the gas upgrade technology chosen. The overall system efficiency η_{sys} ranges from 0.744 to 0.793 both depending on the gas upgrade technology and the background energy system. The MEA process performs best according to both performance indicators, except for the scenario where natural gas combined cycle technology is assumed to be the reference electricity production technology, where the PSA gas upgrade process outperforms the MEA case based on η_{sys}. The PSA process alternative has a higher cold gas efficiency η_{cg} than the membrane case even though the methane recovery ϕ_{CH_4} is lower. This is due to the fact that the SNG produced by membrane separation basically only consists of methane as combustible component while for the PSA separation a considerable amount of hydrogen is present in the final SNG, leading to a higher energy content. Amine-based absorption gives both the highest methane recovery ϕ_{CH_4} and upgrade efficiency $\eta_{upgrade}$, and thus the highest value for η_{cg}.

Table 6: Energy performance indicators for the three gas upgrade technology alternatives.

		MEA	**Membrane**	**PSA**
Wobbe index WI	kWh/m^3	14.7	14.5	14.3
relative density ρ_{rel}	-	0.566	0.605	0.581
methane recovery ϕ_{CH_4}	-	0.982	0.957	0.936
upgrade efficiency $\eta_{upgrade}$	-	0.944	0.876	0.882
specific power consumption for upgrade $p_{upgrade}$	kJ/kWh_{SNG}	97.6	245.5	185.6
cold gas efficiency η_{cg}	-	0.695	0.650	0.660
overall system efficiency η_{sys}	-	$0.790^a/0.768^b$	$0.772^a/0.744^b$	$0.793^a/0.762^b$

[a] Scenarios 1,2 & 4 with Coal condensing steam power cycle as build margin ($\eta_{el,ref} = 0.51$)

[b] Scenario 3 with Natural gas combined cycle (NGCC) as build margin ($\eta_{el,ref} = 0.66$)

The specific power consumption $p_{upgrade}$ is highest for the membrane case as the large raw gas flow is compressed to the highest pressure level among all cases. In consequence the membrane case also shows the lowest upgrade efficiency $\eta_{upgrade}$. Both the membrane and PSA cases have high specific power consumption but still, the net electricity production is higher for the two cases compared to the MEA case. This is partly due to the considerable amount of process excess heat that is consumed by the reboiler in the MEA case, not being fully available for steam generation and power production in consequence. In addition the off-gases available for both the membrane and PSA cases contribute to additional power generation compared to the MEA case. The latter aspect also makes the PSA and membrane cases more flexible in adjusting between power generation and SNG production depending on the economic background conditions, while the choice of amine-based absorption leaves less margins for variation between power generation and SNG production.

5.2 Economic evaluation

The results of the investment cost estimation given in Table 7 indicate that the PSA technology is the most capital intensive alternative, followed by the membrane case and the MEA case, resulting in the lowest total capital cost C_F of 205.5 $M€_{2010}$. The increase in C_F for integrating CO_2 compression and preparation for storage amounts to about 5 and 6 $M€_{2010}$ for both the MEA and membrane cases, respectively.

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Table 7: Fixed costs for the different gas upgrade technologies with and without CO_2 capture.

Investment and operation cost		**MEA**	**MEA with CO_2 capture**	**Membrane**	**Membrane with CO_2 capture**	**PSA**
Total fixed capital investment (TFCI)	M€$_{2010}$	182.9	187.4	179.8	184.7	199.1
Total capital cost C_F	M€$_{2010}$	205.5	210.5	200.5	206.6	223.0
Annualized investment $a\ C_F$	M€$_{2010}$/y	20.6	21.1	20.1	20.7	22.3
O&M cost C_{OM}	M€$_{2010}$/y	10.3	10.5	10.0	10.3	11.2
Labour cost C_{lb}	M€$_{2010}$/y	2.5	2.5	2.5	2.5	2.5
Sum $a\ C_F + C_{OM} + C_{lb}$	M€$_{2010}$/y	33.3	34.0	32.5	33.4	36.0

The distribution of the total fixed capital investment (*TFCI*) illustrated in Fig. 5 clearly shows the dominating influence of the gasification step on the overall costs. The cost contribution of the gas upgrade section varies from 13-22% of the TFCI and is lowest for the membrane case. With increasing electric power generation an increase of the fractional cost of the steam cycle can be observed as well.

