

# Electrical upgrading of biogas

EUDP 12 II



**HALDOR TOPSØE**   
CATALYSING YOUR BUSINESS

  
AARHUS UNIVERSITET

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**DGC** Dansk Gasteknisk Center a/s  
- rådgivning, udvikling og måling inden for energi og miljø



**HMN**   
NATURGAS

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# 1. Summary

## 1.1 Project details

Project title	El upgraded Biogas
Project identification (program abbrev. and file)	
Name of the programme which has funded the project	EUDP 12 II
Project managing company/institution (name and address)	Haldor Topsøe A/S Haldor Topsøe Alle 1 Dk-2800 Kgs. Lyngby
Project partners	Haldor Topsøe A/S Aarhus Universitet HMN NatureEnergy EnergiMidt DGC Ea energianalyse Planenergi
CVR (central business register)	41853816
Date for submission	28.02.2018

## 1.2 Short description of project objective and results

Projektet skulle i en skala på 10 Nm<sup>3</sup>/h demonstrere en højeffektiv proces til opgradering af rå biogas til rørledningskvalitet ved katalytisk omsætning af CO<sub>2</sub> i biogas til metan ved reaktion med brint produceret med dampelektrolyse i en SOEC elektrolyse enhed.

Et fuldskala anlæg skulle designes og de økonomiske aspekter evalueres.

Disse mål er nået på trods af mange problemer med især hjælpeudstyr. Produktion af gas med mere end 97 % CH<sub>4</sub> er blevet demonstreret. Metanatoren har fungeret som forventet og med uændret ydelse gennem projektet, hvilket bekræfter at afsvovlingsenhederne også har fungeret som planlagt.

Den høje effektivitet for SOEC enheden er blevet verificeret med mere end 2200 akkumulerede driftstimer.

De økonomiske analyser har vist at el opgraderet gas kan blive socio-økonomisk attraktiv i fremtidige energisystemer domineret af vedvarende energikilder og drevet af en høj CO<sub>2</sub> pris (800 kr./ton). Privatøkonomisk er processen attraktiv, hvis den producerede ekstra metan kunne opnå samme støtte som traditionel opgraderet biogas.

The project should on a scale of 10 Nm<sup>3</sup>/h demonstrate highly efficient upgrading of biogas to pipeline quality by catalytic conversion of the CO<sub>2</sub> content in the biogas to methane by reaction with hydrogen generated from steam in a Solid Oxide Electrolyzer. A full scale plant should be designed and the economic aspects evaluated.

These project objectives have been met despite numerous problems related to supporting equipment. Production of pipeline quality gas with more than 97 % CH<sub>4</sub> has been demonstrated and the performance of the methanator catalyst unchanged throughout the project verifying that the desulfurization catalysts also have performed as planned.

More than 2200 accumulated operating hours have been demonstrated and high efficiency for the SOEC stacks.

System analyses indicate that el upgraded biogas can become socio economically attractive in future energy system dominated by renewable energy sources and driven by high CO<sub>2</sub> prices (800 Dkr./tonne). If the produced methane, however, would benefit from the same incentive as biogas upgraded by classical means the process would be attractive from a private investment perspective.

### 1.3 Executive summary

A 10 Nm<sup>3</sup>/h biogas upgrading demonstration unit has been designed, constructed and operated at Aarhus university's biogas plant in Foulum.

A 50 kW SOEC unit using 8 stacks in two stack assembly units have generated approx. 16 Nm<sup>3</sup>/h of hydrogen, which was used to convert the approx. 43 % CO<sub>2</sub> in the desulfurized biogas catalytically to methane in a boiling water reactor.

Despite numerous problems with supporting equipment, notably the hydrogen compressor, more than 2000 hours of accumulated operating have been achieved with the SOEC stacks. After some initial fast degradation for the first 500 hours the performance of the stacks has been rather stable. The stack degradation has successfully been mitigated by an operating strategy where the inlet temperature gradually has been increased. In this way the production rate can be kept constant with unchanged high efficiency. The very low stack power consumption of around 3.1 kWh/Nm<sup>3</sup> hydrogen produced has been confirmed.

The methanator performance has been excellent and pipeline quality gas produced. There has been no change in performance as indicated by the temperature profiles. This implies that no degradation by thermal ageing or poisoning of the methanator has occurred confirmings perfect operation of the desulfurization system.

It takes 10-11 hours to heat up the SOEC from cold and 6-8 hours for the methanator. Operating load from hot stand-by can be changed within seconds for the SOEC units and a few minutes for the methanator.

The unit has been remotely controlled most of the time.

A full scale commercial plant has been designed and cost estimated. The following efficiencies has been calculated:

Configuration	A
Percent efficiency	CO <sub>2</sub> Meth.
Exergy eff gas	80.2
District heating	0.8
Oxygen	2.1
Exergy eff Total	83.1
LHV eff gas	76.3
LHV DH	14.0
LHV eff total	90.3

Based on these results the economy has been evaluated both from a socio-economic as well as from a private investment perspective.

The overall conclusions from the socio economic study are:

- In an energy system dominated by renewable electricity production and high CO<sub>2</sub> price (800 kr./ton as predicted by IEA) el upgraded biogas via SOEC can outcompete natural gas, simple upgraded biogas or SNG from thermal gasification
- The socio economic gain in Denmark would be 350 mio kr. per year or 40 kr./GJ per GJ extra methane produced
- The amount of renewable methane from biogas production is increased by approximately 30 % by using el upgrading
- The SOEC unit should operate at full load for 4600 – 4800 hours per year where the average weighted electricity price is 109 kr./MWh
- There exists a strong synergy with biogas engines for CHP production which will operate in the hours with high electricity prices
- The above scenario will not be realized in the near future but could be towards 2050

The results of this private investment economic study strongly indicate that methanation of biogas using hydrogen from SOEC can be economical if some conditions are fulfilled:

- The electricity price to the SOEC is no higher than 250-350 DKK/MWh on an annual average basis (the exact value depends on the hourly price distribution and on the choice of a back-up option for biogas consumption).
- The produced SNG can be sold at a price of 6.00 DKK/Nm<sup>3</sup> or higher.
- The steam output from the methanation is utilized as an input for the SOEC process.

## 2. Project objectives

### 2.1 Objectives

The project objectives were:

- To design and construct a demonstration unit for upgrading 10 Nm<sup>3</sup>/h of raw biogas to pipeline quality BNG (Bio natural gas) by means of SOEC electrolysis and catalytic methanation of the CO<sub>2</sub> in the desulphurized biogas with hydrogen from the SOEC unit

The purpose of the unit is to:

- Demonstrate that the SNG product can meet the Danish natural gas pipeline grid specification
- Demonstrate that the SOEC unit and the bio-SNG unit can operate independently due to the hydrogen storage.
- Demonstrate the bio-SNG and SOEC technologies can operate at continued and stable conditions.
- Evaluate the performance of the bulk sulfur absorbers.
- Evaluate the performance of the final sulfur guards.
- Evaluate the performance of the methanation reactor

Other objectives have been:

- Based on the experimental results evaluate the economics of the process both from a socio-economic and a private investors perspective

- Update the business plans for deploying the process

## 2.2 Main timelines of the project

Establishment of a consortium and supply agreements took much longer than anticipated (>9 months). Haldor Topsøe A/S then performed the detailed design of the demonstration unit using their standard, proprietary design tool and documentation. This formed the basis for issuing bid material for the construction of the BioSNG unit. After negotiations, Zeton in the Netherlands was chosen to construct the unit. In parallel a prereforming SOFC test unit was revamped by FuelCon in Germany to serve as test platform for the SOEC core. A Danish company, Vengcon, supplied the power supplies and control system for the SOEC Cores. Haldor Topsøe A/S separately purchased the biogas and hydrogen compressors from SERA in the UK.

Factory acceptance tests were performed on the BioSNG unit and SOEC unit in Holland and at Haldor Topsøe premises in Kgs. Lyngby respectively.

The BioSNG unit and the compressors were then shipped to Aarhus University's site in Foulum and assembled in a building adjacent to their biogas production plant in January-March, 2016.

A complete start up and operation was performed with SOEC unit at Topsøe, before the entire unit was shipped to Foulum and tied in to the rest of the unit(s) in April 2016. The start-up procedures were refined and the first production of upgraded biogas occurred beginning of June, 2016. A timeline of the major events throughout 2016 and 2017 are provided as Figure 2-1 and Figure 2-2.

The complete demonstration unit is the first of its kind in the world and also the first where two cores with 4 SOEC stacks each has been operated in tandem so it was to be expected that some integration issues had to be resolved, but numerous unexpected problems with the balance of plant equipment had to be addressed as illustrated on Figure 2-1 and Figure 2-2. Some of the less trivial and most time consuming has been:

1. Rupture on the membranes in the hydrogen compressor. The expected lifetime was quoted to be at least 4000 hours and experience with a similar compressor at Topsøe confirm this to be achievable. Nevertheless, operating time less than a 1000 hours has been experienced causing prolonged downtime due to excessive delivery times for spare parts. The vendor has sent technical representative which went over the operating procedures but could not identify any root cause of the problems.
2. The inter stage water separators on the compressors have also caused an undue number of trips.
3. Totally unexpected has been the occurrence of several total power failures on the grid.
4. The control of the steam generator in the SOEC unit was far from optimum resulting in a number trips when too low temperatures exit the generator resulted in system shut down to prevent liquid water ingress to the stacks

Unfortunately, some of the BOP problems have caused catastrophic stacks failures:

- Due to a malfunctioning trip system in the SOEC unit hot stacks was mechanically weakened by ingress of cold, liquid water from the water pumps.
- A breakdown in one of the power supply unit caused one stack group to operate as a battery during start up: the current was running in reverse making the stacks operate in fuel cell mode and using the nickel in the electrodes as fuel. This operation mode of course destroyed the stacks
- A number of stacks were also weakened mechanically by being subjected to abrupt pressure differential (up to 0.5 bar abs) when switching from venting the produced hydrogen to sending it through the compressor. The mechanical weakening has

probably induced micro cracks in the 10  $\mu$  thick electrolyte, which developed into leaks that resulted in exotherms with stack temperatures exceeding 825 °C ultimately causing shutdown of the units(s).

These, and other minor problems, has caused delays to the original time planning and limited the number of accumulated operating hours.

The stacks replacement has caused a substantial budget overrun, but the extra costs has been covered entirely by Haldor Topsøe A/S internal funding.

Many of the issues have been addressed by improved operating procedures and software solutions, but others will need to be addressed by more aggressive, preventive maintenance measures, which will be implemented in a continuation of the project aiming at bringing the SOEC technology from TRL level 6 to 8.

The milestones agreed upon was, nevertheless, all implemented although the milestone concerning transient operation could have been implemented in a more systematic manner.

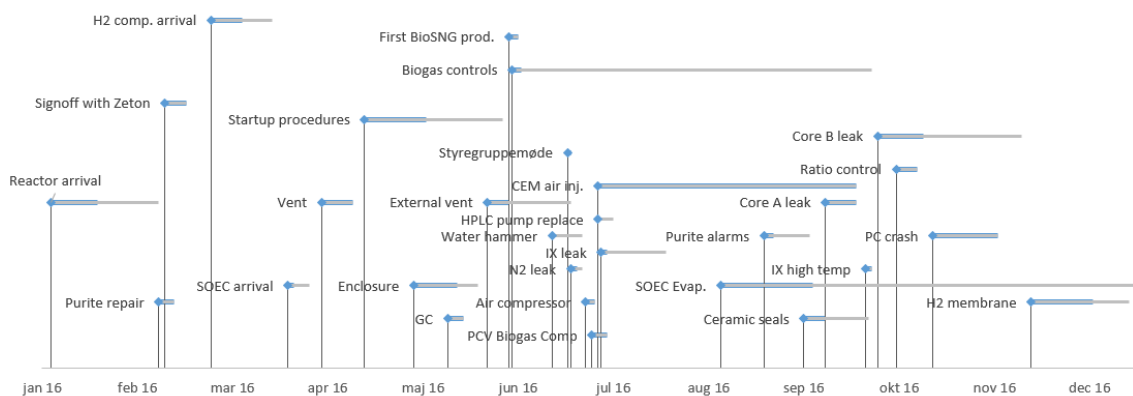


Figure 2-1: Timeline 2016. The gray bar indicate the total time required for each item, whereas blue indicate the actual on site tooltime (time awaiting parts and repair at 3<sup>rd</sup> party workshops have been subtracted in the blue bar).

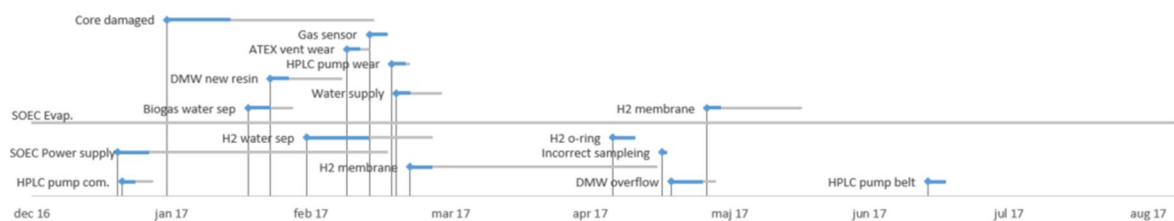


Figure 2-2: Timeline 2017. The gray bar indicate the total time required for each item, whereas blue indicate the actual on site tooltime (time awaiting parts and repair at 3<sup>rd</sup> party workshops have been subtracted in the blue bar).

### 3. Project results

#### 3.1 Plant design

Haldor Topsøe A/S designed the plant using their proprietary computer software and engineering procedures. One objective of the project was to establish a platform so that future scale-up to industrial scale would not require another demonstration plant. This required a minimum size to ensure correct heat transfer characteristics and linear velocities in especially the methanator. On the other hand, the cost of the SOEC stacks are at present rather high due to the limited production capacity and non-automated production technology. The minimum acceptable capacity was found to be 10 Nm<sup>3</sup>/h of biogas.

The main process steps can be seen in the following block diagram (Figure 3-1).

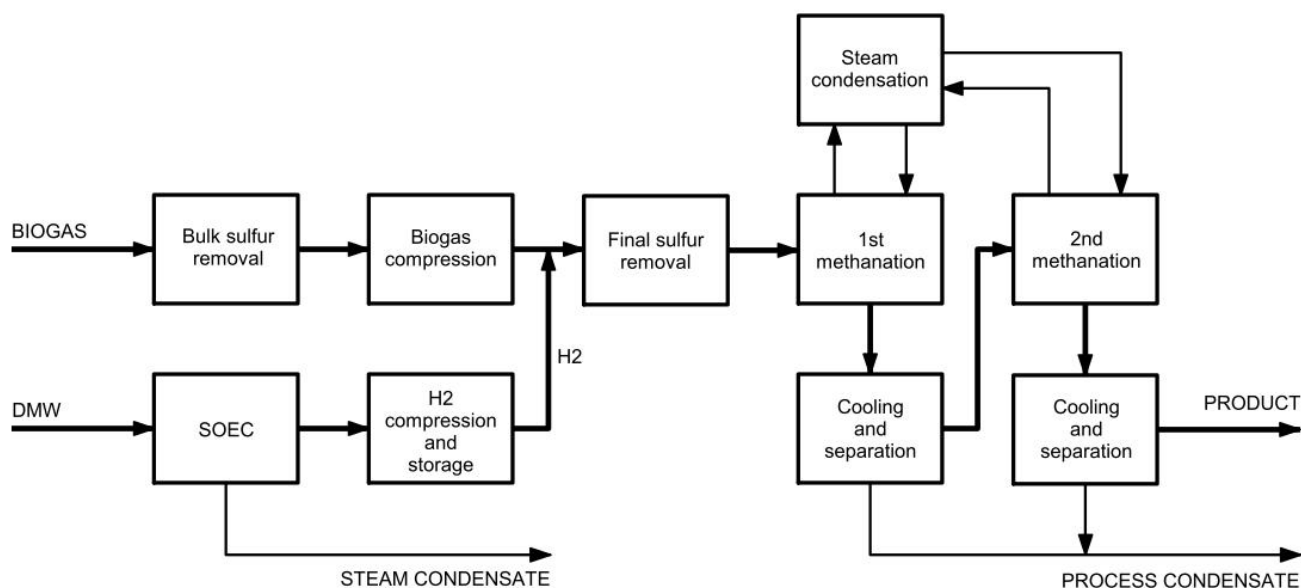


Figure 3-1: Block diagram of the main processes when upgrading biogas

This el-upgraded biogas plant produces synthetic natural gas (SNG) from biogas (carbon dioxide and methane) and hydrogen. The hydrogen is produced in a solid oxide electrolyte cell (SOEC) unit where electricity is used to produce hydrogen from water.

The el-upgraded biogas plant consists of three main units:

1. Bio-SNG unit
2. H2 compressor
3. SOEC unit

The bio-SNG unit consists of the following main process steps:

#### Bulk sulfur removal

Active carbon is used for this step. The process is carried out at ambient pressure and temperature.

#### Compression

The biogas is compressed to achieve the pressure required to achieve a high methane content in the SNG product.

#### Final sulfur removal

A promoted, proprietary Haldor Topsøe zinc oxide absorbent is used to remove the remaining sulfur. This step is placed after mixing with hydrogen.

#### Methanation

The hydrogen and carbon dioxide is converted to methane in the methanator, which is a boiling water reactor, loaded with a high temperature tolerant nickel based catalyst.

#### Hydrogen storage

A hydrogen storage makes it possible to operate the bio-SNG unit and the SOEC unit independently.

Additionally, the following utilities are included in the bio-SNG unit:

#### Water purification unit

Preparation of demineralized water (DMW) from tap water



### Cooling water unit

Circulation and temperature adjustment of cooling water to the process

### Vent system

As a safety measure a separate ATEX vent system encapsulates all process units

The SOEC unit consists of the following main process steps:

1. Evaporation of water
2. Hydrogen production in the SOEC stack from water and electricity.
3. Cooling of effluents and condensation out of non-converted steam

The hydrogen compressor consists of the following main process step

1. Compression
2. Hydrogen export to storage vessel in the bio-SNG unit.

Hydrogen and oxygen is produced from steam in solid oxide electrolyte cells in the SOEC unit. The SOEC unit consists of two cores (core A and core B). There are two parallel steam generators feeding the two cores.

