

# The potential of “Power to Gas” technology integrated with biomethane production

M. Sarić  
J.W. Dijkstra  
S. Walspurger  
W.G. Haije

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## THE POTENTIAL OF “POWER TO GAS” TECHNOLOGY INTEGRATED WITH BIOMETHANE PRODUCTION

M.Sarić<sup>1</sup>, J.W. Dijkstra<sup>1</sup>, S. Walspurger<sup>1,2</sup>, W.G. Haije<sup>1</sup>

<sup>1</sup> ECN, Energy research Centre of the Netherlands,  
Westerduinweg 3, 1755LE Petten, the Netherlands  
(\*M. Sarić : [saric@ecn.nl](mailto:saric@ecn.nl))

<sup>2</sup> Currently employed at Technip Benelux B.V.  
Boerhaavelaan 31, 2713 HA Zoetermeer, the Netherlands

### ABSTRACT

A feasibility study on integration of “Power to Gas” technology with bio-methane production from bio-syngas produced by biomass gasification shows that a significant amount of excess electricity can be accommodated in bio-SNG production. By adding hydrogen produced from intermittent renewable sources to the methanation section, production capacity of methane can be doubled in a flexible operation scheme, while gas quality and cold gas efficiency are hardly affected. Considering two possible limits for H<sub>2</sub> allowed in the natural gas product, it was found that a limit of 10 % v/v can be easily reached. However, a much stricter limit 0.5 % v/v also considered by some pipeline operators cannot be reached with the conventional reactor studied. Simulation and experimental results show however, that with an advanced reactor using sorption enhanced (SE)-methanation the stricter H<sub>2</sub> limit can also be reached. The corresponding CO and CO<sub>2</sub> conversions were close to 100%.

The methanation and SNG upgrade section mode need to deal with a considerable variation in mass flows and variations in the heat balance when switching between hydrogen addition and CO<sub>2</sub> removal mode.

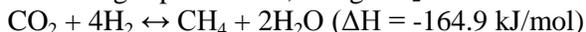
**KEYWORDS:** Power to Gas, Methanation, Bio-methane, Sorption enhanced reaction

### 1 INTRODUCTION

The share of electricity from renewable sources in the European electricity mix is rapidly increasing. As the power generation from wind and solar fluctuates, the match between (renewable) power supply and demand is becoming more challenging. At the same time, there are additional challenges to transmit these increasing volumes of renewable power from wind or solar farms to the end users. Advantageously, the gas infrastructure could accommodate large volumes of electricity converted into gas in case the supply of renewable power is larger than the power grid capacity or than the electricity demand.

The present study focuses on identifying technological opportunities for coupling the power to gas concept to the production of substitute natural gas (SNG) from biomass gasification. The power to gas alternative may materialize when considering that future energy infrastructure with large installed capacity of renewable power production is expected to produce excess electric power during certain periods, power that cannot be accommodated by the grid, leading to grid stability issues.

Using this excess renewable electricity for H<sub>2</sub> production through water electrolysis and subsequent use in producer gas upgrading would offer a method of converting the excess electricity into additional substitute natural gas production, using CO<sub>2</sub> available in the producer gas via the reaction:



During periods when no excess renewable electricity is available, i.e. no extra H<sub>2</sub> can be added; conventional CO<sub>2</sub> removal is applied resulting in a lower amount of substitute natural gas (SNG).

The conversions to SNG rather than direct feed-in of H<sub>2</sub> into the grid potentially circumvents issues with the limited amount of H<sub>2</sub> that is allowed into the natural gas grid. The concept would offer an attractive option for connecting the power grid to the natural gas grid under the condition that operational versatility between H<sub>2</sub> addition (excess power production-*E-excess*) and CO<sub>2</sub> removal (high power demand-*E-demand*) is indeed feasible. Unlike direct H<sub>2</sub> injection in the natural gas grid, the resulting SNG would then fully benefit from existing large scale transport and storage facilities, that have been developed over the years for an efficient and safe distribution of energy to the end consumers, and that are now widely accepted in the society.