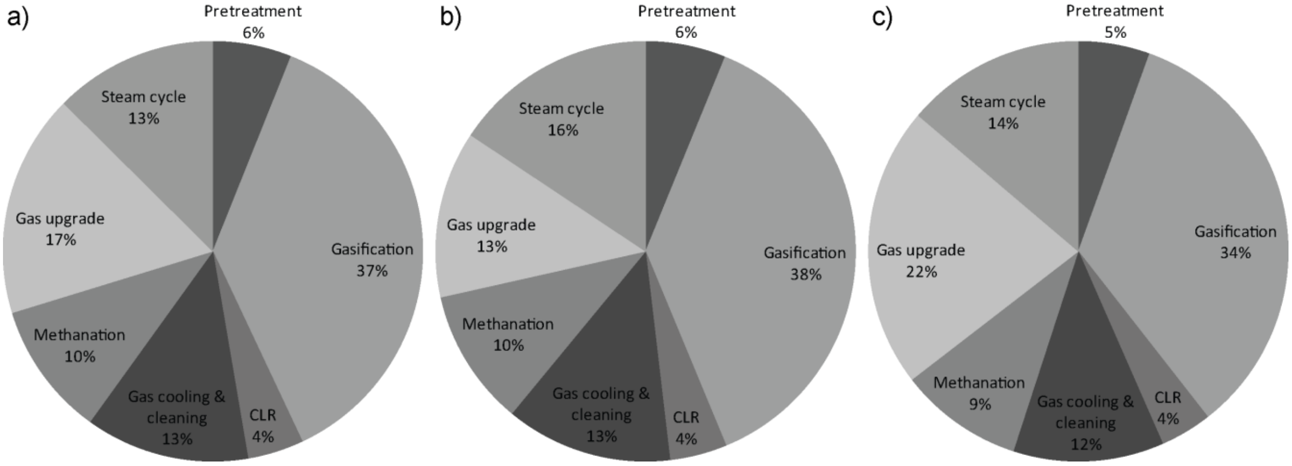

Figure 5: Distribution of total fixed capital investment cost (TFCI) between different plant parts for the three gas upgrade technologies (all without CO_2 storage). a) MEA, b) membrane, c) PSA.

The specific production costs for SNG c_{SNG} vary between 103-127 €$_{2010}$/MWh$_{SNG}$ as illustrated in Table 8. The PSA case results in the highest production costs, even for Scenario 3 where this case shows the highest system efficiency η_{sys}. The MEA gas upgrade technology results in the lowest cost for SNG production in all scenarios. The corresponding subsidy level $c_{subsidy}$ varies between 56 and 78 €$_{2010}$/MWh$_{SNG}$ depending on both the scenario and the gas upgrade technology. These numbers can be compared to a recently introduced production subsidy for biogenic gas production of about 52.4 €/MWh in Denmark [55]. Regarding CO_2 storage it can be stated that a high level of CO_2 charge is necessary to cover the costs of implementing this process option. This is mainly due to the high cost of 22 €$_{2010}$/t CO_2 for transport and storage assumed in this study. In Scenarios 1 and 2 with a CO_2 charge of 30.2 €$_{2010}$/t CO_2, the production costs for SNG actually increase when considering CO_2 storage. In

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Scenarios 3 and 4 c_{SNG} is reduced by 1.8-6.0% when assuming biogenic CO_2 storage as an option.

Table 8: SNG production cost and necessary subsidy level for the four energy market scenarios.

	Scenario 1	**Scenario 2**	**Scenario 3**	**Scenario 4**
SNG production cost c_{SNG}	$[€_{2010}/MWh_{SNG}]$			
MEA	104.4	103.4	118.4	111.2
MEA with CO_2 capture	105.5	104.5	111.3	108.3
Membrane	108.1	107.1	122.7	115.2
Membrane with CO_2 capture	110.5	109.6	116.0	113.1
PSA	112.9	111.9	127.1	119.8
Necessary subsidy level $c_{subsidy}$	$[€_{2010}/MWh_{SNG}]$			
MEA	56.3	57.9	69.3	62.6
MEA with CO_2 capture	57.4	59.0	62.2	59.7
Membrane	60.0	61.6	73.6	66.6
Membrane with CO_2 capture	62.4	64.1	66.9	64.5
PSA	64.8	66.4	78.0	71.2

Figure 6 illustrates the contribution of the different cost items to the production cost of SNG for two scenarios as well as the necessary level of production subsidy. The largest contribution to the production cost is the biomass feedstock cost, followed by the capital investment and the running costs (O&M and labour). Depending on the scenario, electricity generation and CO_2 storage contribute to some extent to a reduction in production costs.