The unit is supplied with demineralized water (DMW) and is split into two streams and the streams are pumped to the steam generators by the DMW pumps. The steam generated is superheated up to 200°C at 7 bar g and sent to pulsation dampers to minimize the pressure pulsations in the electrolyte cells. After the pressure has been reduced to 0.2 bar g, the steam is mixed with small quantities of hydrogen to protect the electrolyte cells from oxidation of the active electrode surface.

In the cores the steam is preheated in the SOEC steam feed/effluent exchangers and then finally heated to the stack inlet temperature, initially 720 °C, in the SOEC steam trim heaters before entering the cathode side of the SOEC electrolyte cells. Each core contains 4 stacks with 75 electrolyte cells in each. The water is split into hydrogen and oxygen and the oxygen is transported in the form of oxygen ion to the anode side where they recombine to oxygen. The steam conversion is designed to be around 60 % in order to have a favorable Nernst potential at the exit of the stacks.

The anode side is flushed with air to reduce the oxygen concentration in the anode product stream. Idaho National Lab has identified 50 % oxygen as a safe level in the cathode off gas and this limit was adopted in the project design. Compressed air is heated in the SOEC air feed/effluent exchangers and then subsequently heated in the SOEC air trim heaters to the same inlet temperature as the cathode side before entering the anode side of the stacks.

The hydrogen rich stream leaving the stacks is first cooled in the SOEC steam feed/effluent exchanger and the two hydrogen rich streams from each core are combined and then further cooled to 40 °C in the H<sub>2</sub> water cooler. The steam condensate is removed in the steam condensate separator and then pumped to drain by the steam condensate pump. The two hot enriched air streams from the two cores are combined and cooled in the air water cooler.

The hydrogen from the SOEC is compressed from 0 to the 43 bar g in a two stage diaphragm hydrogen compressor. As a backup to the SOEC unit, hydrogen is also available from a battery of hydrogen bottles, located approximately 100 meters from bio-SNG unit. A backup hydrogen line is feeding the bottled hydrogen to the hydrogen compressor.

The compressed hydrogen is sent from the hydrogen compressor to the hydrogen storage. Storage of hydrogen enables operation of the bio-SNG when the SOEC unit is not in operation. The hydrogen storage consists of two gas cylinders with a 50 liter volume each.

The biogas is produced in a nearby biogas reactor at Aarhus University, Foulum and the 10 Nm<sup>3</sup>/h biogas feed to the bio-SNG unit corresponds to 5-12 % of the total biogas production, depending on season. From the biogas reactor the biogas is sent to an uninsulated biogas storage vessel, where the biogas is stored at ambient temperature. From the storage vessel the gas is sent to the bio-SNG unit in an uninsulated underground pipeline. The water saturation of the biogas will depend on the ambient temperature of the air and the ground. The biogas enters the bio-SNG unit at 0 mbar g and 10°C. The biogas design content is approximately 56 % methane and 43 % CO<sub>2</sub> with a balance of nitrogen.

The raw biogas design content is on average, 1000 ppmv sulfur compounds, mainly H<sub>2</sub>S, but also organic sulfur compounds are present. Sulfur will poison and deactivate the catalyst in the methanation reactor. Thus, any sulfur present in the feed must be removed.

The bulk part of the sulfur compounds is absorbed on active carbon in the bulk sulfur removal section. The active carbon has an optimal sulfur uptake at temperatures between 10°C and 55°C, a water saturation between 10% and 90% and an oxygen to sulfur ratio of 1.8. The active carbon bed is operating at close to ambient pressure. Condensation in the active carbon bed must be avoided by keeping the gas temperature in the section above the dew point of the gas.

The biogas feed enters the bio-SNG unit and is compressed to 0.3 bar g in the biogas blower. DMW and air is mixed and evaporated in the controlled evaporation and mixing (CEM) unit and added to the biogas. The biogas is mixed with air to a molar ratio between H<sub>2</sub>S and O<sub>2</sub> of about 1.8 and is heated to 40 °C in the traced lines, to achieve a water saturation between 10% and 90%. The relative humidity is measured before the inlet to the sulfur absorbers.

The biogas then passes through the sulfur absorbers with the active carbon beds. These are identical and installed in series. The two external reactor heaters keep the sulfur absorbers at 40°C to avoid condensation in the active carbon beds.

There are two sulfur absorbers in series in order to continue operation after breakthrough of sulfur is detected outlet the first sulfur absorber. A bypass around the first absorber enables replacement of the activated carbon during operation. The oxygen content of the biogas and air mixture is measured by analyzing samples taken at the inlet and the outlet of the sulfur absorbers. The slip of sulfur from the sulfur absorbers was expected to be approximately 25 ppmv.

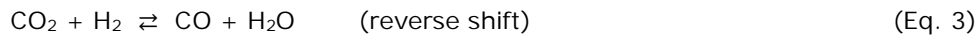
After the bulk sulfur removal the biogas is compressed to 22 bar g in the biogas compressor. The pipes from the sulfur absorbers to the biogas compressor are insulated and traced to avoid condensation.

Small amounts of refractory sulfur are expected to pass through the active carbon sulfur absorbers in the bulk sulfur removal section. This sulfur is removed in the sulfur guards, which are loaded with promoted zinc oxide absorbent from Haldor Topsoe A/S. The absorbent can absorb refractory sulfur best at elevated temperatures and for this reason the temperature has to be maintained above 250°C.

The adsorbent is composed of zinc oxide and absorbs hydrogen sulfide and COS as shown below



As the process gas contains CO<sub>2</sub>, the formation of water will occur according to:



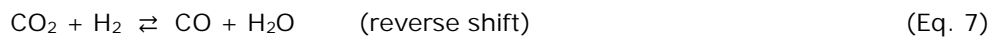
Surplus oxygen reacts exothermically with hydrogen to form water:



The purified biogas from the active carbon vessels is mixed with hydrogen from the hydrogen storage and the process gas mixture is then heated to 286 °C in the sulfur guard trim heater. The sulfur is absorbed in accordance with reactions (Eq. 1) and (Eq. 2) and the remaining oxygen reacts with hydrogen in accordance with reaction (Eq. 4) causing a temperature increase of about 9°C. There are two sulfur guard reactors in series in order to continue operation after breakthrough of sulfur is detected outlet the first sulfur guard. A bypass around the first guard reactor enables replacement of the zinc absorbent during operation. The clean process gas is subsequently sent to the methanation reactor.

In the methanation section the hydrogen and carbon dioxide is converted to methane by catalytic reaction in the methanation reactor. The conversion is limited by equilibrium, which is very pressure dependent. A pressure above 20 bar g is needed to achieve the conversion required to achieve the minimum requirement to the bio-SNG product.

The methanation reactor has two beds loaded with a nickel based methanation catalyst delivered by Haldor Topsoe A/S. In the methanation reactor the hydrogen and the carbon dioxide react to form methane in accordance with (Eq. 5). The methanation catalyst is also active for the reverse water gas shift reaction (Eq. 7), while carbon monoxide will react to form methane in accordance with (Eq. 6).



The ratio of hydrogen to carbon dioxide has to be stoichiometric ( $\text{H}_2/\text{CO}_2 = 4$ ) to achieve the maximum possible methane content in the bio-SNG product. Reactions (Eq. 5), (Eq. 6), and (Eq. 7) produce water which limits the conversion of reactants. A high methane content in the product is achieved by having two methanation stages with inter-stage water separation and removal.

The methanation reactions (Eq. 5) and (Eq. 6) are very exothermic. The methanation reactor is a boiling water reactor (BWR), where the reaction heat is removed from the gas in the reactor tubes to the boiling of water on the shell side. The advantage of the BWR is to ensure a low equilibrium temperature at the outlet of the reactor, resulting in a high conversion of  $\text{CO}_2$  and  $\text{H}_2$  to  $\text{CH}_4$ .

There is a risk of carbon formation from methane on the methanation catalyst. Carbon formation is avoided by keeping a sufficiently high content of hydrogen and water (steam) in the gas.

Outlet the first methanation stage the methane content is limited, the water and the hydrogen content high. There is accordingly low risk of carbon formation in the 1st methanation catalyst bed. The gas outlet the 2nd methanation bed is high in methane and low in water and hydrogen. To protect the methanation catalyst the inlet gas has to be saturated at minimum 85°C and the ratio of hydrogen to carbon dioxide has to be minimum 3.8. The low operating temperature in the 2nd reactor makes, however, the carbon formation fairly slow. Less water or hydrogen in the feed gas can thus be tolerated in shorter periods during upset situations.

After the sulfur guards the process gas is fed into the methanation reactor 1st catalyst bed. There are two catalyst beds (tubes) in the methanator, one for each reaction stage. Each tube is divided into an inert zone and a zone with methanation catalyst. The first zone acts as a heat exchanger, heating the feed gas to around 290°C. The methanation reactions (Eq. 5) and (Eq. 6), and the reverse water gas shift reaction (Eq. 7) takes place in the second, catalytic zone. After the 1st methanation stage the process gas leaves the reactor at around 280-290 °C and is cooled to 85°C in the 1st effluent air cooler, process condensate is removed in the 1st process condensate separator, and the process gas is sent to the 2nd methanation stage. After the 2nd methanation stage the major part of the hydrogen and carbon dioxide is converted to methane. The bio-SNG product is cooled in the 2nd effluent air cooler and process condensate is removed in the 2nd process condensate separator. The final bio-SNG product is produced at 20 bar g. The pressure is reduced and the bio-SNG product is exported to the biogas storage outside battery limit.

The steam produced in the BWR is transported in a steam riser by natural convection to the steam drum, which is placed minimum 1 meter above the BWR. The steam is condensed and the boiler feed water (BFW) is recycled back to the reactor through a downcomer in a closed loop. The heat is removed in the steam condenser by air cooling. The air cooler is a U-tube exchanger, placed inside the condenser drum.

The pressure in the condenser drum is regulated by adjusting an air flow through the steam condenser by means of a speed control on the air blower. An electrical start-up heater is placed on the downcomer. The capacity is 15 kW and will heat up the water, condenser and reactor in approximately 6-8 hours.

In order to simplify the process and facilitate independent operation of the SOEC unit and the Bio SNG unit it was decided not to integrate the steam systems.

The condensate from the methanation process, the biogas compressor and the hydrogen compressor is sent to a common condensate collector, where dissolved gases are released into the vent system. The condensate from the condensate collector is sent to the drain in the floor of the building, which is connected to the chemical sewer.

### *3.1.1 Utilities and online sample points*

The cooling water loop is a standalone unit, placed outside the process building. The cooling water loop supplies cooling water to the compressors and water coolers.

The cooling water consist of 50% tap water and 50% glycol. The cooling water is circulated by a cooling water pump. The cooling water return is cooled by an air cooler. The cooling water temperature is adjusted to 20°C by a mechanical control valve in the bypass. In case the cooling water out of the air cooler is above 30°C (a very warm summer for Denmark), the cooling water can be further cooled to 20°C using a backup extra heat exchanger. The cooling medium in the backup heat exchanger is tap water. An expansion vessel is connected to the system.

The SOEC unit is the main consumer of DMW. The SOEC stack consists of cermet electrodes containing nickel and are very sensitive to impurities in the DMW. For this reason the water purifier is based on a four step cleaning procedure:

#### UV source and bacteria filter

#### Reverse osmosis

Reverse osmosis removes the majority of inorganic salts, organic content and micro-organisms.

#### Electrodeionization unit (EDI)

Removing the majority of dissolved ions still left in the water.

The conductivity of the DMW should be  $<0.5 \mu\text{S}/\text{cm}$  or  $>10\text{M}\Omega$ .

DMW is mainly used as water source for the steam generation in the SOEC unit, but small amounts are also used as water source in the CEM unit. Furthermore, the water in the methanation reactor and steam condenser is DMW supplied from the water purification unit during commissioning of the steam system.

The vent system receives gases from the bio-SNG unit and the compressors due to:

- Releases from pressure safety valves (PSVs)
- Vent of biogas from the low pressure sulfur removal section
- Vent from the biogas compressor
- Vent from the hydrogen compressor
- Released gases from the process condensate collector
- Vent of bio-SNG product

The vent system is oxygen free.

The main ventilation system replaces air in the following enclosures:

- The bio-SNG enclosure (low pressure sulfur removal, methanation)
- Compressor enclosure (hydrogen compressor, biogas compressor)
- SOEC enclosure

The enriched air is removed from the SOEC unit via the ventilation system. If the hydrogen produced in the SOEC unit cannot be exported to the hydrogen compressor, the hydrogen is removed via the ventilation system. There are no emissions from the bio-SNG and the compressor enclosures unless there is a leakage. The air flow is adjusted to 15 air volume changes per hour. The air flow in the ventilation system to the SOEC enclosure is adjusted to ensure a hydrogen concentration below 50% of lower explosion limit (LEL) in the ventilation system. The ventilator is ATEX Zone 2 approved.

Nitrogen is used is for flushing during start-up and shutdown. Nitrogen can be added at the following positions in the plant:

- Inlet low pressure sulfur removal section
- Inlet biogas compressor
- Inlet hydrogen compressor
- Inlet the SOEC unit

A bottle rack of protection gas (5% hydrogen in nitrogen) supplies the SOEC unit with gas for flushing and heating up to  $650 \text{ }^\circ\text{C}$ .

There are 3 online sample points in the bio-SNG unit:

1. Inlet the 1<sup>st</sup> stage of the methanator
2. Inlet the 2<sup>nd</sup> stage of the methanator
3. Bio-SNG product

The samples are analyzed by a micro-GC, analyzing one sample at a time. The gas will continuously flush through all the sample lines, and a four position valve will collect samples.

Some pictures of the unit installed at the Foulum site are shown on Figure 3-2 and Figure 3-3.



Figure 3-2: The pilot plant at AU Foulum site. The tall reactor and steam drum makes the pilot plant reach the rooftop.



Figure 3-3: Main corridor and platform level of the pilot plant

### 3.2 SOEC performance

A start up from ambient conditions takes 10-11 hours (Figure 3-4). The stacks are first heated up with protection gas on the cathode side and air on the anode side to 650 °C. As the temperatures are now above the auto-ignition temperature of hydrogen the cathode gas is switched to 10 % hydrogen in steam.

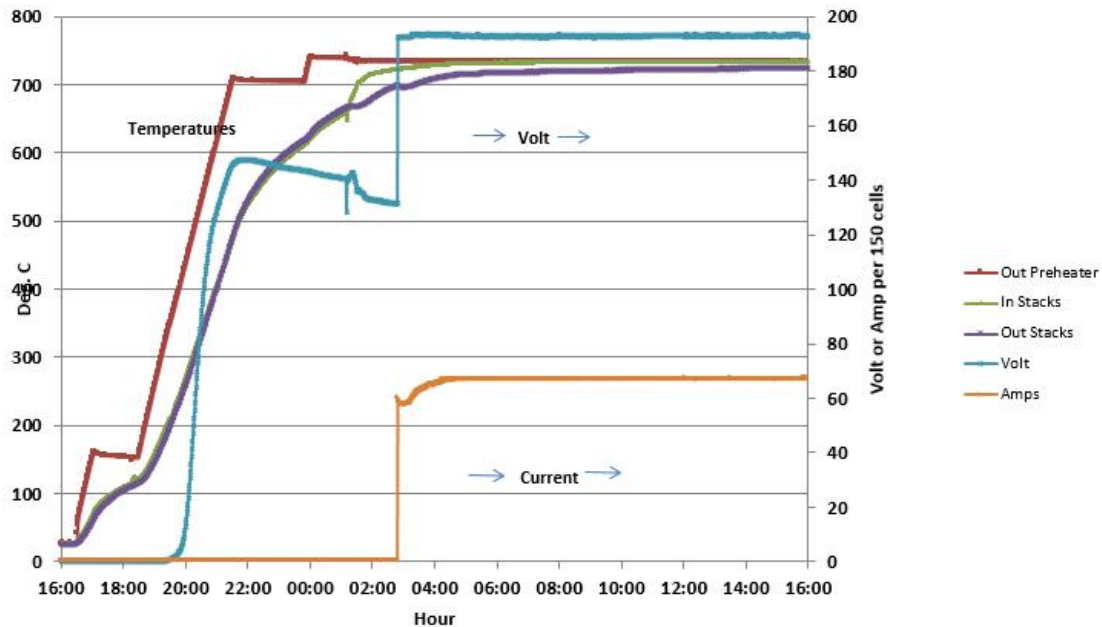


Figure 3-4: Heating of the SOEC unit from ambient conditions. When the unit is at 750°C the electrolysis is engaged.

The voltage recorded until power is switched on is the open circuit voltage developed across the stacks which are coupled two together, e.g. 150 cells in total.

Another start-up strategy has also been applied where a maximum of 180 V has been applied across 150 cells but with a maximum current limited to 3 A. This has minimised the slight delay in reaching full current=production, when applying the thermoneutral voltage around 193 V.

This minimum current of 3 A is also applied in “hot stand-by mode” in order to minimise the response time to below some few seconds when going back to full production.

The investigation of degradation of the stacks has been one of the main goals of the project. Most of the literature on SOEC cell and stack degradation has been carried out with mainly small button cells or short stacks only. Furthermore the measurements have typically been carried out in galvanostatic mode. In other words, the current has been kept constant by varying the applied voltage. This mode of operation would lead to an increase in specific power consumption for hydrogen production and excessive temperature gradient in truly adiabatic stacks.

A better strategy aims at keeping the production (e.g. current) *and voltage* constant by compensating for degradation by slowly increasing the operating temperature at otherwise thermoneutral conditions at 1.29 V per cell. The extra power required for the electrical trim heater upstream the stacks is marginal due to the feed/effluent exchangers.