In this work, a number of possible design cases and operational modes of the methanation process have been considered regarding 1) the feedstock quality (producer gas), 2) the SNG specifications and 3) the catalysts available for methanation process (sweet methanation catalyst vs. sour methanation catalyst). Steady state system performances for these cases have been calculated using the Aspen Plus modelling program. These results have been evaluated comparing quantitative and qualitative parameters including the cold gas efficiency ( $\eta_{cg}$ ), the SNG gas specification (H<sub>2</sub> content, Wobbe index), the system complexity and the power demand for SNG production. The degree of flexibility of the evaluated systems is discussed with the intention of selecting a preferred configuration. Finally, the coupling of H<sub>2</sub> supply from a water electrolysis plant to a biomass gasification plant is discussed with respect to the fluctuations of the excess electricity available from renewables.

One method of further reduction of the H<sub>2</sub> feed-in into the natural gas grid is an integration of H<sub>2</sub>O separation by solid sorption into the methanation reactor. Selective removal of product H<sub>2</sub>O by a high-temperature sorbent shifts the equilibrium methanation reaction to high conversions and thus low levels of H<sub>2</sub> in the product. In a recent publication (Walspurger, S., et al. 2014) it was experimentally shown that with this sorption enhanced methanation the CO and CO<sub>2</sub> conversions close to 100% can be achieved, being able to produce high grade methane with a very low H<sub>2</sub> content at much lower operating pressures than conventionally used. The benefit of using both sweet and sour sorption enhanced methanation were also evaluated in this work.

## 2 METHODS

To explicitly address only the parts of the system affected by the flexible Power to Gas arrangement only the relevant sections are modelled, i.e. gas cleaning, methanation and compression. Battery limits are producer gas from the gasifier after tar removal, as well as hydrogen from an electrolyzer that is outside the battery limits. The product battery limit is SNG delivered at grid pressure.

Throughout the study, a constant feed gas composition is considered as obtained by the ECN Milena indirect biomass gasifier (van der Meijden, C. M. 2010) with downstream Olga technology (Zwart, R.W.R. 2009) for tar removal. The producer gas capacity entering the system boundaries was assumed to be 200 MW<sub>th</sub>, based on lower heating value (LHV). The operating conditions and the producer gas composition entering the system boundaries are given in Table 1. The producer gas is then upgraded to SNG by removal of NH<sub>3</sub>, sulphur components, CO<sub>2</sub> and water. Both conventional sweet methanation and sour methanation process have been considered in the calculations. In the case of a sweet methanation all

contaminants (H<sub>2</sub>S and others) are thoroughly removed upstream the methanation section. In contrast, for a sour methanation most of the gas cleaning processes will be placed downstream the methanation reactors and only NH<sub>3</sub> and HCl are removed before the methanation.

H<sub>2</sub> produced by the electrolyzer is assumed to be 100% pure, and available at the methanation pressure (the methanation pressure was subject of sensitivity study in this work and has been varied from 20-60 bar) and 300°C.

Table 1 Feed stream composition and conditions after Olga tar removal (van der Meijden, C. M., et al. 2010) and specification limits of G-gas (Donders, R., et al. 2010)

Producer gas composition	[vol%]	G-gas specs	[mol%]
H <sub>2</sub>	21.3	O <sub>2</sub>	< 0.5
CO	25.9	H <sub>2</sub>	Max1 < 0.5, Max2 < 10
CO <sub>2</sub>	12	N <sub>2</sub>	~2
CH <sub>4</sub>	10.3	CO <sub>2</sub>	< 8
C <sub>2</sub> H <sub>6</sub>	0.2	CO	< 0.5
C <sub>2</sub> H <sub>2</sub>	0.3	Aromatic hydrocarbons	1
C <sub>2</sub> H <sub>4</sub>	3.4	BTX* [ppm]	500
C <sub>6</sub> H <sub>6</sub>	0.6	H <sub>2</sub> S [mg/m <sub>n</sub> <sup>3</sup> ]	< 5
C <sub>7</sub> H <sub>8</sub>	0.1	H <sub>2</sub> O [dew point]	< -10°C
N <sub>2</sub>	0.9		
H <sub>2</sub> O	24.8		
	[ppmV]		
H <sub>2</sub> S	314		
COS	35		
NH <sub>3</sub>	2308		
<b>Gas conditions</b>		<b>Gas conditions</b>	
T [°C]	180	Pressure [bar]	60
Pressure [bar]	1	Wobbe Index [MJ/m <sub>n</sub> <sup>3</sup> ]	> 43.46