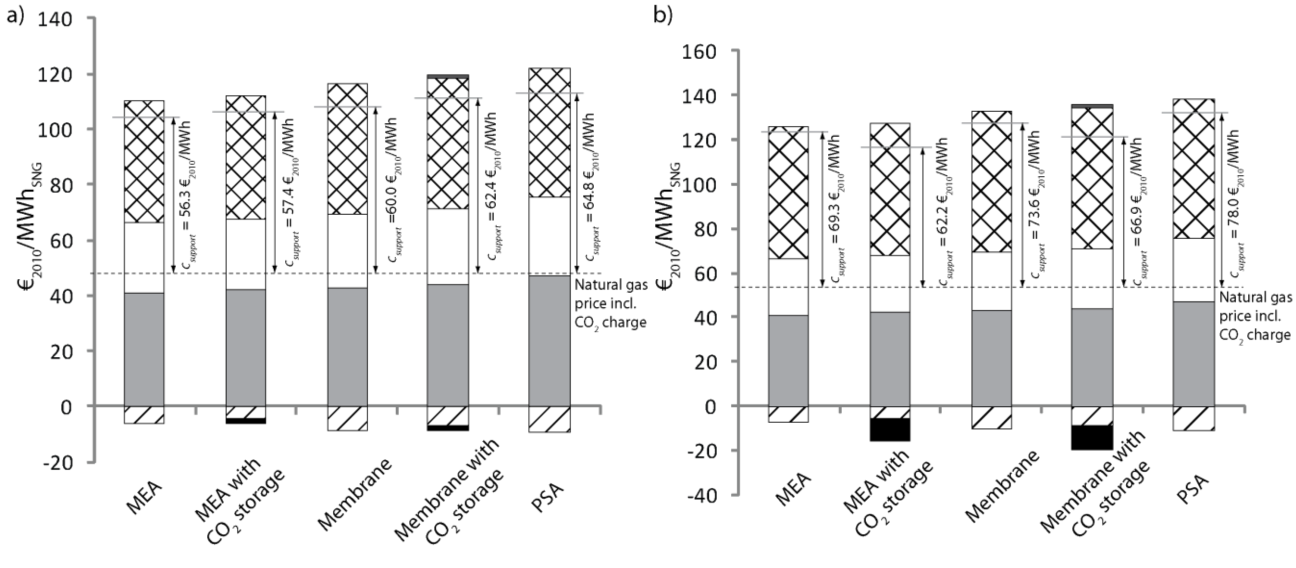

Figure 6: Contribution of the different cost items to the production costs for SNG for Scenario 1 (a) and Scenario 3 (b). (■ O_2 ▩ Biomass ☐ O&M and labour ▒ Investment ▨ Electricity ■ CO_2).

With the methodology applied for capital cost estimation an uncertainty of about ±30% is to be expected. In addition, the conversion of C_F to annual costs using the annuity factor a is subject to uncertainty. The value of 0.1 chosen for a might for example correspond to an estimated economic lifetime of 20 years and a fractional interest rate per year of 8%. The figures chosen for these two values – and the annuity factor a in consequence – depend on the nature of the investment. A strategic investment can have a rather long economic lifetime while a short term investment will have a shorter economic lifetime resulting in a higher annualized capital cost. In order to evaluate the influence of these considerations on the production cost for SNG a sensitivity analysis was performed. Mathematically, a variation of the total capital cost and the annuity factor both result in a change in annualized capital cost. It was therefore chosen to vary the total capital cost C_F (keeping the cost for operation and maintenance c_{OM} at its base values) being representative for changes in both above-mentioned factors. Figure 7 illustrates the changes in c_{SNG} for two scenarios when C_F is varied. It can be observed that the trend is similar for all cases investigated with a relative change of c_{SNG} of about ±12% for the extreme changes in C_F. The PSA case – having the highest value in C_F – is the most sensitive to changes in C_F but the differences between the three technologies are marginal. The cost figures used have been chosen rather conservatively (see Appendix A) so the base case rather overestimates the fixed capital cost. Considering the lower range for the investment cost in the sensitivity analysis, the necessary subsidy level for SNG production is lowest for scenario 1 and the MEA case with c_{SNG} = 92.0 €$_{2010}$/MWh$_{SNG}$ and $c_{subsidy}$ = 43.9 €$_{2010}$/MWh$_{SNG}$. This combination of technology and scenario even requires the lowest subsidy level when considering CO_2 storage (c_{SNG} = 92.8 €$_{2010}$/MWh$_{SNG}$ and $c_{subsidy}$ = 44.7 €$_{2010}$/MWh$_{SNG}$).