This strategy was applied for the first time in this project on full size stacks and the results are shown on Fig. 3-5

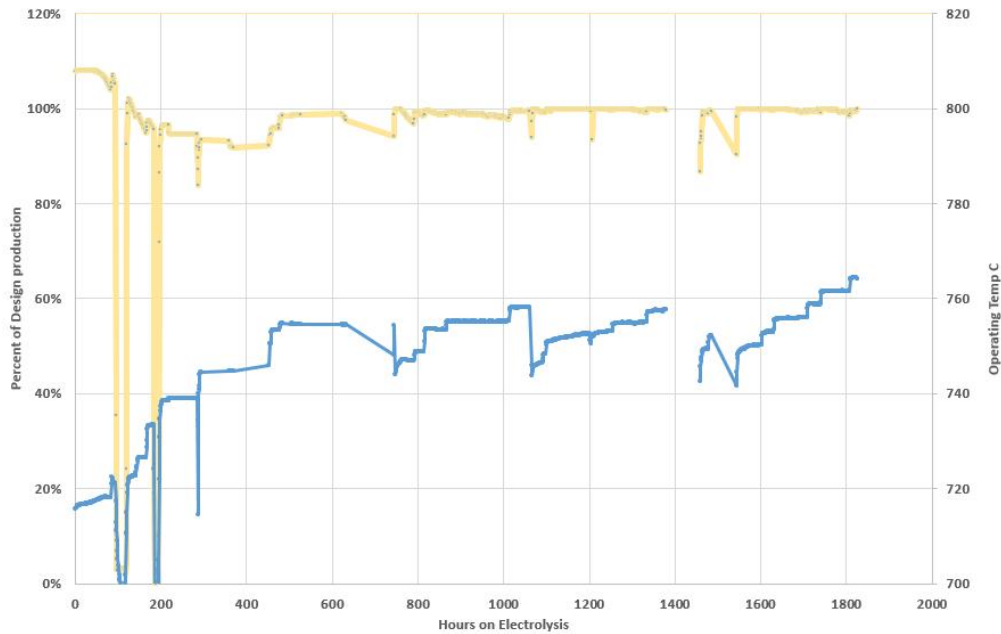


Figure 3-5: Operating hours on the SOEC unit. The production is kept at design production of 100% (yellow). Ageing of the stacks are compensated by increasing the temperature (blue).

It can be observed that the degradation rate as expressed by the rate of inlet temperature increase to keep production constant was rather high during the first 500 hours of accumulated operating time. It was thus necessary to increase the temperature from 715 °C to 755 °C.

After this initial period the performance of the stacks, however, stabilised and it was only necessary to increase the inlet temperature by a modest 10 °C in the next 1500 hours of accumulated operation.

It has been interesting to note that the performance of the stacks improved after every shutdown and restart of the units as indicated in a drop of around 15 °C in the needed inlet temperature. This could indicate that a certain reversible poisoning of the fuel electrodes have been occurring. The poison(s) could then be removed by the treatment with protection gas/hydrogen during the cooling down/heating up cycles. It should be pointed out that this is just a working hypothesis, which need to be checked by installing a poison trap on the steam in the next project.

### 3.3 Methanator performance

The startup time from cold to operating conditions was found to be approx. 8-9 hours as illustrated on Figure 3-6. When the cooling water on the reactor shell side reaches operating conditions of 265°C, the methanation reactor is ready. At this temperature (hot standby), the methanation reaction will stabilize within minutes after gas is introduced. The exothermic reaction gives rise to a hotspot just below 700°C



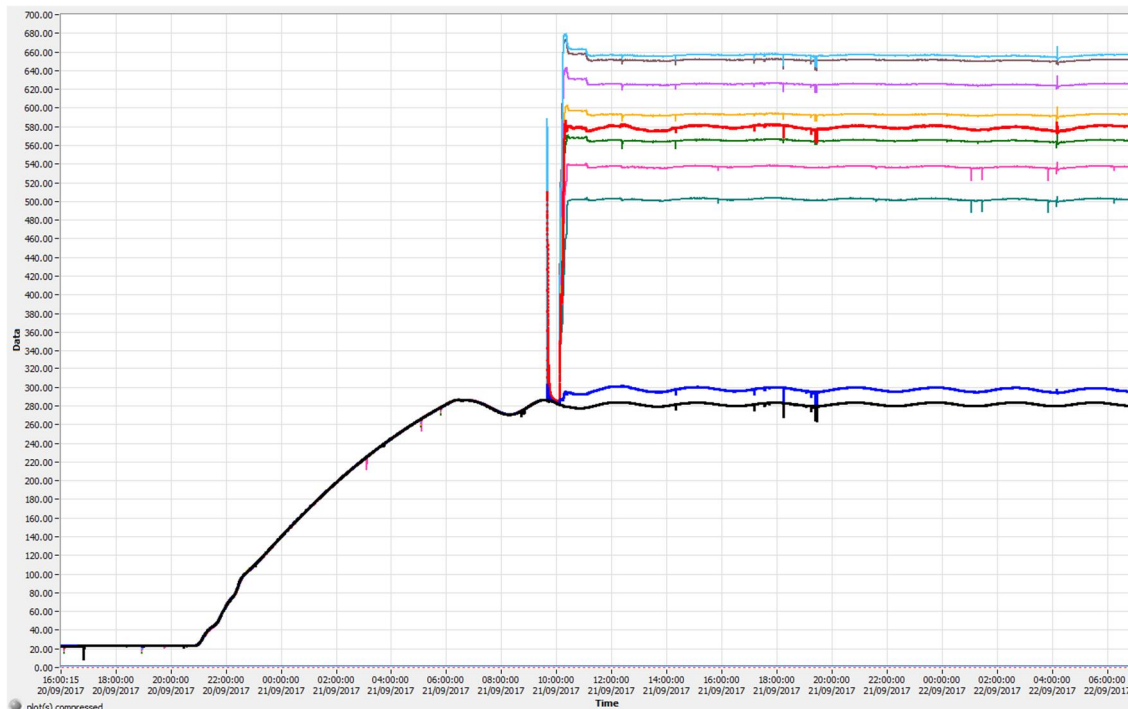


Figure 3-6: Heating of the methanation reactor from ambient conditions. When the cooling water on the reactor shell side reaches 230°C the methanation reaction can be performed. The exothermic reaction gives rise to a hotspot just below 700°C.

A typical temperature profile from the methanator 1st tube with a 2.4 m loading of catalyst is shown on Figure 3-7.

### Temperature profile in Methanator

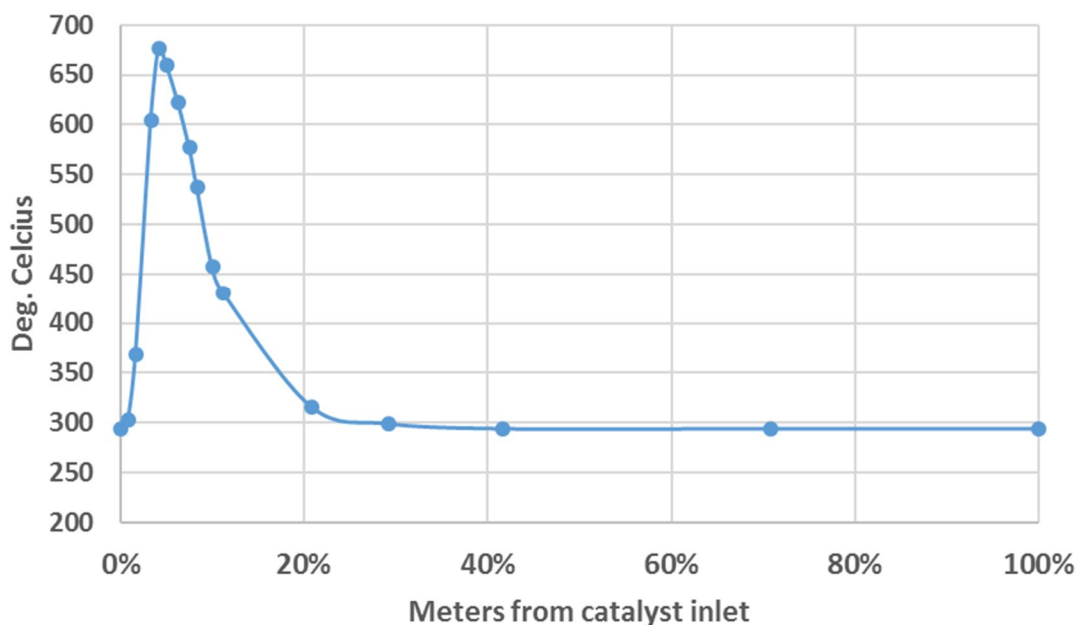


Figure 3-7: Temperature profile of the 1st reactor tube. A hotspot is observed at the inlet of the reactor.

It can be seen that the catalyst is very active. The hot spot temperature of 676 °C is already reached 10 cm down the bed and the conversion of the CO<sub>2</sub> to methane is nearly complete 25 % down the bed. This may seem to indicate that the reactor volume is oversized, but a certain buffer is included in order to account for thermal ageing as well as for poisoning.

Comparing the shape of the temperature profiles from start to end of the project as shown on Figure 3-8, it can be concluded that the profiles are unchanged indicating negligible ageing or

poisoning has occurred throughout the accumulated 1630 hours the methanator has been operated. This also proves that the desulphurization has performed excellently.

### Temperatures profile in Methanator

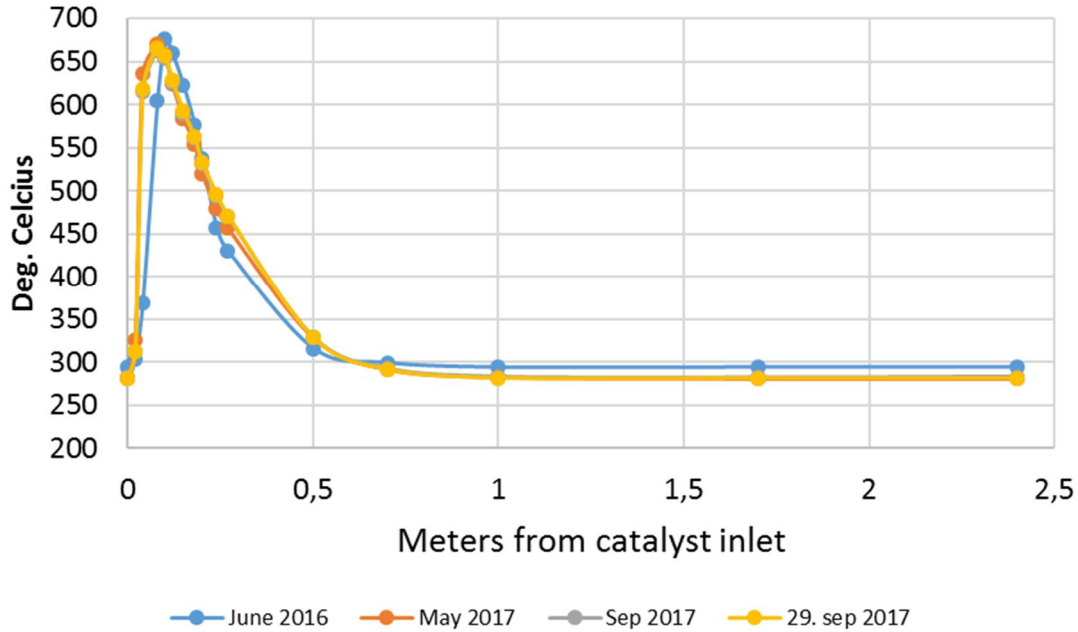


Figure 3-8: Temperature profiles thru the first reactor tube from the first testruns in 2016 till end of the project in 2017.

If the methanator is left in hot standby conditions, it only takes 7-8 minutes to reach steady state temperatures (Figure 3-9). The very rapid increase in hot spot temperature from 280 to 675 °C has not harmed the catalyst or reactor, despite more than ten such occurrences.

### Start of methanator 17 October, 2016

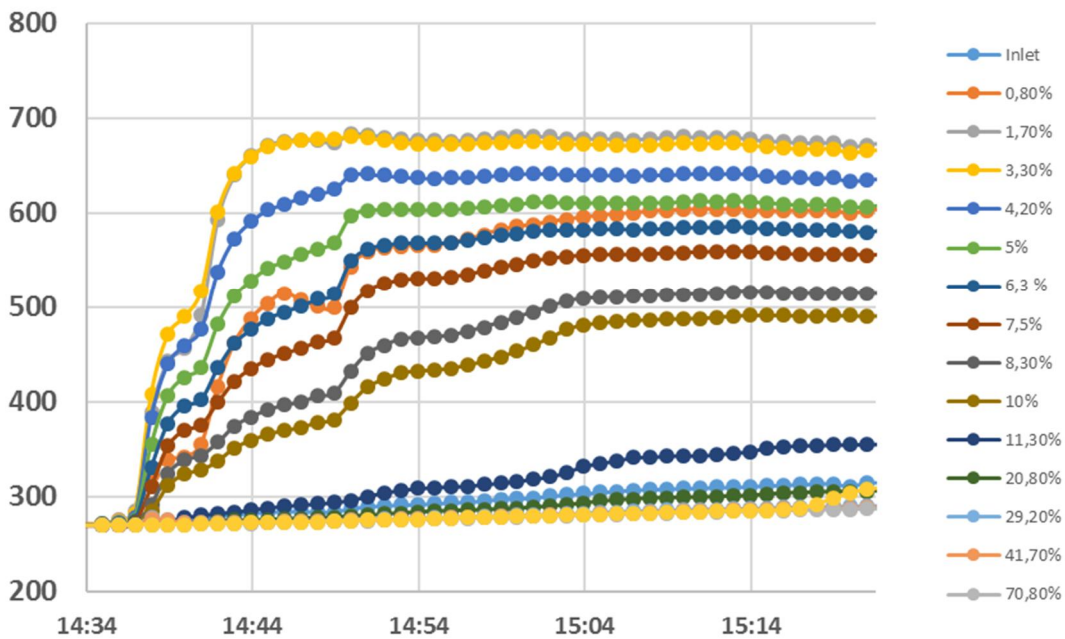


Figure 3-9: Introducing gas to the methanation reactor. The reaction hotspot reaches steady state after 7-8 minutes.

### 3.4 Gas quality

The gas operators are responsible for maintaining the gas quality delivered to end-users. The technical specifications on gas quality are defined in the standards Gasreglementet C-12 (Denmark), EN 16726 "Gas Quality Standard" (Europe), DVGW G260/G262 (Germany) as well as EN 16723-1 "Specifications for Biomethane for Injection in the Natural Gas Network" (Europe) and EN 16723-2 "Automotive Fuels Specification" (Europe). The formal vote for the new European standards EN 16723-1 and -2 closed in March 2017, and the standard is ready to become official, however there are still some issues under discussion (GasInfrastructureEurope).

The upgraded biogas needs to fulfill the same standard as the natural gas currently supplied to the network. Special requirements for national grids are still common, and some of the concerns covered are specific to natural gas delivered to the grid in Denmark.

A simplified process of upgrading biogas to natural gas can be seen in Figure 3-10.

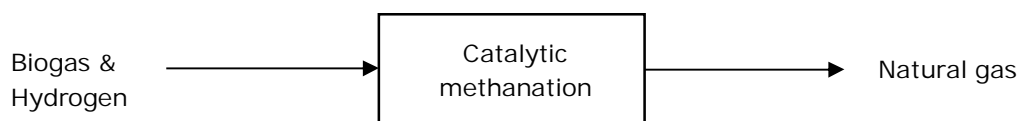


Figure 3-10: Simplified process showing the input and output gasses

#### 3.4.1 Biogas quality

The term biogas covers a mixture of gasses, with the main components being methane and carbondioxide.

A number of impurities are to some extent allowed in biogas. The impurities originate mainly from the digestion of molecules containing nitrogen or sulphur, but may also be introduced as organic material contamination or process disturbances (like stirring, overloading, change in feed material). Table 3-1 lists the typical impurities.

Table 3-1: Typical impurities found in raw biogas. Impurities are all components besides methane and carbondioxide. (Hjuler, 2017)

Contaminant	Typical amount in raw biogas	Spec. (DK C-12 & EN 16723)
Water vapor	5 - 10 %	Not allowed, dew point < -8° C
Carbon monoxide	0 - 3 %	Max. 0.1 %vol
Nitrogen	0 - 3 %	No direct limit, but restricted as: Methane spec min. 97.2 %vol
Oxygen	0 - 1 %	Max. 0.5 %vol
Hydrogen sulfide	0 - 10'000 mg/Nm3	Max. 5 mg/Nm3
Volatile organic compounds (VOCs)	0 - 4500 mg/Nm3	No direct limit
Hydrogen	20 – 600 ppm	Max. 2 %vol
Halogenated hydrocarbons	20 - 200 mg/Nm3	Not allowed
Long-chain hydrocarbons	0 - 200 mg/Nm3	Only small amounts of gaseous alkanes allowed (prim. ethane)
Ammonia	0 - 100 mg/Nm3	Max. 3 mg/Nm3
Siloxanes	0 - 41 mg/Nm3	Max. 1 mg/Nm3

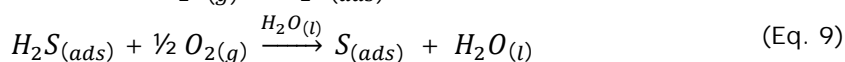
Many of the listed impurities are of little concern when sourcing biogas from a continuously operated facility solely based on livestock manure and agricultural bio-waste. This means carbon monoxide, VOCs, halogenated and long chain hydrocarbons, ammonia and siloxanes are of little or no concern in this project. The impurities nitrogen and hydrogen sulfide however require strict management.

- Nitrogen acts as an inert. It passes thru the entire upgrading process and ends up in the final product. Nitrogen results in a lower heating value/Wobbe Index of the final gas as well as undesired NOx emissions in the flue gas. The presence of nitrogen will require a higher conversion to methane of the catalytic methanation process. The result is a restriction in the operational window of the plant.
- Hydrogen sulfide is highly toxic, even in trace amounts. It leads to corrosion in engines, deterioration of lubrication oils and SOx emissions in flue gas. Because of these issues, the specification is very strict, requiring the concentration of hydrogen sulfide to be below 5 ppm as well as requiring hourly data reporting at point of injection to the grid. As for the upgrading of biogas with catalytic methanation, a specific problem with catalyst poisoning is also of concern. This results in a desire to remove all hydrogen sulfide in the biogas.

### 3.4.2 Removal of hydrogen sulfide from raw biogas

Although most biogas facilities include a process to reduce hydrogen sulfide levels, this project focuses on upgrading directly from raw biogas. In this way the provided solution will fit into any biogas facility and will cover all processes downstream the biogas reactor to grid injection.