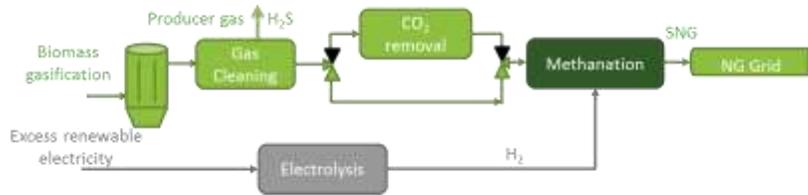
\*BTX = benzene, toluene and xylene

Water for production of steam (that is required to prevent carbon formation, see discussion below) is available at 1 bar and at 20°C. For this study SNG specifications (Table 1) were based on the Dutch G-Gas specification (Groningen gas). The most stringent limit (max1) implies that H<sub>2</sub> must be below 0.5 mol% corresponding to the current specification for the gas infrastructure in the Netherlands. However, in other (European) countries, specifications are less strict, and in the future gas specifications might be less tight. Therefore, this study will also consider the less strict limit, (max2) of 10 mol% H<sub>2</sub> content. The Wobbe index is based on the lower value for the Dutch G-Gas.

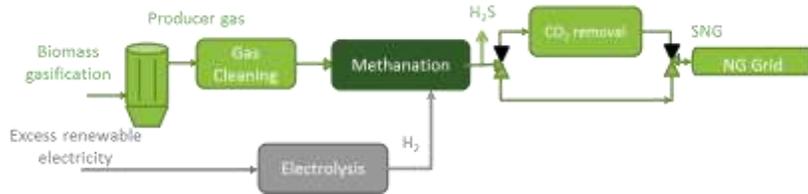
It was assumed that steam export is at 10 bar and 230°C. This steam can then possibly be used for biomass pre-treatment, Olga tar removal or for the electrolyzer, but these units are outside the battery limits. When required, steam at a different pressure levels could also be produced but this is not considered.

## 2.1 Systems studied

Figure 1 gives a schematic overview of the layout of a system for a producer gas upgrade to SNG for studied methanation cases. The operating units for either case are listed in Table 2.



a) Sweet methanation



b) Sour methanation

Figure 1 Scheme of a producer gas conversion to SNG

The systems were evaluated for two operating modes:

- *E-excess mode*: the  $H_2$  produced from the excess renewable electricity is added before the methanation unit. The  $CO_2$  removal unit is bypassed for this mode. The amount of hydrogen in the feed is chosen to match the stoichiometry ratio for methane production.
- *E-demand mode*: the  $CO_2$  removal unit is in operation, and no hydrogen is added. The  $CO_2$  removal (only in the *E-demand mode*) section for the sweet case consists of an amine (MEA) scrubbing unit. The  $CO_2$  is removed in order to reach the stoichiometry ratio for methanation ( $(H_2 - CO_2)/(CO + CO_2)$ ) of 3. For the sour case the Sulfinol process is used. This process was selected because it can deal with large variations in the acid gas concentrations (Khol, A. and Nielsen, R. 1997), that are expected when switching between *E-demand* and *E-excess* mode in the sour case.

The configuration with 3 methanation reactors in series with the recycle over 2<sup>nd</sup> reactor was evaluated, since it was shown (Saric, M., et al. 2013) that it has the highest potential in terms of the steam export. The catalyst maximum temperature was 650°C, using a high temperature Lurgi catalyst (Herrlet, U. 2012). The inlet reactor temperatures of reactors 1, 2 and 3 were 300, 250 and 220°C, respectively.