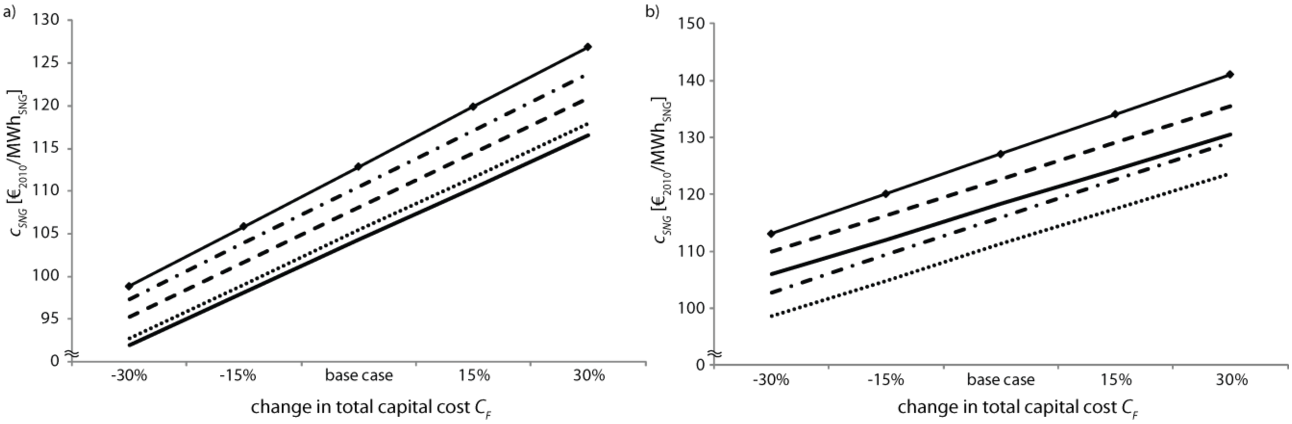

Figure 7: Sensitivity of SNG production cost c_{SNG} to total capital capital cost C_F for Scenario 1 (a) and Scenario 3 (b). Solid line – MEA case, dotted line – MEA case with CO_2 storage, dashed line. membrane case, dash-dotted line – Membrane case with CO_2 storage, solid line with diamonds – PSA case.

7. Conclusions

Three different gas upgrade technologies for production of SNG have been investigated for their thermodynamic and economic performance. Amine-based absorption results in the highest cold gas efficiency η_{cg} of 0.695 and even performs well considering the overall system efficiency. For a 100MW$_{th,LHV}$ biomass input plant, membrane and PSA technology for gas upgrade give a decrease of SNG yield of 4.1 and 3.2 MW, respectively. This can be weighed against an increased electricity production of 1.3 MW for membrane-based and 1.8 for PSA-based gas upgrade. The overall system efficiency varies between scenarios and cases with amine-based absorption and PSA technology performing about 2 %-points better than membrane technology. Amine-based absorption is the least flexible of all three technologies offering little potential for variation between SNG production and electricity generation while the two other technologies have a higher degree of freedom with the off-gases being burnt for steam generation. Changes in operating parameters for the membrane or PSA system allow adaption of the SNG gas yield and quality and the amount of off-gases available for steam generation. This enables a better adaptation to varying commodity prices. All three technologies offer potential for efficiency improvement, but using sensitivity analysis on process parameters, the same level of sophistication was retained for all three technologies in order to enable a fair comparison. Regarding the process integration aspects, it might also have been considered to use the off-gases from the separation process in the combustion unit of the indirect gasifier. This would have led to higher cold gas efficiencies for both the membrane and PSA process alternatives. Due to uncertainties on the flammability limit and the associated increased complexity of both the combustion system and the overall process control, this option was not considered.