Removal of hydrogen sulfide (and mercaptans) is done by adsorption to alkaline activated carbon. The technology is fully proven, active carbon is relatively cheap and has a high capacity. One downside with active carbon is the batch operation of the absorption reactors. In large scale facilities, the continued exchange of active carbon could induce a significant cost. The high adsorption capacity of hydrogen sulfide to active carbon is due to multiple adsorption processes both physical and chemical (Yan, 2002). In order to surpass the low adsorption capacity of physical adsorption the chemical adsorption must be favored at all times. The mechanism of chemical adsorption will require a thin water film on the active carbon to aid the chemical reaction. Further the reaction requires oxygen as aid to oxidize hydrogen sulfide to sulfur. The simplified steps are shown in (Eq. 8 and (Eq. 9 (Yan, 2002):



Other chemical aids are often added to alkaline activated carbon to aid capture of heavier mercaptans and to further boost the capacity. These aids are spent during operation and the capacity limit where breakthrough is observed will thus vary from plant to plant due to different feed and operational strategies.

Removal of hydrogen sulfide directly from raw biogas means the required water vapor is already present, whereas treatment of biogas from pipeline or storage tanks will require vapor injection as the biogas has been dried. The chemical adsorption further requires oxygen.

This results in two drawbacks of active carbon solutions:

- Handling of pure oxygen in hydrogen plants presents a severe safety risk
- A source of pure oxygen is costly

As the hydrogen sulfide level typically is in the ppm range (Table 3-1) a common short cut for both issues is to add air as the source of oxygen. This implies a small contamination of nitrogen from air is accepted in the treated biogas. To ensure sufficient oxygen for the chemical process producers will specify a desired oxygen/air surplus. The Solcarb KS3 used in this project recommends an oxygen surplus of 80%.

Historical levels of hydrogen sulfide in the raw biogas at the Foulum biogas facility is shown on Fig 3-11 . The hydrogen sulfide level is generally in the range of 400 – 800 ppm.

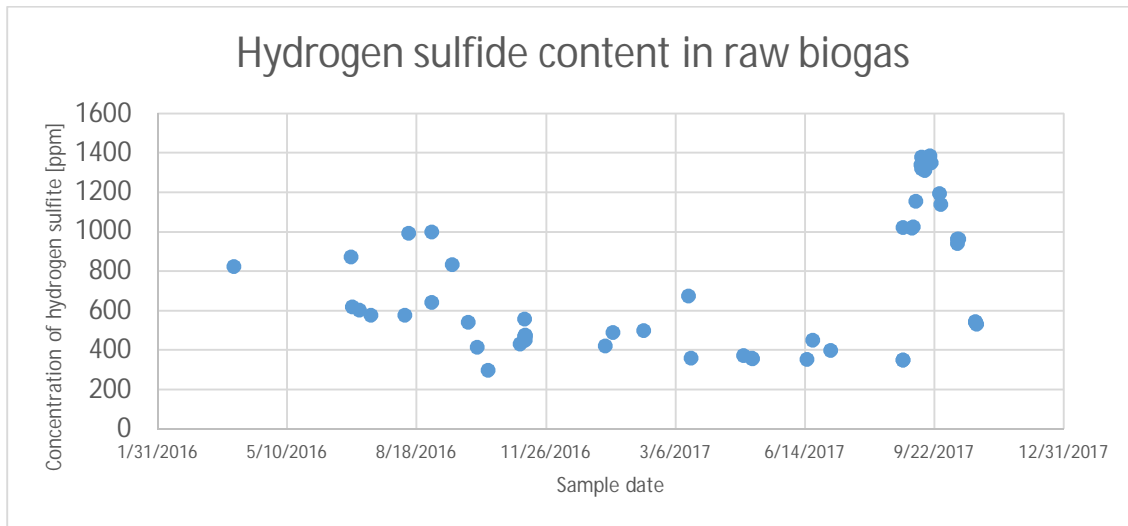


Figure 3-11: Historical levels of hydrogen sulfide at the Foulum biogas facility. Analysis is performed on the raw biogas.

During the two-year period the hydrogen sulfide level peaked at 1400 ppm. This specific high was due to plant maintenance (change in feed and operational procedure), and during this period the nitrogen contamination from air injection (air: 21 % oxygen and 79 % nitrogen) would result in:

$$N_2 = \frac{1400 \text{ ppm} \cdot 1.8}{10^6} \cdot 0.79 \cdot 100 \% = 0.95 \% \quad (\text{Eq. 10})$$

According to Table 3-1, the raw biogas may already contain up to 3 % nitrogen. Levels of nitrogen in raw biogas directly from the reactor is typically not as high as the handling/post-processing operations increases the risk of air contamination. Analysis on the raw biogas from the Foulum reactor has shown a nitrogen level around 0.4% (Appendix 4.2.1).

Possible options to reduce nitrogen contamination in raw biogas:

- Improved biogas reactor design, reducing the risk of air contamination.
- Change in biogas feed, as some nitrogen rich feeds could be minimized
- Using an oxygen rich stream as replacement for air dosing.
- Pretreatment of biogas, reducing the overall level of H<sub>2</sub>S

As such an additional nitrogen contamination of up to 1 % (Eq. 10) in the treated biogas reduces the allowed operational variance in the biogas upgrading plant, and at this level it will require continued monitoring to avoid quality issues.

A study of the supplier-recommended oxygen surplus (+80%) showed no need for dosing of surplus oxygen. Using only the stoichiometric requirement, the active carbon fully performed with no breakthrough after processing of 10'000 Nm<sup>3</sup> of biogas, equivalent to 45 days of operation at full upgrading capacity. It is possible that surplus oxygen is required as the active carbon nears the final adsorption capacity. However, the obtained results on nitrogen contamination shows that the risk to gas quality by air injection can be managed.

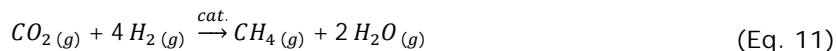
### 3.4.3 Hydrogen quality

The hydrogen needed by the catalytic methanation process should be free from any impurities. Typical production pathways for bulk volumes of hydrogen involve reforming or electrolysis. Both processes produce a hydrogen source of high purity. In this project the source is electrolysis of steam using solid oxide electrolyser cells (SOEC). The water used for steam production is purified with reverse osmosis and ion exchange resin making the only hydrogen-impurities of concern water and oxygen. Water is easily removed by cooling of the hydrogen gas. The hydrogen from the SOEC is confirmed dry as no water is observed in the drain after the compression. The product from the SOEC is pure hydrogen and oxygen in two separate streams. The ceramic membrane in the cell is however sensitive to pressure spikes

and thermocycles. Over time (and due to unplanned shutdown) microcracks will form. It is possible that micro cracks in the cell allow a small amount of oxygen to ingress in to the hydrogen. At the operating temperatures of the SOEC such oxygen would, however, immediately react with the hydrogen on the nickel electrode to form steam.

### 3.4.4 Natural gas quality

In case the stoichiometric ratio of carbon dioxide and hydrogen deviates from 1:4 (Eq. 11) noticeable amounts of either carbon dioxide or hydrogen will be present. The chemical equilibrium of the reaction can be pushed towards methane by removing the water produced by the reaction



A GC is installed at the reactor inlet, at mid stage and at the reactor outlet. The quality of the produced gas is measured every three minutes. The results from the test-run in June 2016 is shown in Figure 3-12. The results show that the produced gas is on spec. as natural gas.

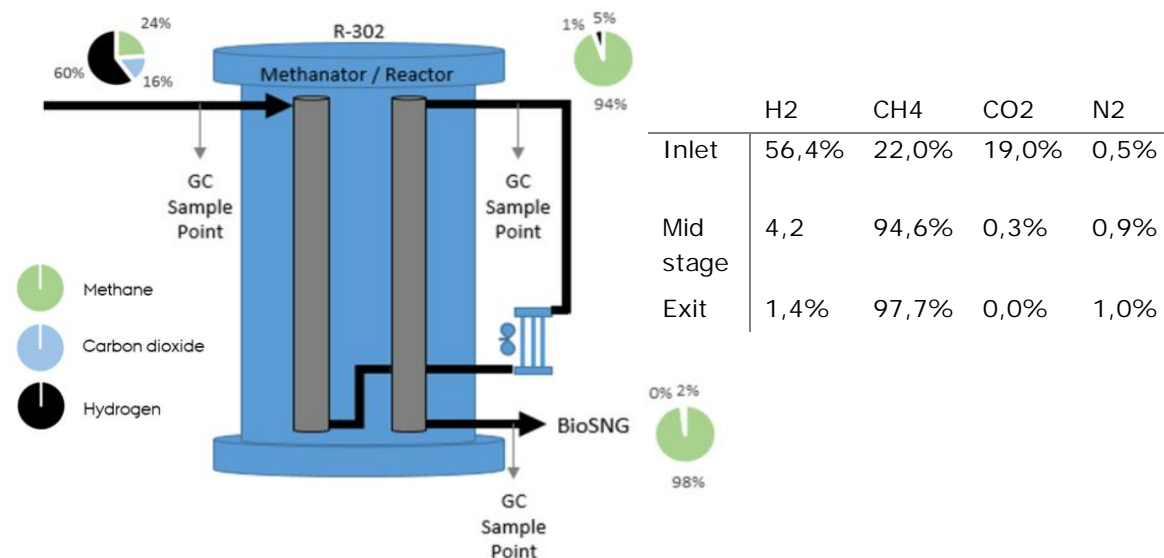


Figure 3-12: Sketch of the reactor tubes series and gas composition from simulation (left). GC analysis data of the gas composition at the reactor testrun in June 2016 (right). GC apparatus calibration was still ongoing at this time.

Nitrogen contamination was observed on the GC analysis data (Figure 3-12). During the test-run in June 2016, no air injection was performed (for H<sub>2</sub>S removal/biogas pretreatment), indicating that the nitrogen contamination originates from the biogas source. After the initial test-run air injection was enabled to ensure H<sub>2</sub>S removal with active carbon. This increases the nitrogen contamination to a general level of 1.5% as seen in Figure 3-12. The required purity of methane when delivering pipeline quality is 97.2 mol % (GasreglementC12; Hjuler, 2017), indicating gas of pipeline quality is produced as long as other non-methane contaminants can be kept low.

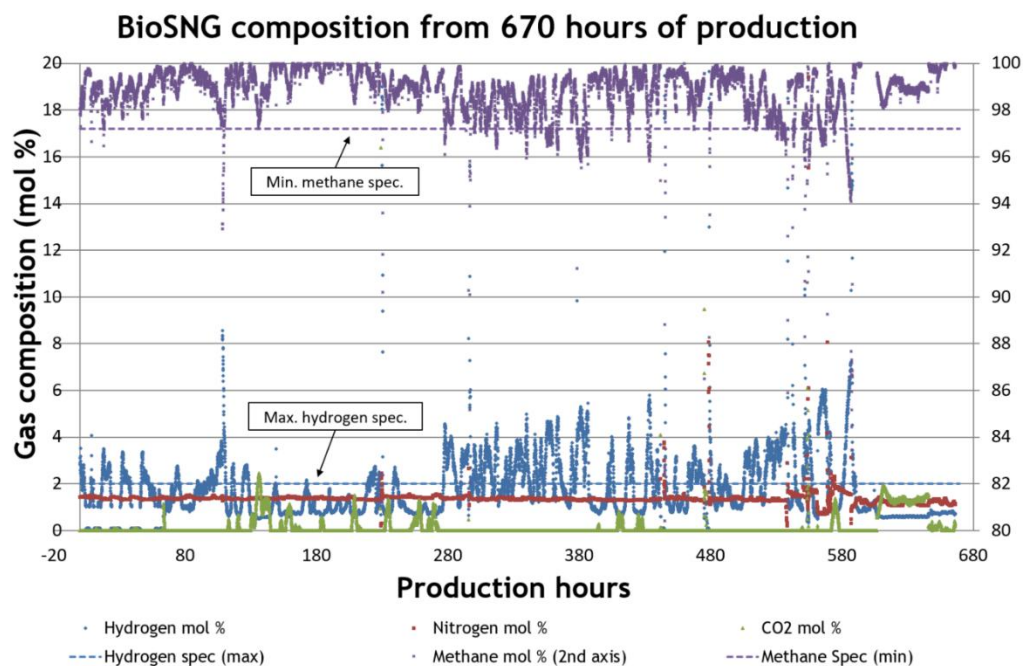


Figure 3-13: Gas quality from 670 hours of biogas upgrading operation. The quality data shows only minor drops on methane purity, but continuous periods where the specification on hydrogen level is exceeded. After 600 hours the operational philosophy was changed to allow a small surplus of carbon dioxide and not hydrogen. This improved the overall gas quality.

The main contaminants of interest are summarized in Table 3-2. During the first 600 hours of operation the limit of max 2 mol % hydrogen was exceeded (Figure 3-13). At the same time the CO<sub>2</sub> level was far from the allowed limit of 3.0 mol % (GasreglementC12). After 600 hours the operational philosophy was changed to allow a small surplus of carbon dioxide and not hydrogen. This improved the overall gas quality to be well within the quality requirements.

From Figure 3-13 it can be observed how surplus of hydrogen is closely related to a low CO<sub>2</sub> level.

Table 3-2: Selected parameters from the pipeline specification on natural gas in Denmark

Analysis	Unit	Specification
Methane	mol %	Min 97.2
Carbon dioxide	mol %	Max 3.0
Hydrogen	mol %	Max 2
Odorant (THT)	mg/Nm <sup>3</sup>	10.5

The total level on non-methane contaminants allowed is 2.8 mol %. With a stable nitrogen level of 1.5 mol %, the operational window of all other non-methane components is restricted to a narrow window of 1.3 mol % (hydrogen and CO<sub>2</sub> combined). During all production hours shown above the air injection was set at the recommended oxygen surplus of +80%. The results on gas quality led to the study showing no change in the performance of the active carbon when air injection was reduced to a level with no surplus of oxygen, thus reducing the nitrogen content.

### 3.4.5 Hydrogen in the natural gas

The permissible content of hydrogen in the pipeline quality natural gas is discussed in several work groups under the European Committee for Standardization. In most grids 10% hydro-

gen would not be a problem, but the limit value of gas turbines might be as low as 1 % hydrogen and of steel tanks and gas engines a limit value of 2 % hydrogen may apply. (Hjuler, 2017) In Denmark the current level is set at max 2%.

### 3.5 Design of full scale plant

Based on the results obtained from the demonstration project a full scale 20 million Nm<sup>3</sup>/year biogas upgrading plant has been designed and cost estimated by Haldor Topsøe A/S. The simplified flow scheme is shown on Fig. 3-14

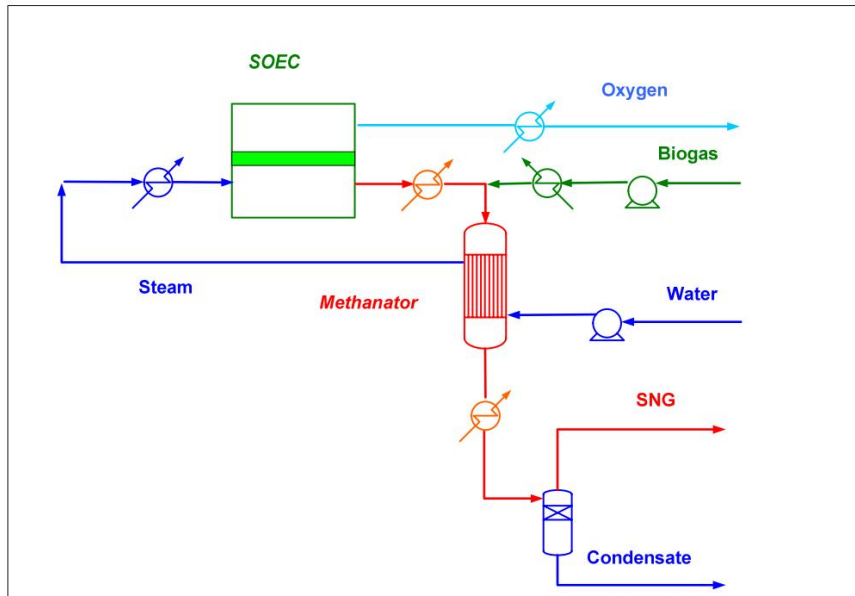


Fig. 3-14 Simplified flow sheet of plant configuration

The following efficiencies both on an exergy basis and energy basis calculated are shown in Table 3-3

Table 3-3

Configuration Percent efficiency	A CO <sub>2</sub> Meth.
Exergy eff gas	80.2
District heating	0.8
Oxygen	2.1
Exergy eff Total	83.1
LHV eff gas	76.3
LHV DH	14.0
LHV eff total	90.3

The exergy flows are depicted on Fig. 3-15



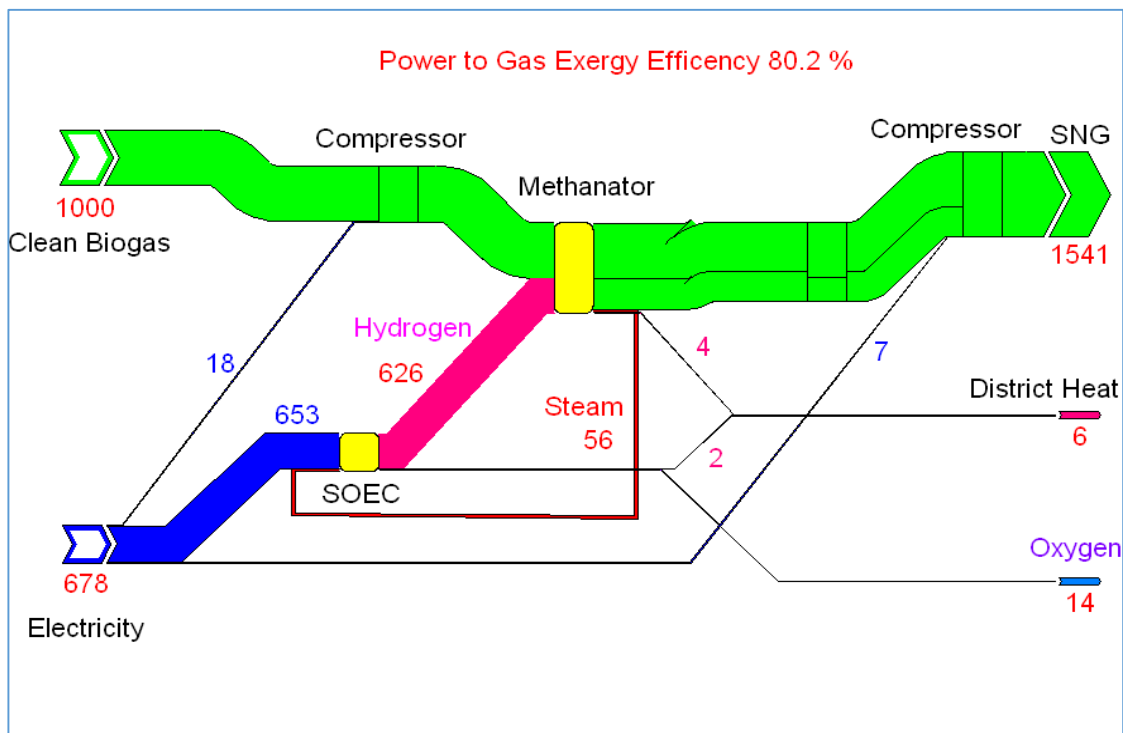


Fig. 3-15 Exergy flows in full scale plant

It should be noted that the design assumes that pressurized stacks will become available and the hydrogen compressor can thus be avoided. Operating the SOEC stacks at ambient pressure would decrease the exergetic efficiency by approximately 4 %.