The methanation pressure was varied between 20 and 60 bar. At the inlet of the methanation section, the conditions are such that thermodynamically it is possible that carbon can be formed. The addition of steam to the methanation reactor feed may be required to limit carbon formation and to ensure sufficient catalyst lifetime. The steam-to-carbon ratio of 0.25 is required to prevent carbon formation in sweet methanation reactors which was obtained by using an isothermal equilibrium Gibbs reactor model (Aspen Plus).

In the case of sour methanation the methanation catalyst hydrogenates unsaturated hydrocarbons, COS and mercaptans and reforms aromatic and higher hydrocarbons. As opposed to the sweet methanation catalyst, that requires an exact stoichiometric ratio between  $CO/CO_2$  and  $H_2$  for methanation, the sour methanation catalyst does not have such a restriction. Experimental data on industrial sour methanation catalyst (Lee, A. L. 1978) showed that catalyst was not losing activity even when non-stoichiometric ratio of  $H_2/CO$  was used.

Table 2 Operation units sweet and sour methanation

Sweet methanation		Sour methanation	
<i>Producer gas cleaning units</i>			
Water Scrubber	NH <sub>3</sub> and HCl at 1 bara and 20°C	Water Scrubber	NH <sub>3</sub> and HCl at 1 bara and 20°C
Compression	To 7 bar with upstream water removal by knock-out		
Hydrodesulphurization unit (HDS)	Hydrogenation of organic sulfurs, at 300°C		
H <sub>2</sub> S removal	Adsorption using ZnO at 300°C		
Pre-reforming reactor	To convert benzene and toluene. Inlet T=500°C, S/C =1.33		
CO <sub>2</sub> removal unit (optional, only in <i>E-demand</i> mode)	Amine (MEA)scrubbing Regeneration: steam at 130°C, energy duty of 3 GJ of steam/ton CO <sub>2</sub> removed (Häring, H-W. 2007)		
Compression to methanation pressure	Sensitivity study for pressures between 20-60	Compression to methanation pressure	Sensitivity study for pressures between 20-60
H <sub>2</sub> addition (optional only in <i>E-excess</i> mode)		H <sub>2</sub> addition (optional only in <i>E-excess</i> mode)	
<i>Methanation section</i>			
3 reactors in series with recycle over the 2 <sup>nd</sup>	Max. T =650°C S/C=0.25	3 reactors in series with recycle over the 2 <sup>nd</sup>	Max. T =650°C S/C=2, <i>E-demand</i> S/C=0.25, <i>E-excess</i>
<i>SNG upgrade units</i>			
		H <sub>2</sub> S removal	adsorption using ZnO at 300°C
		CO <sub>2</sub> removal (optional, only in <i>E-demand</i> mode)	Sulfinol process, CO <sub>2</sub> removed to 6.60 mol% of CO <sub>2</sub> in the SNG Regeneration: steam at 130°C, 3.6 GJ steam/t CO <sub>2</sub> removed.
Drying of SNG	Water knock-out and glycol, modelled as a black-box without energy use	Drying of SNG	Water knock-out and glycol, modelled as a black-box without energy use
SNG compression	Compression to 60 bar	SNG compression	Compression to 60 bar

In the *E-excess* mode of the sour methanation case it was assumed that H<sub>2</sub> is added to obtain the required stoichiometric ratio for methanation. In the *E-demand* mode of sour methanation the non-stoichiometric ratio is supplied to the first methanation reactor. This is because the CO<sub>2</sub> removal unit is placed after the methanation reactors (Table 2) in a sour methanation configuration, and thus there is not enough H<sub>2</sub> in the producer gas entering the first methanation reactor for obtaining the required stoichiometric ratio. The tendency for carbon formation in the methanation reactors will largely vary for

sour *E-excess* and *E-demand mode* since molar ratio of H/C and O/C atoms for these cases will differ substantially. Therefore, it was decided to check in Aspen Plus at which S/C ratio for both cases the thermodynamic carbon formation can be avoided. This was done using an isothermal equilibrium Gibbs reactor model.