The determined production costs for SNG indicate a rather high level of production subsidy in the range of 56-78 $€_{2010}/MWh_{SNG}$ that is needed to render the process economically viable. A sensitivity analysis on the influence of changes in the total capital costs shows that decreased investment costs by 30% can bring the SNG production cost down to levels needing subsidy levels of about 44 $€_{2010}/MWh_{SNG}$. CO_2 storage from biomass – when included in the emission trading system – only becomes an option at CO_2 charges higher than 50 €/ton CO_2. This is due to increased investment costs, but in particular due to rather high costs for CO_2 transport and storage that have been assumed in this study.

Acknowledgments

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Appendix A – Equipment cost estimation data

Table A-1: Purchased equipment cost at base rating C_E in M€$_{2010}$ and cost correlation data for major process units.

Unit	C_E for scale in present study	scale in present study	Original unit cost[a]	Base scale	Scale factor[b]	Ref.
Pre-treatment						
Biomass receive and unload[c]	0.98	38.04 wet t/h	3.5 MUS \$$_{2007}$	198.06 t coal/h	0.62	[56]
Biomass storage	0.98	38.04 wet t/h	0.65 M€$_{2002}$	33.5 wet t/h	0.65	[57]
Biomass dryer[d]	0.72	$1.08 \cdot 10^6$ m^3/h	$0.2\frac{(m^3/h)^{0.8}}{1000}[MSEK_{2003}]$			[58]
Gasification						
Indirect gasifier[e]	10.7	100 MW$_{th,LHV}$ (20% MC)	8 M€$_{2003}$	8 MW$_{th,LHV}$	0.72	[59, 60]
Gas cleaning						
Tar reforming[f]	0.72	8.1 Nm^3/s	3.23 MUS \$$_{2010}$	12.Nm^3/s	0.6	[61]
Fabric filter	0.073	15.6 m^3/s	68.8 kUS \$$_{2002}$	15.6 m^3/s	-	[25]
Water scrubber	3.4	9.0 m^3/s	3.0 M€$_{2002}$	12.1 m^3/s	0.7	[57]
Methanation island						
Guard bed	0.028	6.6 Nm^3/s	0.024 M€$_{2002}$	8 Nm^3/s	1	[57]
Methanation reactor[g]	2.6	13.3 m^3/s 7.1 Nm^3/s	reactor vessels, bayonet heat exchangers, and catalyst material			[10, 62-65]
Gas upgrade						
Amine-absorption unit[h]	4.72	3.9 Nm^3/s raw gas, 46.7 vol-% CO$_2$	Absorber, stripper, condenser, reboiler, heat exchangers, and pumps			[66, 67]
Membrane[i]	varying scale depending on case		500 \$$_{2006}$	cost per m^2 mounted in skid		[68]
PSA[j]	5.08	610.7 kmol/h	32.6 M€$_{2002}$	9600 kmol/h	0.7	[57]
Gas drying (Glycol)[k]	varying scale depending on case		20000 US \$$_{2000}$	1180 m^3/h	0.7	[69, 70]
Catalytic burner	varying scale depending on case		US \$$_{1988}$ – fixed bed reactor, sized based on gas flow rate and level of energy recovery			[71]
Steam cycle						
Turbine incl. steam system[l]	varying scale depending on case		5.9 M€$_{2002}$	10.3 MW$_e$	0.7	[57]
Standard equipment						
Fan/Blower[m]	varying scale depending on case and unit					[64]
Compressor[n]						[64]
Flash tank[o]			vertical vessel with demister			[64]
Heat exchanger[p]			heat exchangers, heaters and coolers			[64]

[a] cost data has been updated to 2010 using the composite Chemical Engineering Plant Cost Index (CEPCI) and converted to € on a 2010 exchange rate basis (1 €$_{2010}$ = 0.755 US \$$_{2010}$ = 9.54 SEK$_{2010}$). In case the year basis for the currency is not indicated, the year of publication of the reference minus one year for reports and minus two years for journal articles is assumed.