The results from the design study have formed the basis for the economic analyses of the concept.

### 3.6 Economic analysis

In work package 5.2 the economics of deploying el upgraded biogas technology by SOEC has been investigated from two different perspectives:

- A socio-economic perspective evaluating the benefit to society
- A private, plant owner perspective

### 3.7 Socio-economic analysis

Ea Energianalyse carried out the socio-economic analyses using the Balmorel model. This model designs an energy system satisfying the demand for electricity, heat and transport services at the lowest possible cost to society given a set of assumption concerning fuel prices, energy technology investment costs and efficiency as well as externalities exemplified by a cost of CO<sub>2</sub> using specified production and consumption profiles. The model considers the central and Northern European energy system with respect to the electricity sector constrained by the capacity of the interconnectors. A further constraint is that Denmark produces at least enough electricity to cover domestic consumption on a yearly basis. The analyses have been carried out for 2030 and 2050 using the following assumptions:

For fuel prices the projections from IEA's 450 ppm scenario in world Energy Outlook 2016 has been used as shown on Figure 3-6.

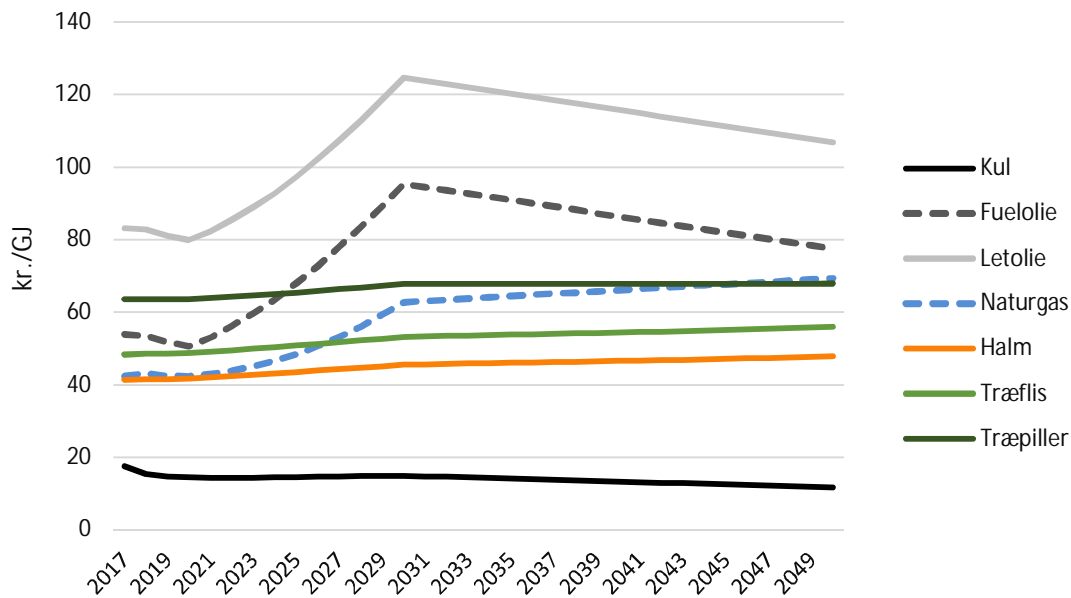


Figure 3-16: Fuel prices at Central plants (incl. transport costs).

Coal and oil prices are projected to decline due to the low demand caused by a large share of renewables. The biomass prices are based on studies by Ea Energianalyse. It is assumed that the natural gas grid will be fully depreciated in 2030.

It can be argued that flexible operation of SOEC plants will help to balance production and consumption in the electricity grid thus reducing the cost of operation and investment in the grid. This would imply a negative socio economic electricity tariff but in this study a tariff of 0 DKK/MWh is assumed in the base case scenario.

With respect to CO<sub>2</sub> prices 216 DKK/MT in 2030 and 800 DKK/MT in 2050 are used based on IEA's projection in WEO 2016.

It is assumed that no new investments in coal- or oil-fired power plants will be done in good agreement with present developments.

The price of electricity will not surprisingly have a major impact on the economy for el up-graded biogas plants. In the base case scenario, the ordered prices realized in Denmark in 2015 as well as the projected prices for 2030 and 2050 are shown on Figure 3-147.

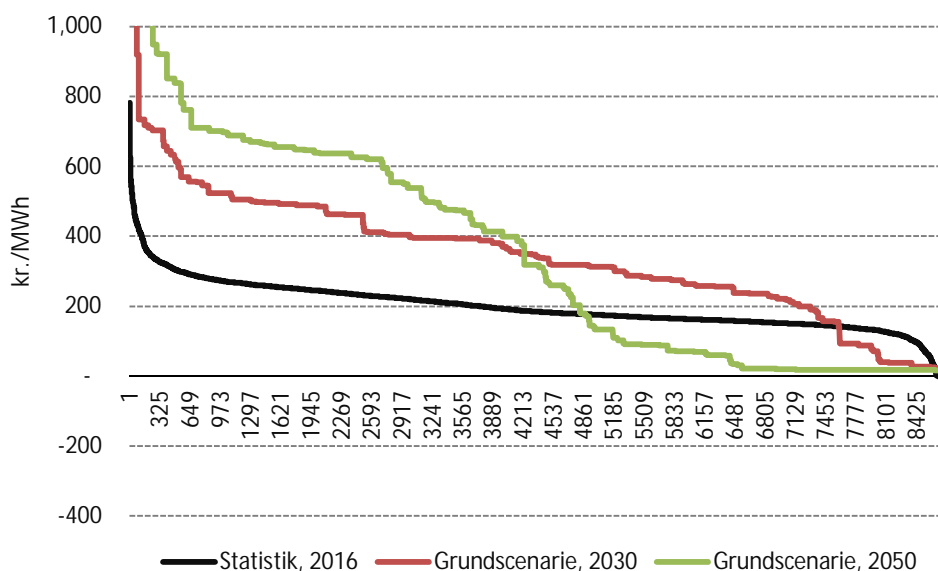


Figure 3-147: Electricity Prices in West Denmark

Due to the foreseen large share of fluctuating wind and solar power there will be a significant number of hours with low el prices: approx. 40 % of the time the prices will be below 100 kr./kWh. On the other hand, there will also very high prices caused by the coal and natural gas fired plant setting the price and the high CO<sub>2</sub> price.

With respect to demand for gas of pipeline quality the break down is shown in Table 3-44

*Table 3-4: Demand for pipeline quality gas in the base case Scenario in PJ/year*

	2030	2050	Refer- ence
Transport	1,6	3,0	1
Individuel opvarmning	16,0	13,0	2
Processheat for industry*	15,0	15,0	3
Total	32,6	31,0	

The demand for gas in the transport sector is relatively low as it is assumed that transport mainly will be covered by liquid biofuels and electricity.

On the supply side the quantities shown in Table 3-55 are assumed.

*Table 3-5: Biogaspotential assumed in the analysis (PJ biogas).*

	2030	2050	Ref.
Biogas from manure, bedding, organic industry- and household waste etc.	28	28	4
Biogas from straw *	20	7	
Total biogaspotential assumed	48	35	

*40 PJ straw available for energy use. 4 PJ straw in 2030 and 27 PJ straw in 2050 assumed reserved for 2nd generation ethanol. The potential for straw assumes a utilization degree to biogas of 55 %.*

The decrease in the potential based on straw is due to the allocation of this raw material to 2nd generation ethanol production, an assumption which could be challenged.

The cost data for the production plants are shown in Table 3-6

Table 3-6: Costdata assumed for the SOEC electrolysis, methanation and simple upgrading plants (in kr. 2016).

		2030	2050	
SOEC-plant*	Investment	3,59	2,76	Mkr./MWe
	Fixed O&M	96.000	71.700	Kr./MWe/år
CO2-metanation*	Investment	5,52	5,52	Mkr./MW H <sub>2</sub>
	Fixed O&M	165.600	165.600	Kr./MW H <sub>2</sub> /yr
Simple bio-gasupgrading**	Investment	2,87	2,59	Mkr./MW gas
	Fixed O&M	71.600	64.900	Kr./MW gas/yr
	Variable O&M	24	28	Kr./MWh gas

\* Data from Haldor Topsøe, 2017. Connection cost 0,37 Mkr./MW-e added.

\*\* Energistyrelsen, January, 2017: Teknologikataloget, kapitler om opgraderet biogas ( ).

The analyzed, integrated el upgrading biogas plant is shown on Figure 3-158.

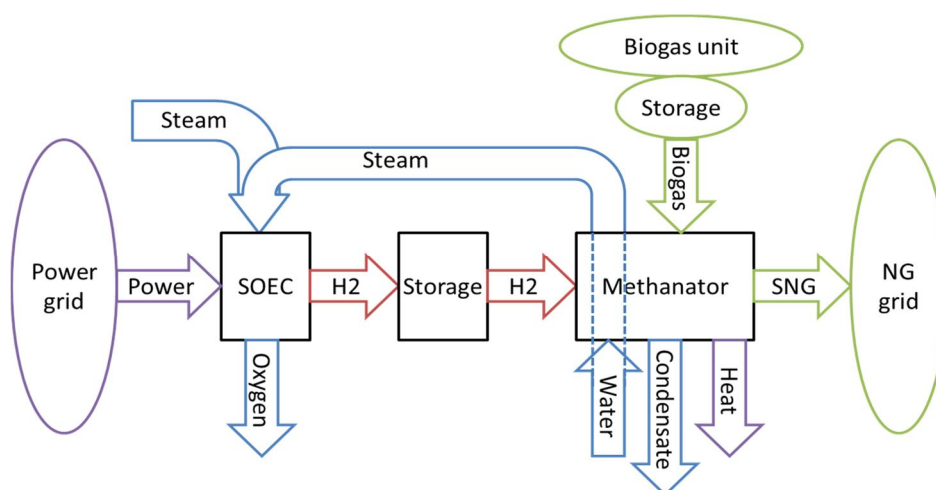


Figure 3-15: Overview of the plant for electrical upgrading of biogas.

In addition to the unit shown on Figure 3-158 the model can invest in a biogas engine for CHP production as well as a simple biogas upgrading unit which discards the separated CO<sub>2</sub>.

These units will now have to compete with natural gas, thermal gasification plants, heat pumps, biomass fired boilers, solar heating etc.

As the overall energy system is highly complex and the optimum depends on many parameter a scenario analyses have also been carried based on the scenarios shown in Table 3-7.

Table 3-7: Scenarios analyzed.

Scenarioname	Description
EloppgradOmk +30	30 % higher investment costs for el upgrading plant
EloppgradOmk +50	50 % higher investment cost for el upgrading plant
Fossile_hoeje	Fossil fuel prices according to New Policies scenario in IEA's WEO 2015. Natural gasprice: 85 kr./GJ (instead of 69 kr./GJ)
CO2_400kr	CO <sub>2</sub> -price of 400 kr. /ton in 2050 and 108 kr./ton i 2030 (e.g. 50 % lower than Base case)
Eltarif_30kr	Socioeconomic el-tariff 30 kr./MWh
VE-tilskudsrevet	Renewables favored by feed in tariffs instead of tax/quota driven. General RE-subsidy of 150 kr./MWh-e and a low CO <sub>2</sub> -price of 150 kr./ton. Results in lower el prices.
Naturgas_ude_2050	Natural gas phased out in Denmark in 2030.

Based on the assumptions above, the model selects the optimum outcome based on a 4 % discount rate (the recommended for socio-economic analyses).

The overall electricity production in Central and Northern Europe is predicted to be as shown on Figure 3-169 depending on the scenario.

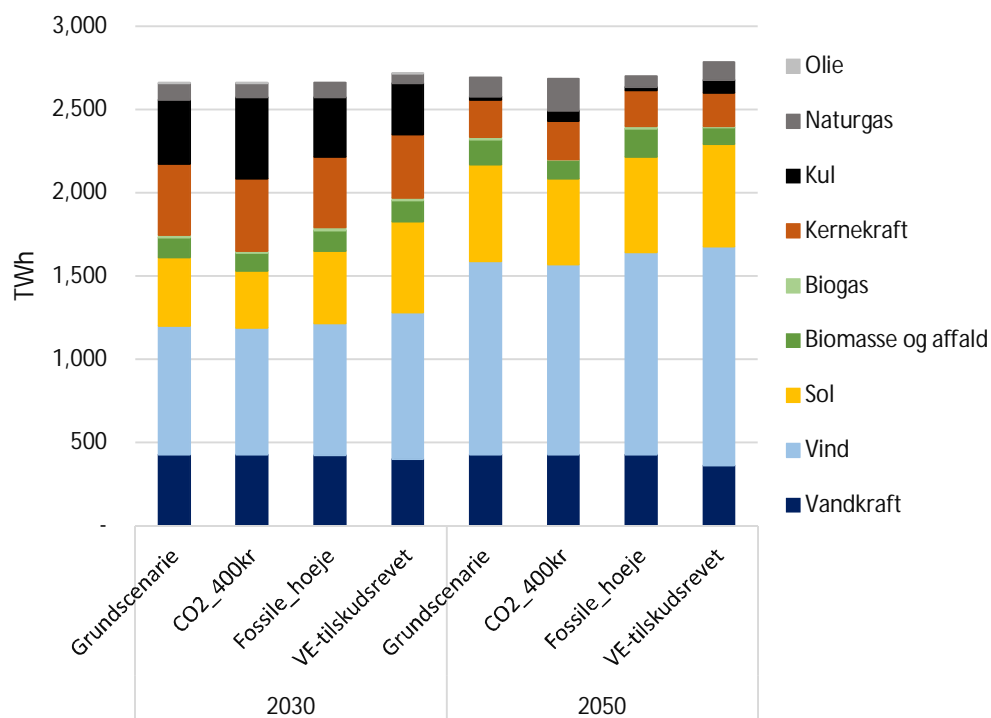


Figure 3-169: Electricity production from different feedstocks

It can be noted that renewables amount to 82-89 % of the electricity production in 2050 and wind- and solar power delivers 62-69 % due to their increased economic competitiveness. The high share of wind and solar power results in the large price fluctuations for electricity as shown on f.inst. Figure 3-16.

For Denmark the consumption of pipeline quality gas from the different sources: Bio-SNG from thermal gasification, el-upgraded biogas, simple upgraded biogas and natural gas are shown on Figure 3-20 in the different scenarios in 2050.

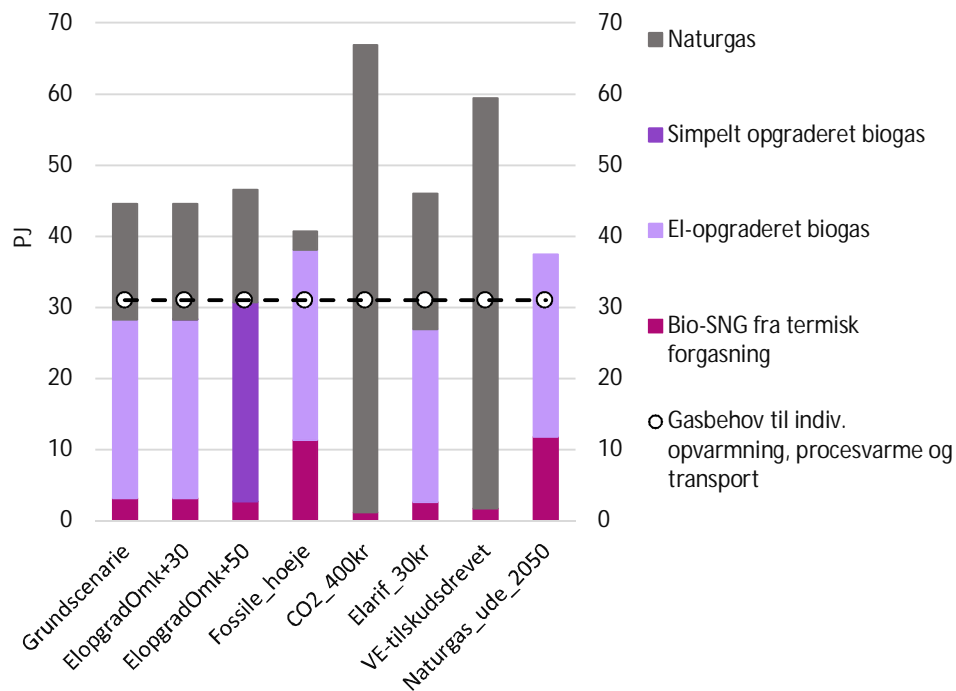


Figure 3-20: Different scenarios on how NG is sourced in 2050.

The dotted line in Figure 3-20 shows the combined (31 PJ) demand from transport, household heating and process heat. The rest is used for power and heat in general.