For the sorption enhanced methanation system the last reactor is the sorption enhanced reactor. This is because most of the conversion takes place in the first section of the methanation process resulting in a temperature rise in the first and second methanation reactors that approaches the limits that the catalyst may withstand. It is thus reasonable to implement a sorption enhanced process in place of the third reactor that is used mostly to enhance the conversion such that tight product specifications may be matched. In order to simulate this sorption enhanced reactor a 3-step model was constructed. First, a stoichiometric reaction step with quantitative conversion of H<sub>2</sub> to H<sub>2</sub>O was implemented, followed by removal step of a pre-specified amount of H<sub>2</sub>O, and finally a Gibbs minimisation reaction step to calculate the equilibrium composition corresponding to the remaining water. The pre-defined amount of H<sub>2</sub>O removed by the sorbent was set to 90% of the water as present in the feed and produced by full conversion of all H<sub>2</sub> in the feed. Periodically, the sorbent needs to be regenerated by pressure reduction or heating, but this step is not considered in the model.

In this study eight combinations of cases and modes were considered:

- Sweet methanation case in *E-demand* and *E-excess modes*
- Sour methanation case in *E-demand* and *E-excess modes*
- Sorption Enhanced (SE)-sweet and sour methanation cases in *E-demand* and *E-excess mode*

The heat integration was performed using a pinch analysis for the hot and cold streams in the system, assuming a minimum temperature approach of 10°C.

The results were assessed using the following performance indicators:

Energy balance criteria:

- SNG produced
- Cold gas efficiency
- Steam export
- Power use

SNG gas quality criteria:

- H<sub>2</sub> level in SNG (two limit levels considered: *max1* and *max2*), see below
- Wobbe index (minimum value)

The cold gas efficiency ( $\eta_{cg}$ ) of the process was calculated relative to the lower heating value of producer gas and H<sub>2</sub> according to:

$$\eta_{cg} = \frac{(LHV \cdot F)_{product}}{(LHV \cdot F)_{feed}} \quad (1)$$

The Wobbe index is defined as:

$$Wobbe = \frac{HHV}{Rd^{0.5}} \quad (2)$$

Where:

F- volumetric flow [m<sub>n</sub><sup>3</sup>/s], HHV - high heating value of SNG [MJ/m<sub>n</sub><sup>3</sup>], LHV - lower heating value [MJ/ m<sub>n</sub><sup>3</sup>], Rd - relative density of SNG to air

In case of the Wobbe index, only the lower value of 43.5 MJ/m<sub>n</sub><sup>3</sup> for the Dutch G-Gas is used.

For the H<sub>2</sub> content two possible limit values are considered. The most stringent limit (max1) implies that H<sub>2</sub> must be below 0.5mol%, corresponding to the current specification for the gas infrastructure in the Netherlands (Zachariah, J. L., et al. 2007). However, in other (European) countries specifications are less strict, and in the future gas specifications might be less tight. Therefore, this study will also consider the less strict H<sub>2</sub> limit (max2) of 10 mol % H<sub>2</sub> content.

In the simulations, the main design variable which is the operating pressure of the methanation reactor is varied.

### 3 RESULTS AND DISCUSSION

Table 3 summarizes the process simulation results at a methanation operating pressure of 46 bar in terms of the key performance parameters. Given that the unit must switch between *E-demand* and *E-excess*, both cases must be operated at the same pressure, delivering comparable SNG qualities that are expressed in terms of Wobbe index and hydrogen content.

Table 3 shows that by H<sub>2</sub> addition the SNG production capacity can be almost doubled in either methanation case. The SNG production increases from 174 MW in *E-demand* mode to the 341 MW in *E-excess* mode. Effectively 83% of the lower heating value of the imported hydrogen is converted into SNG, while electricity demand is 288 MW. Sweet and sour methanation have similar performance numbers, only in the H<sub>2</sub> mol% in the SNG streams differ to such an extent that measures need to be taken.