[b] Cost scaling is done using the standard correlation $\mathit{Cost\ B} = \mathit{Cost\ A} \cdot (\mathit{Capacity\ B} / \mathit{Capacity\ A})^{\mathit{scaling\ exponent}}$; in case no scaling exponent is given or can be derived a value of 0.6 is used.

[c] Large downscaling from coal treatment data introducing some uncertainty, but compared to scaling from other sources [72] equipment cost derived are rather high and the estimate is considered conservative.

[d] Gas flowrate of warm incoming air used in correlation. The cost correlation given in the reference includes conveying from and to dryer as well as engineering and foundation costs. The cost is back-calculated to purchased equipment cost C_E at base rating using a factor of 3 (based on Table 2 and eq. (6)).

[e] Estimated costs for a complete gasifier installation. The scaling coefficient is based on boiler costs as presented in [60]. The costs is back-calculated to purchased equipment cost C_E at base rating using a factor of 3.8 and a temperature factor f_T of 3.1 (based on Table 2 and eq. (6)).

[f] Based on capital cost figure for complete gasifier and tar reformer equipment where the cost of the reformer are stated to be 1/3 of the total equipment cost. Gas flow determined from data given in reference [61].

[g] vertical reactor vessel diameter based on mean superficial velocity of 0.14 m/s [63], max. 4 m in diameter. Vessel height based on correlation regressed by Gassner [10] for fluidised bed methanation reactors. Bayonet heat exchanger [64] for each reactor using the corresponding heat duty, overall heat transfer coefficient of 130 $W/(m^2 \cdot K)$, driving temperature difference of 50 K, and a area safety factor of 1.1. Catalyst material cost based on a space velocity of 4.5 Nm^3/(kg catalyst h) [62] and specific catalyst cost of 110 $€_{2010}$/kg (90-120 $\$_{2004}$/kg [65]).

[h] The correlations by Chapel [66] are for flue gas treatment and therefore used in this study outside the range of validity for the CO_2 concentration. The equipment sizing for all parts except the absorber are still considered valid as they are independent of the gas flow but only dependant on the amount of CO_2 separated. The absorber is considered to be oversized using Chapels correlation as the higher concentration of CO_2 in the product gas actually leads to a lower absorber diameter when designing for a similar gas flow rate. The cost estimate is therefore considered conservative.

[i] A large range of membrane costs can be found in literature. The chosen data is at the higher range and cost estimates can be considered conservative. The membrane area determined by the calculation model is increased by a safety factor of 10% for the cost calculation.

[j] The equipment cost has been back-calculated to base condition cost C_E for use with eq. (6) using a pressure factor f_P of 1.3 accounting for vacuum operation.

[k] The flow rate at base scale given in [69] is assumed to be at standard conditions common in natural gas processing (15ºC and 1 atm).

[l] Steam system consists of water and steam system, steam turbine, condenser and cooling. Scaled on steam turbine size [57]. The equipment cost determined by the correlation given are back-calculated to base condition cost C_E for use with eq (6) using a pressure and temperature factor of $f_P = 1.9$ and $f_T = 2.1$, respectively.

[m] Centrifugal radial fan, low grade stainless steel, up to $\Delta P = 0.15$ kPa.

[n] Centrifugal compressor, low grade stainless steel, maximum pressure ratio per stage of 4.

[o] Based on general heuristics [25] and a safety factor of 1.5.

[p] Shell-and-tube fixed tube heat exchanger, carbon steel, area safety factor of 1.1. Individual heat transfer coefficient based on the corresponding streams. For coolers and heaters exchanging heat with the steam cycle, a general heat transfer coefficient for the steam cycle side of $h = 8000$ $W/(m^2\ K)$ and a driving temperature difference of 20 K is assumed.

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Appendix B – Modelling assumptions

Table B-1: Modelling assumptions for the different process units.