The highest gas consumption is in the two scenarios with low CO<sub>2</sub> prices where the gas used is almost exclusively natural gas. With the assumed natural gas prices in 2050 it is calculated that the CO<sub>2</sub> price should be at least 500 kr./ton for upgraded biogas to be competitive compared to natural gas. Biogas is not competitive in any of the scenarios in 2030 due to the assumed low CO<sub>2</sub> price of 216 kr./ton.

El-upgraded biogas outcompetes natural gas, simple upgraded biogas and gasification SNG in the base case scenario as well in the scenario with 30 % higher investment cost. If the investment cost is 50 % higher the simple upgraded biogas becomes cheaper.

In the scenario with higher fossil fuel prices and in the scenario with no natural gas there is a significant contribution from SNG generated via thermal gasification of biomass.

The production of biogas, simple upgraded biogas and el upgraded biogas in the different scenarios are shown on Figure 3-.

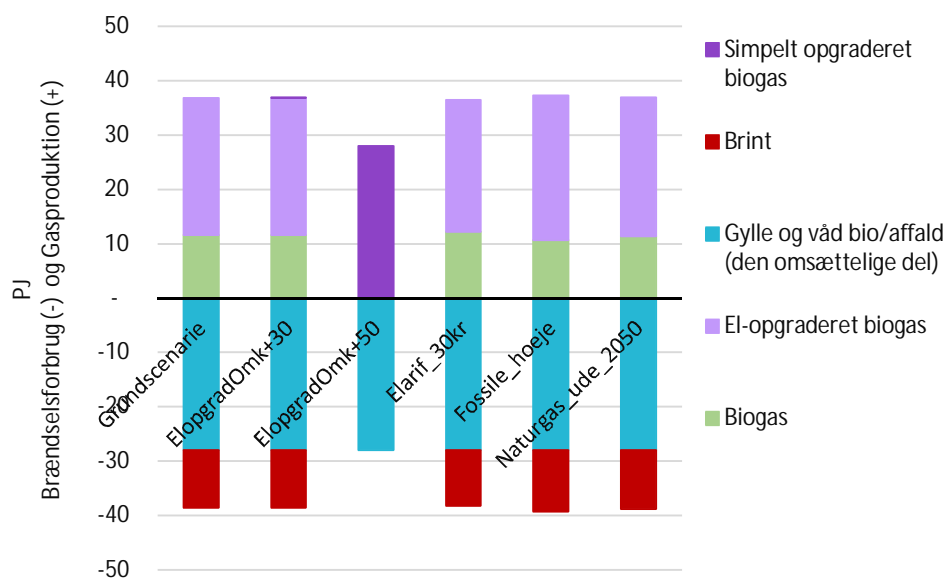


Figure 3-21: Production of different biogases in 2050

It is seen that the full biogas potential is used in all scenarios and the use of el upgrading increases the potential for renewable methane. The combined output of el upgraded biogas corresponds to approx. 1,6 GW and the needed SOEC capacity for the hydrogen production is 640-650 MW.

The fraction of the biogas which is not upgraded is used in engines for CHP production. The optimum amount of operating hours for the SOEC units are 4600-4800 hours per year which has a weighted electricity price of 109 kr./MWh which is considerably lower than the average electricity price of 388 kr./MWh.

As a check on above modelling results the levelized cost of energy (LCOE) has been calculated in the base case scenario. The ranking found is with LCOE in kr/GJ in parentheses:

El-upgraded biogas (91) < simple upgraded biogas (99) < natural gas (111) < thermal gasification of straw (115) < thermal gasification of wood chips (129)

The result is found to be sensitive to variations in the cost of electricity used in the SOEC unit. If the consumption weighted electricity price increases to around 140 kr./MWh then simple upgraded biogas is cheaper than el upgraded.

The socio economic benefit of upgrading the biogas with electricity via SOEC technology is calculated to be 353 mio. Kr per year or 40 kr./GJ for the extra gas produced.

In above base scenario there is a significant import of woody biomass to Denmark for production of el, heat and transport. If the biomass demand should be covered by national resources and natural gas was not available, the economic potential for upgrading of biogas as well as production of biofuels from thermal gasification would increase considerably.

The overall conclusions from the socio economic study are:

- In an energy system dominated by renewable electricity production and high CO<sub>2</sub> price el upgraded biogas via SOEC can outcompete natural gas, simple upgraded biogas or SNG from thermal gasification
- The socio economic gain in Denmark would be 350 mio kr. per year or 40 kr./GJ per GJ extra methane produced

- The amount of renewable methane from biogas production is increased by approximately 30 % by using el upgrading
- The SOEC unit should operate at full load for 4600 – 4800 hours per year where the average weighted electricity price is 109 kr./MWh
- There exists a strong synergy with biogas engines for CHP production which will operate in the hours with high electricity prices
- The above scenario will not be realized in the near future but could be towards 2050

In order to arrive to this scenario, it is necessary, that the investment cost for the technology can be decreased to the foreseen level, the SOEC unit can be operated in a flexible manner with rapid start and stops (from a minimum hot load < 4 %) and a socio-economic tariff of 0 kr./MWh or lower can be obtained. Moreover if battery storage costs become very low or other countries than Denmark invest heavily in power-to-gas technology it is doubtful that the duration curve with low electricity prices for a significant part of the year will become reality.

### 3.8 Private investor perspective

Planenergi has analyzed the economic aspects of the technology as seen from a private investor perspective.

They have modelled the system shown on Figure 3-172. In addition to the SOEC electrolysis and methanation unit they have included both an option for biogas storage as well as hydrogen storage. When the electricity prices are high the raw biogas can be sent to either a A) biogas engine or B) a SOFC generating power and heat or C) a simple amine upgrading unit producing pipeline gas quality by discarding the CO<sub>2</sub>. Finally, there is also an option D) just to flare the gas. It is assumed that there is a constant flow of biogas year round at a constant biogas price.

Three operating strategies have been investigated:

- a) The SOEC and the methanation are both operated constantly.
  - This leads to high hydrogen and SNG production costs in hours with a very high electricity price.
  - No storages are required in this operation strategy.
- b) The SOEC and the methanation are only operated when electricity prices are sufficiently low.
  - Here, a biogas storage can be used for storing the constant biogas in-flow during those hours when the methanation is not operating.
  - Alternatively, the biogas can be flared or utilized for a different purpose during those hours when the methanation is not operating.
- c) The SOEC is only operated when electricity prices are sufficiently low, but the methanation is operated constantly.
  - This requires the SOEC to be scaled such that it can produce more hydrogen than the methanation process can use during each hour.
  - The excess hydrogen is injected to a hydrogen storage for usage during hours where the SOEC is not operating.

A combination of strategy b) and c) could also be envisaged.



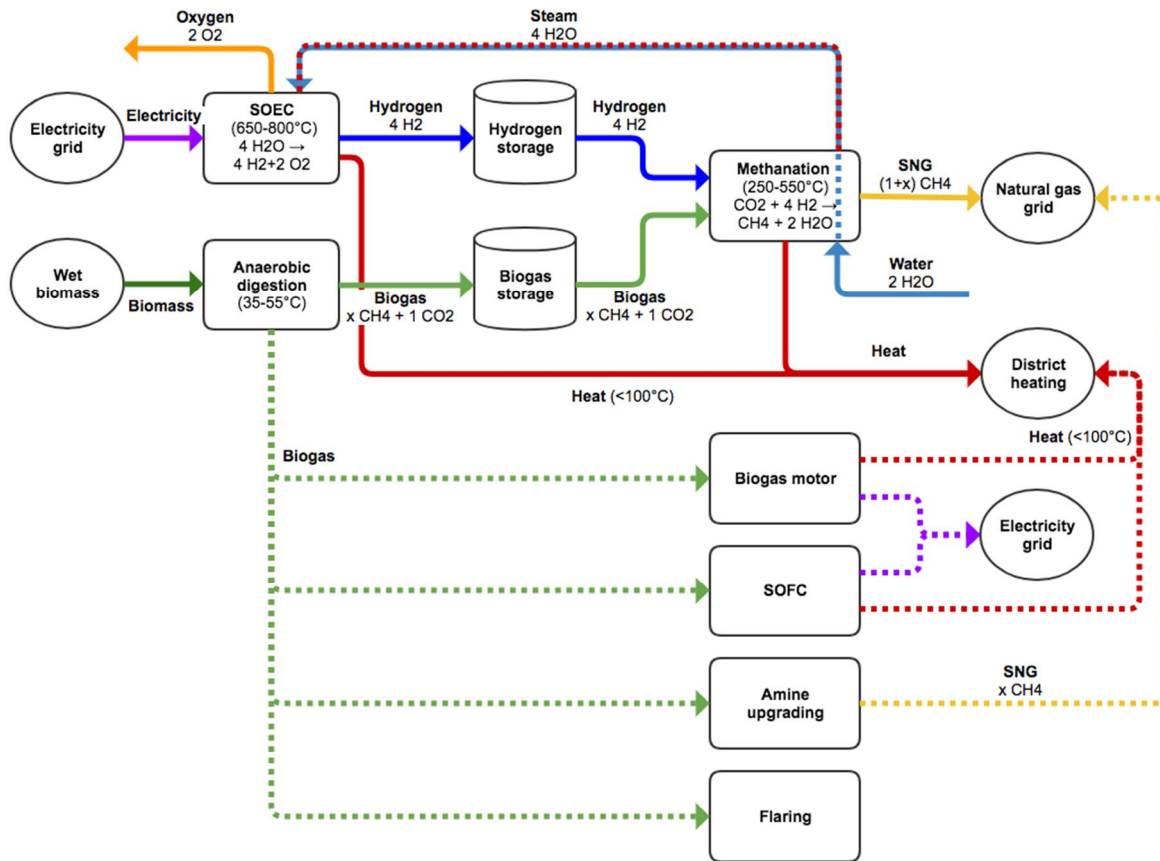


Figure 3-17: A schematic representation of the layout of the model system. The model assumes a constant in-flow of biogas, regardless if the methanation process is running or not. The energy conversion units with the dotted lines (biogas motor, SOFC, amine upgrading, flaring) serve as back-up options for consuming the biogas in case the methanation is not running. Only one of the back-up options is active during each model run.

The techno-economical model used is simulated in the software *TRNSYS*, which simulates the technical (e.g. energy flows, efficiencies) and economic (e.g. gas production cost, profit) operation of the SOEC, Sabatier methanation unit and gas storages throughout a whole year in a 1-hour resolution.

The four back-up types A-D therefore make up four different variations of the *TRNSYS* model. In case A), the biogas must be purchased from the source (at 3.25 DKK/Nm<sup>3</sup>), but is simply flared off. In case B), the upgraded SNG is assumed to be sold at the same price at the methanised SNG, i.e. 6.00 DKK/Nm<sup>3</sup>. In cases C) and D), the produced electricity is assumed to be sold to the grid at the same variable price as electricity for the SOEC is purchased, and the produced heat is assumed to be sold at a fixed price of 200 DKK/MWh = 56 kr./GJ.

All numerical inputs to the *TRNSYS* model, are listed in this chapter. All assumptions for the technical input values (e.g. installed capacities, gas flows, efficiencies) to the *TRNSYS* model are shown in Table 3-. All assumptions for the economic input values (e.g. investment costs, O&M costs, energy prices) are shown in Table 3-3. Flow charts for the energy inputs and outputs of the system are shown in Figure 3-183 and Figure 3-194.

Table 3-8: Technical input values to the TRNSYS model. The system is scaled for a biogas consumption of 2500 Nm<sup>3</sup>/h.

Parameter	Value	Unit
Gas flows		
Biogas flow (60% CH <sub>4</sub> , 40% CO <sub>2</sub> )	2500	Nm <sup>3</sup> /h
Hydrogen production at full load	4010	Nm <sup>3</sup> /h
Sabatier production at full load	2570	Nm <sup>3</sup> /h
Heat of combustion (LHV)		
Biogas (60% CH <sub>4</sub> , 40% CO <sub>2</sub> )	21.37	MJ/Nm <sup>3</sup>
Hydrogen	10.78	MJ/Nm <sup>3</sup>
SNG	34.71	MJ/Nm <sup>3</sup>
Electrolyzer		
Electrolyzer capacity	12.32	MW electricity in
Electrolyzer hydrogen output	12.00	MW H <sub>2</sub> out
Electrolyzer steam consumption	2.07	MW heat in
Electrolyzer HT heat boost (for steam from Sabatier)	0.211	MW electricity in
Electrolyzer useful heat out (<100°C)	0.652	MW heat out
Methanation		
Methanation capacity	24.77	MW SNG out
Methanation biogas input	14.84	MW biogas in
Methanation electricity input (compressors)	0.418	MW electricity in
Methanation steam out	2.07	MW heat out
Methanation low-temperature heat out (<100°C)	1.05	MW heat out
Biogas motor		
Biogas motor capacity	2500	Nm <sup>3</sup> /h biogas in
Biogas motor, electrical efficiency	42	%
Biogas motor, thermal efficiency	51	%
Solid oxide fuel cell		
Solid oxide fuel cell capacity	2500	Nm <sup>3</sup> /h biogas in
Amine upgrading		
Amine upgrading capacity	2500	Nm <sup>3</sup> /h biogas in
Amine upgrading electricity input at full load	0.475	MW electricity in
Amine upgrading output	1500	Nm <sup>3</sup> /h SNG out

Table 3-3: Economic input values to the TRNSYS model, for a system scaled for a biogas consumption of 2500 Nm<sup>3</sup>/h. For the biogas motor, the fixed and variable O&M costs should be summed.

Parameter	Value	Unit
Energy prices		
Biogas purchase price	3.25	DKK/Nm <sup>3</sup>
SNG sales price	6.00	DKK/Nm <sup>3</sup>
Heat sales price	200	DKK/MWh
Oxygen sales price	0	DKK/Nm <sup>3</sup>
Electricity purchase price	2015 price variation	DKK/MWh
Economic parameters		

Price level	Fixed 2017 prices	
Real interest rate	4	%
Investment lifetime, amine upgrading	15	Years
Investment lifetime, all other components	10	Years
Investment subsidy	0	% of investment
Max. no. of operation hours per year	8000	hours
€ to DKK conversion rate	7.45	DKK/€
Investments		
Solid oxide electrolyzer	3.22	Million DKK/MW
Sabatier (incl. biogas & SNG compressors)	5.52	Million DKK/MW
Amine upgrade	4.26	Million DKK/MW
Biogas motor	1.00	Million DKK/MW
Solid oxide fuel cell	11.20	Million DKK/MW
Flaring	0.00	Million DKK/MW
Biogas storage	19255	DKK/MWh
Hydrogen storage	48470	DKK/MWh
Operation & maintenance costs		
Solid oxide electrolyzer	3.00	% of investment
Sabatier	3.00	% of investment
Amine upgrade	1.92	% of investment
Biogas motor, fixed O&M	1.00	% of investment
Biogas motor, variable O&M	59.60	DKK/MWh
Solid oxide fuel cell	3.00	% of investment
Flaring	0.00	% of investment
Biogas storage	8.00	% of investment
Hydrogen storage	2.00	% of investment

The model excludes all taxes and tariffs regarding electricity purchase and sales, CO<sub>2</sub> emissions and the plant operation. The annualized capital expenses are calculated using the annuity payment formula:

$$CAPEX_{annualized} = \frac{r \cdot CAPEX}{1 - (1 + r)^{-n}} \quad (\text{Eq. 12})$$

Here *CAPEX* denotes the total investment cost, *r* is the interest rate and *n* is the investment lifetime in years.

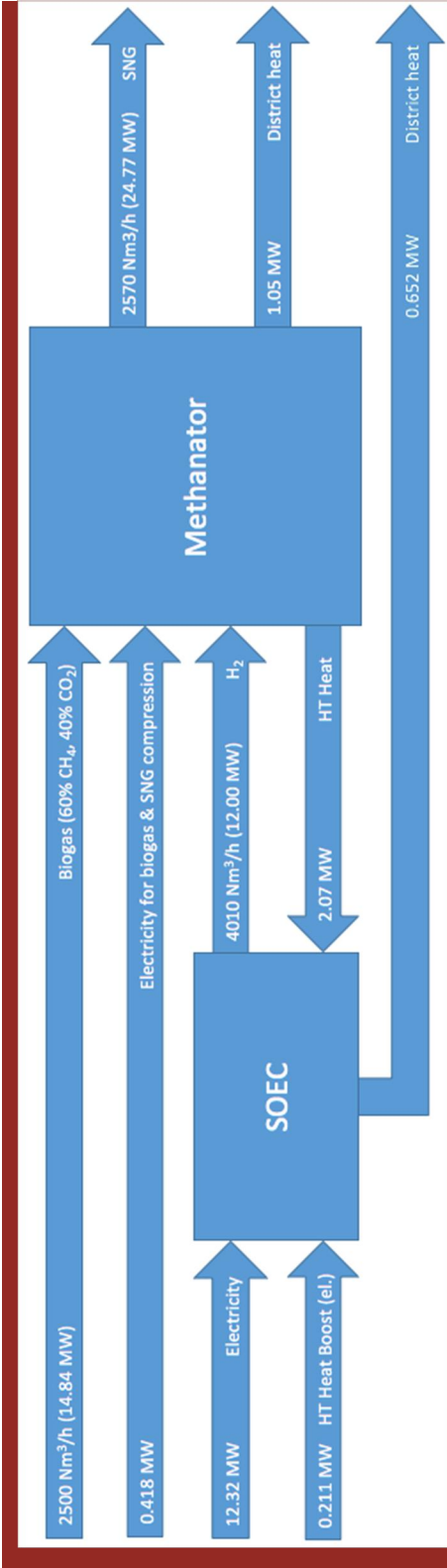


Figure 3-183: The total energy flows in the modelled system, shown in terms of gas flows (Nm<sup>3</sup>/h) and power (MW). The values here assume a 100% scaling of the SOEC and methanation units.