Table 3 Comparison between sweet and sour methanation

	Sweet case	Sour case
<b><i>E-excess</i></b> <b>[46 bar]</b>	<i>SNG<sub>produced</sub></i> =341.6 MW Electricity = 288.2 MW Steam export: 50 MW $\eta_{cg}$ =85 % <i>W</i> =50.48 MJ/Nm <sup>3</sup> H <sub>2</sub> =1.64%	<i>SNG<sub>produced</sub></i> =341.77 MW Electricity = 288.7 MW Steam export: 55 .MW $\eta_{cg}$ =85 % <i>W</i> =50.57 MJ/Nm <sup>3</sup> H <sub>2</sub> =1.18%
<b>20-60 bar</b>		
H <sub>2</sub> <0.5%	<i>Never</i>	<i>Never at stoichiometric feed</i>
H <sub>2</sub> <10%	<i>Always</i>	<i>Always</i>
<i>Wobbe</i> > <i>min</i>	<i>Always</i>	<i>Always</i>
<b><i>E-demand</i></b> <b>[46 bar]</b>	<i>SNG<sub>produced</sub></i> =174.1 MW $\eta_{cg}$ =87.0% <i>W</i> =49.73 MJ/Nm <sup>3</sup> H <sub>2</sub> =1.26% Heat demand:-23.30 MW	<i>SNG<sub>produced</sub></i> =174.06 MW $\eta_{cg}$ =87.0% <i>W</i> =44.10 MJ/Nm <sup>3</sup> H <sub>2</sub> =0.58 % Heat demand:-18.63 MW
<b>20-60 bar</b>		
H <sub>2</sub> <0.5%	<i>Never</i>	>50 bar
H <sub>2</sub> <10%	<i>Always</i>	<i>Always</i>
<i>Wobbe</i> > <i>min</i>	<i>Always</i>	<i>Always</i>
<b>Main conclusion</b>	H <sub>2</sub> <0.5% <i>never met</i> H <sub>2</sub> <10% <i>always met</i> <i>Wobbe always met by far</i>	H <sub>2</sub> <0.5% <i>met at p &gt; 50 bar</i> H <sub>2</sub> <10% <i>always met</i> <i>Wobbe always met by far</i>

Figure 2 shows that the sweet methanation case is never capable of meeting the lower H<sub>2</sub> concentration of the two envisaged limits. The sour case may comply with both limits, but only in the case of *E-demand* and for a methanation pressure higher than 50 bar. This is because it was assumed that sour methanation in *E-demand* mode operates under non-stoichiometric conditions as suggested by Lee, A.L. 1997. However, the assumption that sour methanation catalyst can operate under non-stoichiometric conditions should be confirmed experimentally. In the sour *E-excess* mode the stoichiometric conditions are required, and thus the stricter H<sub>2</sub> limit could not be reached. However, by controlling H<sub>2</sub> addition in

this mode this stricter H<sub>2</sub> concentration limit can be also obtained at the expense of the SNG production capacity.

In the recent work of Walspurger, S. 2014, et al. it was experimentally shown that conversions close to 100% can be reached by using of sorption-enhanced methanation with a mix of commercially available adsorbent and catalyst. The application of this technology can be interesting in the studied systems, since it can enable control of the H<sub>2</sub> in the SNG. Figure 3 shows that if sorption enhanced methanation is used, the stricter limit can be easily reached in either mode and for both methanation cases at much milder operating conditions. For the sweet *E-excess mode* already at 30 bar operating pressure the max 2 limit of 0.5 H<sub>2</sub> mol% specification is reached.

Table 3 shows that when switching between two operation modes the main challenge will be the heat management of the conversion chain since it is either steam exporter in the *E-excess* or steam demanding in the *E-demand mode*.

While some opportunities for integration do exist with the biomass gasification plant and possibly with the electrolyzer unit, better heat management may be achieved with novel systems comprising high temperature solid oxide electrolyzer with dual operational modes that are currently under investigation ( Tremel, A. 2014)and could appear of great advantage for feasibility of cost effective power to gas.