Process unit	**Modelling assumptions**
Raw gas process (Fig. 1)	refer to [9]
Amine-based absorption	CO_2 separation efficiency: 0.9965 CH_4 loss: 0.0005 all other gases are assumed to remain in CH_4 rich stream Gas inlet and outlet T = 40 ºC Absorber ΔP = -0.1 bar reboiler heat demand (at 150ºC): 3.3 MJ/kg CO_2 separated recoverable heat (90 to 40ºC): 20% of reboiler heat load cooling demand (55 to 40 ºC): 80% of reboiler heat load amine solution pumps power consumption: 25kJ/kg CO_2 separated[a]
Membrane unit *(same model used for CO_2 separation and H_2 separation)*	separation factors and membrane area determined based on model by Pettersen and Lien [44] as a function of molar stage cut (MSC), pressure ratio (PR), membrane thickness (assumed to be 1000 Å), and permeability for the different gas species (in [barrer]: H_2O: 148, H_2: 17.8, CO_2: 8.9, CO: 0.45, CH_4: 0.3, N_2: 0.27)[b]
PSA unit	CH_4 recovery: 0.95 (same recovery assumed for H_2) recovery for remaining species except CO_2: 0.5 CH_4 / CO_2 molar ratio in CH_4 rich stream: 0.97 / 0.03 (assuming a 97% purity for a binary mixture) $P_{absorption}$ = 9 bar; P_{purge} = 0.1 bar Gas inlet and outlet T = 40 ºC Adsorber ΔP = -0.2 bar
TEG gas dryer	Adsorber ΔP = -0.2 bar Gas inlet and outlet T = 40 ºC H_2O separation efficiency: 0.99 reboiler heat demand (at 200ºC): 8.7 MJ/ kg H_2O separated[c]
Catalytic reactor (incinerator)[d]	relative air(oxygen)-to-fuel ratio: 1.05 incinerator ΔP = -0.1 bar gases preheated to 250ºC prior to incinerator by heat exchange with incinerator exhaust gas flow incinerator exhaust temperature limited to 600ºC by dilution with exhaust gas cooled to 250ºC
General	Heat exchanger pressure drop $\Delta P = -\,0.02 \cdot P_{in}$ Maximum compressor stage compression ratio: 4 Compressor isentropic stage efficiency: 0.77 Pump efficiency: based on pump efficiency curve for water[e] Compressor and pump mechanical efficiency: 0.98

[a] Based on [37] a power consumption for all pumps of 17.8 kJ/kg CO_2 can be determined, 25 kJ/kg CO_2 are used as a conservative estimate.

[b] The permeabilities and membrane thickness represent a cellulose acetate membrane with permeability data taken from [43, 73].

[c] specific reboiler heat load [MJ/kg H_2O separated] : 2.1 + 265 · *specific TEG circulation rate [m^3 TEG/kg H_2O separated]* (taken at an average specific TEG circulation rate of 0.025 m^3 TEG/kg H_2O (3 gal TEG/lb [d] basic design consideration taken from [76].

[e] standard in Aspen Plus.

Nomenclature

Letter symbols/Abbreviations:

a	annuity factor
c/C	specific/absolute cost
CEN	European Committee for Standardisation
CEPCI	Chemical Engineering Plant Cost Index
CHP	combined heat and power
DEA	diethanolamine
EASEE	European Association for the Streamlining of Energy Exchange
ENPAC	Energy Price and Carbon Balance Scenarios
f	cost factor
GCC	Grand Composite Curve
HHV	higher heating value
LHV	lower heating value
$\dot{m}$	mass flow
MC	moisture content
MSC	molar stage cut
NG	natural gas
NGCC	natural gas combined cycle
O&M	operation and maintenance
p	specific power consumption
P	power
PR	pressure ratio
(V)PSA	(vacuum) pressure swing adsorption
SNG	synthetic natural gas
t	time
TEG	tetraethylene glycol
TFCI	total fixed capital investment
w	specific work
WI	Wobbe index

Greek symbols

η	efficiency
ϕ	recovery
ρ	density

Subscripts and superscripts

BUILD	buildings (incl. services)
cg	cold gas
CONT	contingency
DEC	design, engineering, and construction
E	equipment
el/ELEC	electrical
ER	erection
F	total capital
INST	instrumentation and controls
lb	labour
M	material
OM	operation and maintenance
op	operation
OS	off-sites
P	pressure
PIP	piping
ref	reference
rel	relative
SP	site preparation
sys	system
T	temperature
ts	transport and storage
UTIL	utilities
WC	working capital

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