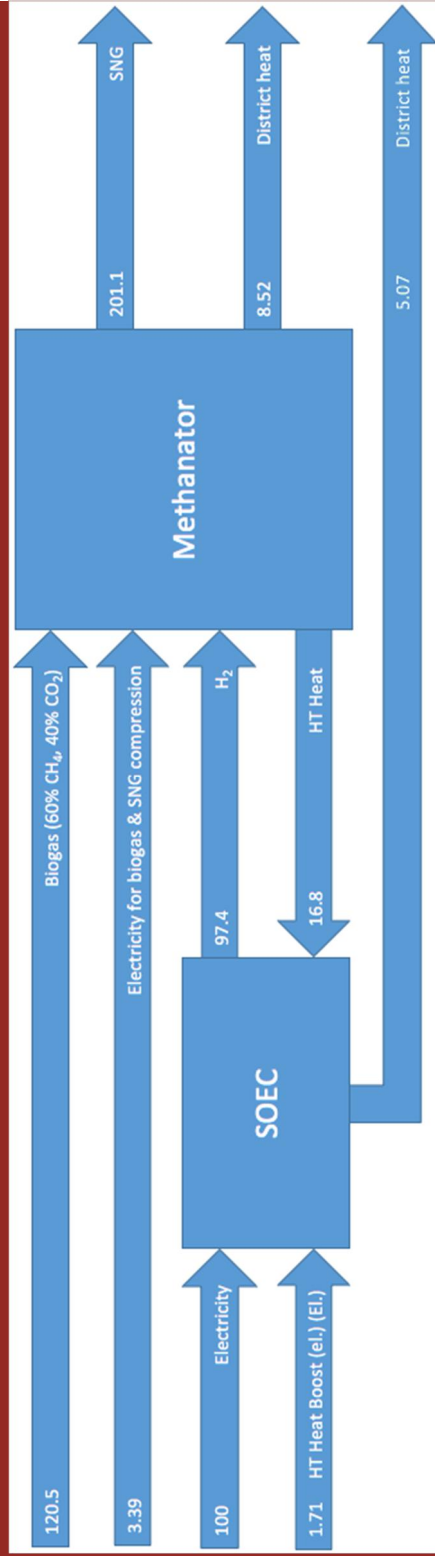


Figure 3-19 The total energy flows in the modelled system, shown in relative units where the electricity input for the electrolyzer has been set to 100 units.

Concerning the electricity prices, the following methodology was used:

The electricity price time series for the year 2015 was used. The average electricity spot price in the year 2015 was 170 DKK/MWh. As already mentioned, the electricity price used in this analysis is excluding all tariffs and taxes.

The analysis was performed for four different electricity price levels; an average spot price of 170, 250, 350 and 450 DKK/MWh. Time series for the different variations were constructed by scaling the 2015 time-series with a multiplication factor. To illustrate this, Figure 3-205 shows the electricity price time series for two of these cases; the non-scaled 2015 spot price time series (which has an average price of 170 DKK/MWh) and the same time series scaled to an annual average price of 450 DKK/MWh.

It should be noted that by scaling the 2015 electricity spot price with a multiplication factor (rather than scaling the price by the addition of a constant), the absolute price fluctuations also increase compared to the original 2015 profile.

Figure 3-216 shows duration curves for the electricity price, both for the non-scaled 2015 prices and for the scaling to an average annual price of 350 DKK/MWh. In the current project, Ea Energianalyse has investigated the socio-economic feasibility of biogas methanation with SOEC electrolysis and made projections for the years 2030 and 2050. In Figure 3-216, the simulated electricity price profile calculated for the year 2030 by Ea Energianalyse is shown for comparison with the electricity price profiles used in the current private-economic analysis. The annual average price of Ea Energianalyse's 2030 electricity price is 347 DKK/MWh. As can be seen in the figure, the duration curve for the 2015 prices scaled (using a multiplication factor) to an average price of 350 DKK/MWh is very similar to the calculated 2030 prices by Ea Energianalyse.

In all scenarios, a certain cut-off electricity price was set for the operation of the SOEC. In hours with the electricity price exceeding the cut-off price, the SOEC does not operate. This was one of the parameters included in the optimization of the system.

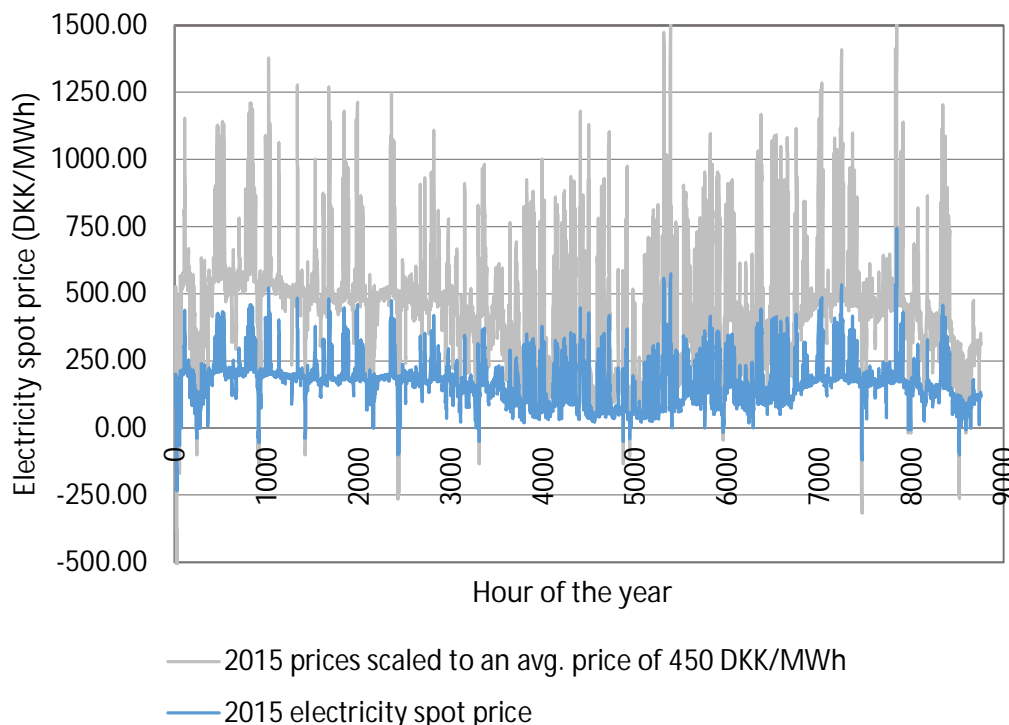


Figure 3-205 Two of the electricity time series used in the analysis: The 2015 electricity spot price time series, with an average price of 170 DKK/MWh and a scaled version of the same time series with an average price of 450 DKK/MWh.

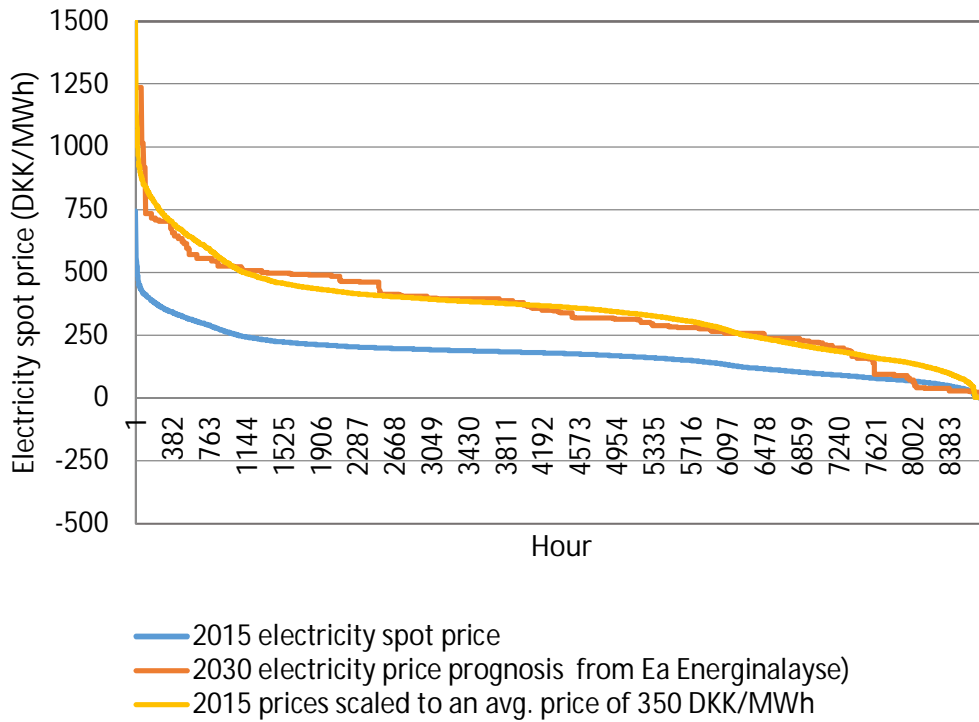


Figure 3-216: Duration curves for the electricity spot price in 2015, the scaling to an average annual price of 350 DKK/MWh and the 2030 price projection calculated by Ea Energianalyse. The duration curves with an average price of 350 DKK/MWh is very similar to the 2030 projection.

The TRNSYS model was optimized using the optimization software GenOpt. The objective of the optimization algorithm was in all cases to maximize the annual profit from the system's operation.

The optimization was performed for the following model variations:

- Four different choices of back-up option for biogas consumption (options A-D)
- Four different average electricity price level (170, 250, 350 and 450 DKK/MWh),

Furthermore, for each model variation, the optimization was performed with two different settings for the scaling of the electrolyzer and methanation capacity:

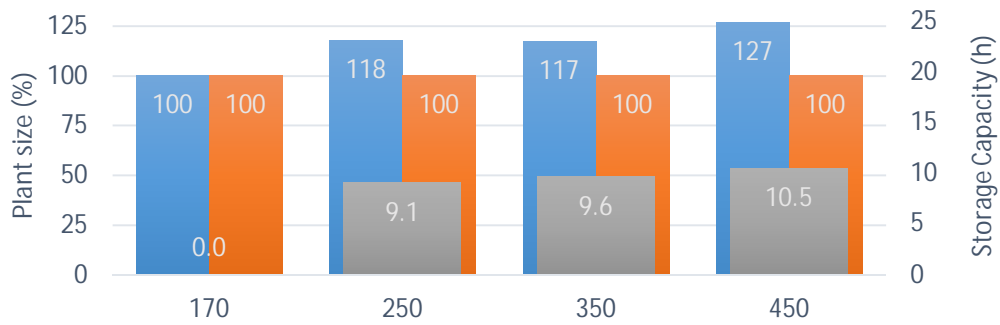
- A fixed 100% capacity of the electrolyzer and the methanator.
- A variable capacity of the electrolyzer and the methanator. The scaling of the electrolyzer and the methanator were included as optimization parameters.

These variations lead to a total of 32 scenarios. In 16 of these scenarios, the electrolyzer and methanator scaling were included as optimization parameters, while in the remaining 16 scenarios the electrolyzer and methanator scaling was fixed at 100%.

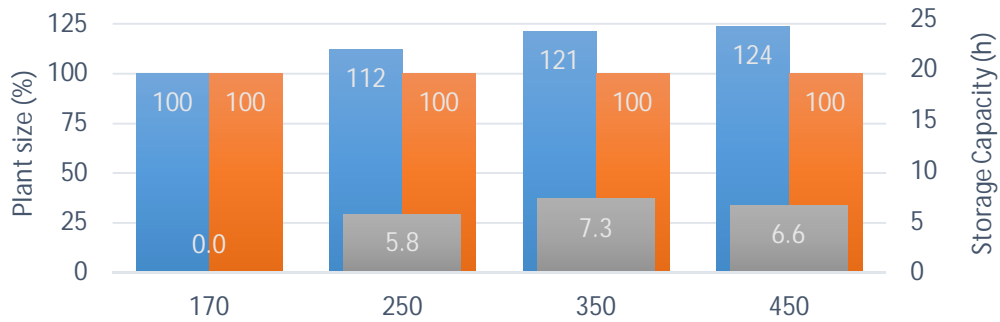
The results of the optimization are shown on Figure 3-227.

The optimal electrolysis scaling was found to equal 100% in all scenarios with an average electricity price of 170 DKK/MWh. The optimal hydrogen storage capacity was found to equal zero for this electricity price level. For scenarios with an average electricity price of 250 DKK/MWh or higher, the optimal electrolyzer scaling was found to be between 110% and 127%, and the hydrogen storage capacity was found to be between 5.8 and 10.5 hours of full production. The optimal biogas storage capacity was found to equal zero in all analyses.

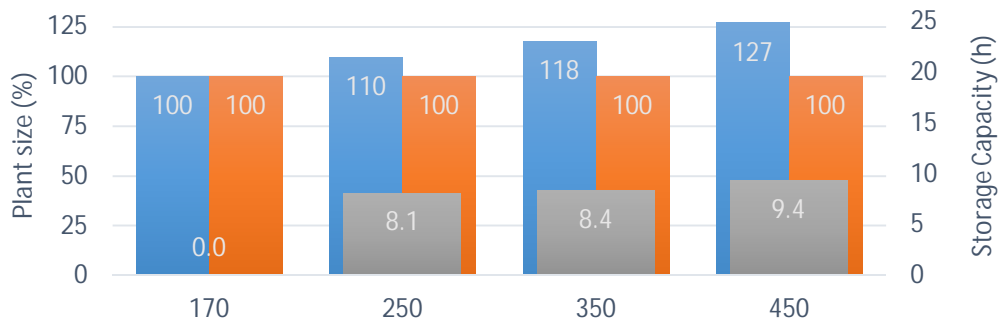
a) Flaring



b) Amine upgrading



c) Biogas motor



d) Solid oxide fuel cell

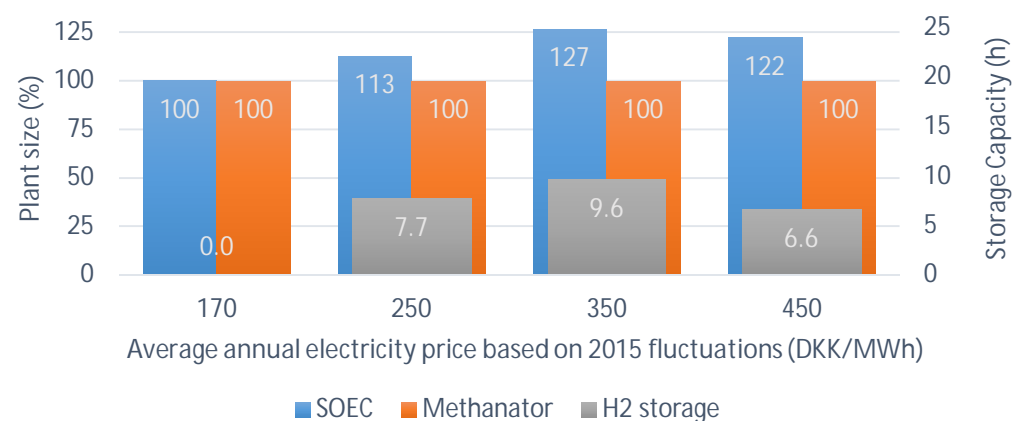


Figure 3-227 The results of plant size optimization. The scaling of the electrolyzer and the methanator were optimized, as well as the storage capacity for hydrogen and biogas. The optimization of the biogas storage size yielded zero in all cases.

The electricity cut-off price for the SOEC, i.e. the price above which the electrolysis does not operate, was one of the optimization parameters. The results for the cut-off price are shown on Fig. 3-28. For the scenarios with the lowest electricity price, no hydrogen storage is utilized and the cut-off level is rather high. For the scenarios with average electricity prices of 250 DKK/MWh or higher, the electrolysis is slightly oversized and small hydrogen

storages are included (as shown in Figure 3-227). This adds flexibility to the system operation and lowers the optimal electricity-cut off price in these scenarios.

The maximum annual production is 20.6 million Nm<sup>3</sup> of SNG per year, corresponding to full load when outages and maintenance have been accounted for. There is only a 5% difference between the scenario with the lowest SNG production (19.5 million Nm<sup>3</sup>/year) and the scenarios with the highest production. This shows that even for the highest electricity price, the methanation is operated at least 95% of the time (after accounting for outage and maintenance hours).

The back-up options are, in general, only operated for a very limited number of hours during the year. For the lowest electricity price level, the back-up options are only utilized in times of outages and maintenance. As the electricity price increases, the number of hours per year where it is economical to turn off the electrolysis and methanation process and instead operate the back-up facility slightly increases. The amine-upgrading back-up option has the highest number of operation hours of the four back-up options, with 1133 hours for the scenarios with an average electricity price of 450 DKK/MWh. Approx. 740 of these hours are due to outages and maintenance, but the remaining 400 hours (which corresponds to 5% of the maximum annual operating hours) are due to high electricity prices that lead to high hydrogen production costs.

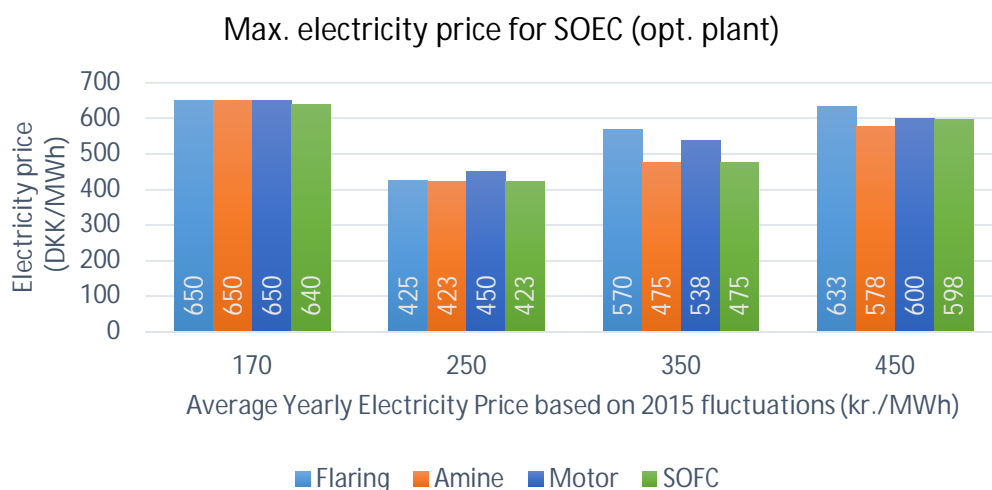


Figure 3-238: The optimal electricity cut-off price for the optimized plant scale, shown for the different electricity prices and biogas back-up options.

The annual profit for each case is shown in Figure 3-249. The profit is calculated as the difference between the income and the expenses.