In order to use intermittent electricity, all equipment used in the process should be able to cope with significant flow rate and temperature fluctuations, while maintaining SNG quality in the specifications range at all time. For the two modes of sweet methanation the results indicate that there is a factor 3.4 difference in the volume flows to the methanation reactors, 2 to the SNG compressor and 5 in recycle compressor. The turndown ratio between the *E-demand* and *E-excess* modes in the methanation sour case is 2.5 for the methanation reactors and 2 for the SNG compressor. In the sour *E-demand* recycle from the 2<sup>nd</sup> to the 1<sup>st</sup> methanation reactor maintaining the temperature at 650°C is not required. Thus, turndown ratio for the sour methanation section is significantly lower than for the sweet methanation section.

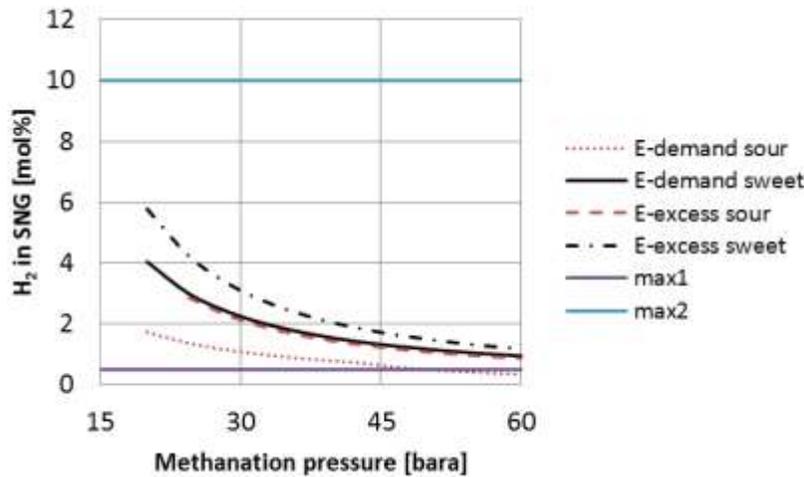


Figure 2 H<sub>2</sub> content in the SNG at different operating methanation pressures for sweet and sour *E-demand* and *E-excess* modes

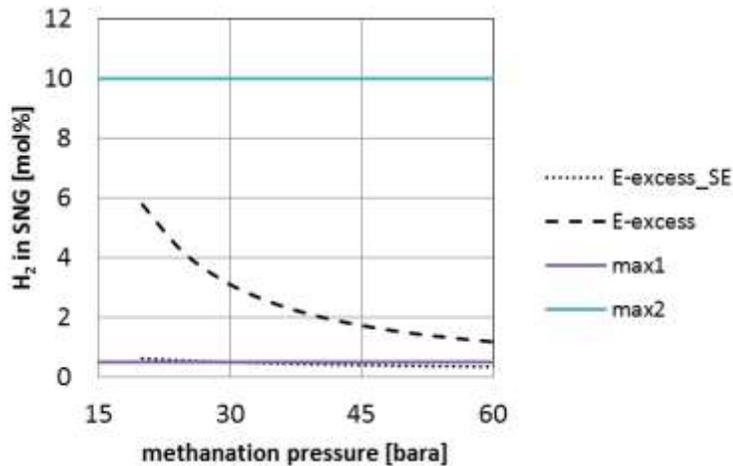


Figure 3 H<sub>2</sub> content in the SNG at different operating methanation pressures in the *E-excess mode* with and without sorption enhanced water removal