The scenarios with biogas flaring, biogas motor and SOFC yield a profit for the two lower electricity price levels (170 DKK/MWh and 250 DKK/MWh), but a loss for the two higher price levels (350 and 450 DKK/MWh). In the flaring scenario, the biogas purchased from the biogas plant when the methanation is not running is flared, and thus only contributes to the expenses, and not to the income. The investment costs for the flaring equipment are neglected in the model. In the scenarios with a biogas motor or an SOFC the investments make up a larger part of the investment than in the other scenarios. This additional investment is not fully recovered by the electricity and heat sales due to the relatively few operating hours obtained for the motor and SOFC.

In the scenarios with the amine upgrading back-up option, the sales price of the amine-upgraded SNG is assumed to be the same as that of the methanised SNG (6.00 DKK/Nm<sup>3</sup>). The investment in the methanation facility is at least partially recovered by the sales of the upgraded gas, and the amine back-up option is more economical than using flaring, a biogas motor or an SOFC as a back-up option for consuming the biogas. The scenarios with the



amine upgrading back-up option yield the highest profits out of all the back-up options, and yields a profit for an average electricity price of 350 DKK/MWh or lower.

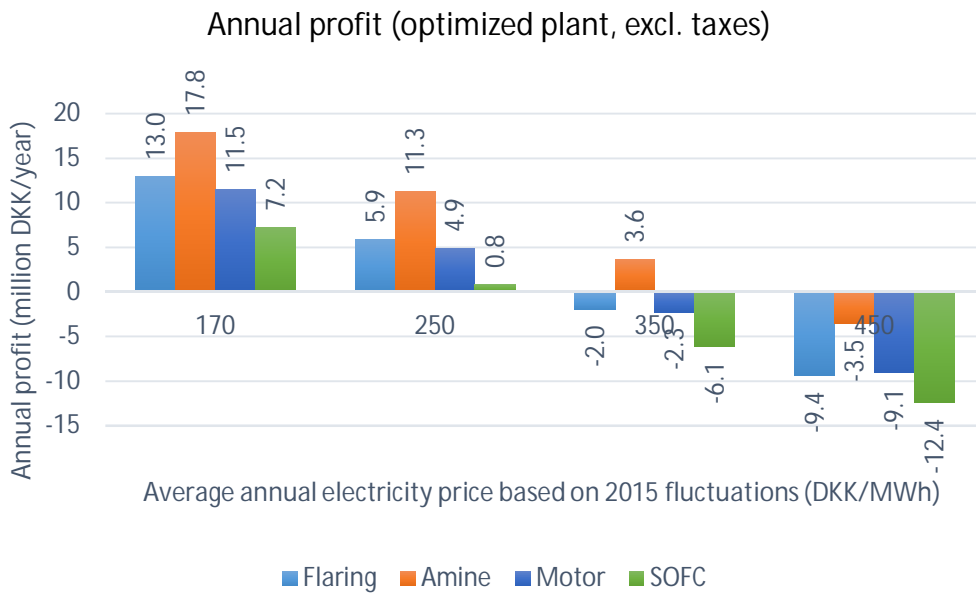


Figure 3-249: The annual profit (or loss, in case of negative values) from the operation of the system.

The average production cost per cubic meter of SNG in the model is shown in Figure 3-3030. In the case of flaring, the production cost includes the purchase cost of the flared biogas. In the case of amine upgrading, the production cost is the weighted average of the production costs from the methanation and from the amine upgrading. In the case of the motor and SOFC, the calculation of the production cost includes the income from the sales of heat and electricity.

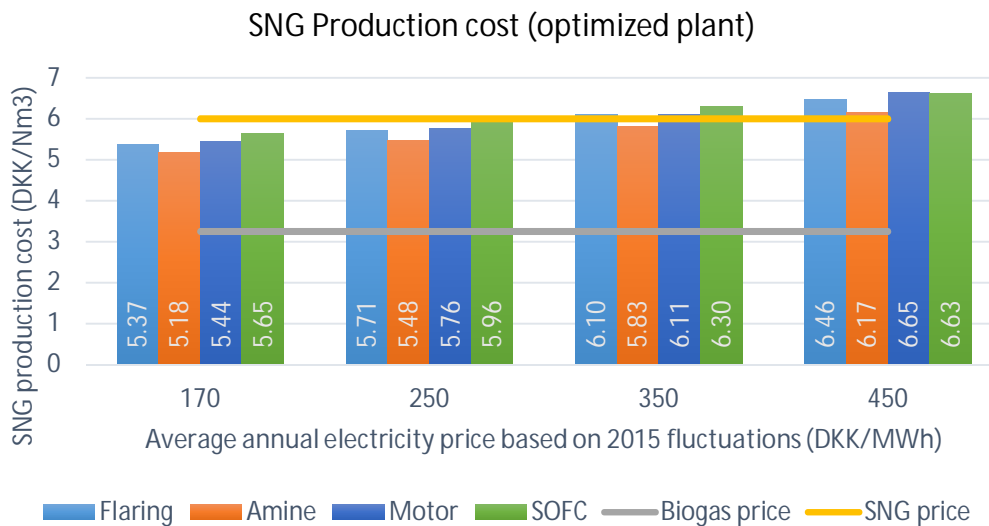


Figure 3-30: The average production costs for the SNG in the model. The gray horizontal line denotes the purchase price of biogas in the model and the yellow horizontal line denotes the selling price of the SNG in the model. The production costs must be lower than the SNG selling price (6.00 DKK/Nm3) for the production to be profitable.

A 100% electrolyzer scaling and no hydrogen storage were found to yield the highest profit for the lowest electricity price level (170 DKK/MWh), regardless of the back-up option. For this electricity price, the optimal electrolysis cut-off price was found to be 650 DKK/MWh. The unscaled 2015 electricity spot price profile only exceeds this price once throughout the

year. This means that in these scenarios the back-up options almost only operate when the electrolysis and methanation plant has outages or is under maintenance. For the higher electricity price levels (250-450 DKK/MWh), some over-dimensioning of the electrolysis (110-127% scaling relative to the methanation) and the introduction of small hydrogen storages (approx. 6-11 hours' production) is found to yield a higher profit than a 100% scaling with no storage. The optimal biogas storage capacity was in all cases found to equal zero.

In the cases with over-dimensioned electrolysis and where hydrogen storages are present, the system has greater flexibility and can produce SNG even when the electricity price is above the cut-off price, as long as hydrogen can be extracted from the storage. As a consequence of this flexibility, the methanation process runs most of the time (7600 hours/year or more in all scenarios), meaning that the back-up facilities obtain rather few operating hours (1160 or less operating hours/year). In all scenarios, operation and maintenance accounts for 735 hours/year where the electrolysis and methanation are not running, and where the back-up facilities must consume the biogas (regardless if the back-up operation during these hours is economical or not). The back-up facilities therefore only obtain 400 hours/year or less during hours in which the model is actually free to choose between the methanation process or the back-up process. This corresponds to only 5% of the full annual operating hours of the plant (8000 hours/year, when outage and maintenance hours have been subtracted).

The main reason for the few operating hours obtained by the back-up units for consumption of biogas is that it is economically difficult to compete with the SNG sales price of 6.00 DKK/Nm<sup>3</sup>. To put the SNG price into perspective with the electricity prices, 6.00 DKK/Nm<sup>3</sup> SNG equals 594 DKK/MWh SNG. The electricity produced by the motor or the SOFC can only very rarely be sold at a price above 594 DKK/MWh, and the heat produced by the motor or the SOFC is assumed to have a sales price of 200 DKK/MWh at all times. The efficiency of the motor and the SOFC is furthermore lower than that of the methanation reactor, meaning that they yield less useful energy output for each unit of purchased biogas.

The amine upgrading back-up facility has the best economy of the four back-up options for biogas consumption. Despite the few annual operating hours of the back-up facilities, it is still more profitable to invest in an amine upgrading facility than to simply flare off the gas. This is because in the model, the biogas must be purchased at 3.25 DKK/Nm<sup>3</sup> during all hours of the year. The flaring is therefore not free of cost, and in the model flaring the gas is less economical than investing in an amine upgrading facility that runs 700-1100 hours/year and sells the upgraded gas at 6.00 DKK/Nm<sup>3</sup>.

The results of this study strongly indicate that methanation of biogas using hydrogen from SOEC can be economical if some conditions are fulfilled:

- The electricity price to the SOEC is no higher than 250-350 DKK/MWh on an annual average basis (the exact value depends on the hourly price distribution and on the choice of a back-up option for biogas consumption).
- The produced SNG can be sold at a price of 6.00 DKK/Nm<sup>3</sup> or higher.
- The steam output from the methanation is utilized as an input for the SOEC process.

The investment costs, O&M costs and durability of the SOEC follow the anticipated development (reflected in the assumptions used in this analysis) in the next 5-10 years.

### 3.9 Dissemination of results

The national as well as international interest in the project concept is considerable resulting in many requests for presentations. A number of lecturers have accordingly been given:

- IEA Working Group on Small Hydrogen Plants and Electrolyzers, Hartford, Connecticut, September 26-27, 2013
- Electrochemical Society, SOFC XIII Conference, Okinawa, Japan, October 10, 2013
- Key Note lecture at ECS Fall Meeting, San Francisco, October 30, 2013
- DTU/DI Energilagrings Konference, Lyngby, November 14, 2013
- Invited Key Lecture, Dutch Sustainable Hydrogen Society, Holland, November 27, 2013 including interview for Dutch chemistry trade journal
- Invited Lecture at Research Coordination Network Carbon Capture and Utilization, Columbia University, New York, April 2, 2014
- Invited Presentation: "2<sup>nd</sup> Workshop Methanation and 2<sup>nd</sup> generation Fuels", Nürnberg, June 12, 2014
- Invited Presentation at ASME Turbo Conference, Düsseldorf, June 16, 2014
- Faraday Discussions, York, July 12-15, 2014
- IEA Biomass Conversion Conference, Berlin, October 27-30, 2015
- IEA Annex 33, Hydrogen Implementing Agreement, Paris, March 9-11, 2016
- 3<sup>rd</sup> Nürnberg Methanation Symposium, Nürnberg, April 8-9, 2016
- EUDP Board Meeting Presentation, Foulum, August, 2016
- Gordon Conference on Green Chemistry, Stowe/Flake, VT, August 1-5, 2016
- World Green Energy Forum, Gyeongju, S. Korea, November 8-10, 2016
- Biogasbranchens Temadag, København, March 7, 2017
- International Methanol Conference, Taastrup, May 10, 2017
- ECS Springs Meeting, New Orleans, LA, May 30, 2017

In connection with the Scandinavian Biogasconference: Biogas2020 the complete project results were presented at an afternoon session on November 9, 2017.

The following articles and papers have been published:

- "Elopgradering af biogas startet med succes", Gasenergi nr. 5, 2016
- "SOEC Enabled Biogas Upgrading", Proceedings of the 12th European SOFC & SOE Forum, Luzern, Switzerland, 2016
- "Vindmøllemetan" kan konkurrere med biometan", FIB nr. 59, Marts 2013
- "Operating experiences with a 50 kW SOEC unit integrated with a catalytic methanation unit for biogas upgrading", Proceedings of EFC2017, December 12-15, 2017, Naples, Italy

and are attached as Annex 1-4

In addition an invited visit has been paid to the Audi biogas + alkaline electrolysis + methanation plant in Wertle, Germany, April 2, 2014

### 3.10 Utilization of project results

The results obtained in the project is already being used by Haldor Topsøe A/S to offer complete desulfurization and methanation plants for biogas upgrading on normal commercial terms including guarantees. This has been made possible by the demonstration of the feasibility as well as durability of the developed design.

At the start of the project the SOEC technology from Haldor Topsøe A/S had only been tested at a single stack level for hydrogen production in laboratory settings so the experience gained in the project has provided essential data to increase the confidence in the technology. Nevertheless, the SOEC stacks and cores need some further developments and demonstrations are deemed necessary in order to bring the technology readiness level from the present level 6 to TRL 8. The results obtained in this project have been used to refine operating and preventive maintenance procedures, which will be implemented in a new EUDP project.

The results from the present project have, however, confirmed the assumptions for the design of full scale, complete plants with respect to consumption and investment parameters. The experience has also further underlined the benefits which could be obtained if pressurized SOEC were developed.

The economic analyses have demonstrated that political initiatives are needed with respect to incentive schemes in order to make the business case attractive for private investors.

The business model of Haldor Topsøe A/S for the roll out of SOEC based technology have been adjusted accordingly so that market implementation is foreseen to start in 2025 and large scale implementation from 2035. The SOEC based eCOS plants for carbon monoxide ion demand production has, however, already been commercialized.

The present and coming demonstration projects are obviously a very important part of the market preparation, not only for biogas upgrading but also for hydrogen, methanol/DME, ammonia and other SOEC based technologies. This should lead to a market uptake from 2020 – 2025 provided the changes in political regulatory measures are implemented.

Clear environmental and energy supply security benefits have, however been identified and the technology could be a key enabler for realizing the ambitious Danish energy policy objectives.

## 4. Project conclusion and perspective

From above results it can be concluded that:

1. SOEC electrolysis can produce hydrogen from steam electrolysis at very high efficiency approaching 98 %
2. The stability of the stacks is promising, but longer tests are needed to confirm multi-year stability
3. Stack degradation can be countered by a gradual increase in operating temperature with negligible impact on efficiency
4. Desulfurization and CO<sub>2</sub> methanation technology is ready for commercial deployment
5. Both the SOEC and methanation units can cope with rapid transients
6. Methanation of biogas using hydrogen from SOEC can be economical if some conditions are fulfilled:
  - a. The electricity price to the SOEC is no higher than 250-350 DKK/MWh on an annual average basis (the exact value depends on the hourly price distribution and on the choice of a back-up option for biogas consumption).
  - b. The produced SNG can be sold at a price of 6.00 DKK/Nm<sup>3</sup> or higher.
7. In an energy system dominated by renewable electricity production and high CO<sub>2</sub> price el upgraded biogas via SOEC can outcompete natural gas, simple upgraded biogas or SNG from thermal gasification
8. The socio economic gain in Denmark would be 350 mio kr. per year or 40 kr./GJ per GJ extra methane produced
9. The amount of renewable methane from biogas production is increased by approximately 30 % by using el upgrading
10. The SOEC unit should operate at full load for 4600 – 4800 hours per year where the average weighted electricity price is 109 kr./MWh

The above scenarios will not be realized in the near future but could be in future sustainable energy scenarios.

The clear environmental and energy supply security benefits have, however been identified and the technology could be a key enabler for realizing the ambitious Danish energy policy objectives.

#### 4.1.1.1 References

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## 4.2 Appendix

### 4.2.1 Biogas quality at AU Foulum

Nitrogen and oxygen contamination in raw biogas at Foulum biogas facility measured using GC at Dansk Gasteknisk Center.

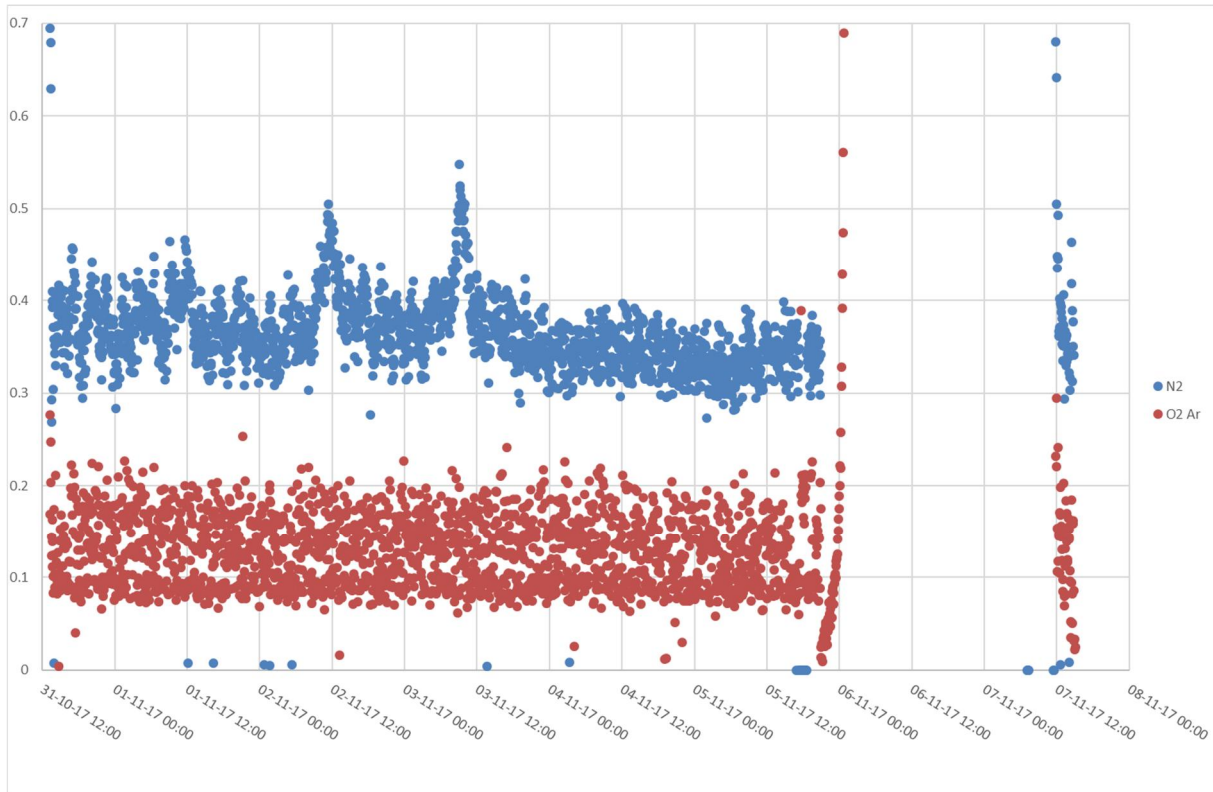


Figure 4-1: Continuous analysis of nitrogen and oxygen in raw biogas. Due to a plant trip the sampling line was flushed with air on the 6<sup>th</sup> of november. Results show a low and stable contamination of O<sub>2</sub> (red) and N<sub>2</sub> (blue) in the raw biogas. Data and analysis performed by DGC.