Table 4 Sour methanation vs. sweet methanation

Sweet methanation	Sour methanation
H <sub>2</sub> <0.5mol% can never be reached.	H <sub>2</sub> <0.5mol% can be obtained, but only for <i>E-demand</i> pressure >50 bar. For the <i>E-excess</i> mode a study with operation in non-stoichiometric conditions should be performed.
However, with <i>sorption enhanced</i> methanation this is not an issue	The methanation pressure can further be decreased by using <i>sorption enhanced</i> methanation
Minimum Wobbe index always met, but above target, dilution required	+ In <i>E-demand</i> mode the Wobbe index is controlled by the CO <sub>2</sub> removal rate in the acid gas removal unit
	+ simpler design ( less process units) and lower turndown ratio in methanation section
	+ lower steam demand for the <i>E-demand mode</i> (benefit depends on total system heat balance) and higher steam export for the <i>E-excess</i> (benefit depends on whether there is use for this steam)
- H <sub>2</sub> S to be removed lower than 10ppb	+ less deep requirement for H <sub>2</sub> S removal (H <sub>2</sub> S+COS content in SNG < 5mg/Nm <sup>3</sup> )
+ Industrially proven catalyst	+ Indications in literature that sour methanation catalyst does not loose activity in the presence of aromatic hydrocarbons, capability of catalyst to convert hydrocarbons (to be confirmed) Limited information on commercial scale experience
- Stoichiometric ratio required to avoid soot formation	++ Can operate in non-stoichiometric conditions

For the methanation reactors the turndown ration of 3.4 can be reasonably anticipated since it was assumed that reactors are fixed bed reactors. Reactors will be designed for the largest flows. Thus, at lower flows, the gas hourly space velocity in the catalytic bed will be lower, and thermodynamic conversion can be obtained without further issues at the condition that side reaction such as carbon formation or heavier hydrocarbon formation does not become a problem. It is suggested to study the dynamics of switching between the two modes in the reactors, as the local effects of heat production and removal in the reactor might change which could give a risk of temporary hot-spots (van Dijk, H.A.J. et

al. 2010). Variation in flow will have impact on the recycle ratio and SNG compressors efficiency. A likely measure can be to install multiple compressors in parallel. Finally, it will be challenging to design a system for heat integration that can cope with the significant differences in flows, a separate study is to be devoted to this issue.

A comparative assessment of sweet and sour methanation is given in Table 4. It can be concluded that the sour methanation has several benefits over the sweet methanation, as less process steps are required, lower turn down ratio in methanation and a little better performance than the sweet methanation process. The catalytic aspects are however more challenging.

## 4 CONCLUSIONS AND RECOMMENDATIONS

The value of the producer gas in the “Power to Gas” concept assessed by the present study has revealed to be an attractive option for the production of renewable methane and further transport and storage in the existing gas infrastructure. Beyond allowing for large scale storage of fluctuating renewable power, it enables the introduction of renewable energy in the whole energy system, from power production to use in industry and households. The results show that renewable hydrogen addition to a producer gas originating from the gasification of biomass allows for doubling the SNG production compared to using the producer gas alone for SNG production. When intermittent renewable hydrogen supply is considered, the process must be able to cope with a turndown ratio of 3.5 behind the hydrogen addition point to ensure continuous operations. Both conventional sweet methanation and sour methanation processes were considered in the conversion chain from producer gas to SNG with and without renewable hydrogen addition. While no significant differences were found in process efficiency between sour and sweet methanation, the sour methanation process would simplify the conversion chain by combining reforming of higher hydrocarbons and the methanation in a single process unit. Moreover, the results showed that the conventional sweet methanation does not meet the most stringent limit for the H<sub>2</sub> content of the flue gas of 0.5 mol%, in the considered plant configuration while sour methanation could allow for reaching the specification at the operating pressure higher than 50 bar when operating under non-stoichiometric conditions. To overcome these hurdles, sorption-enhanced methanation is a promising technology, since 100% conversion and this stricter H<sub>2</sub> limit at much milder operating conditions can be easily obtained. This really determines an area of further RD&D.

Nevertheless, for both sweet and sour methanation concepts, the system study shows that the main challenge will be the heat integration design since the methanation section is either steam exporter in the electricity excess or steam demanding in the electricity demand mode. While some opportunities for integration do exist with the biomass gasification plant and possibly with the electrolyzer unit, better heat management may be achieved with novel systems comprising high temperature solid oxide electrolyzer with dual operational modes that are currently under investigation and could reveal to be of great advantage for feasibility of cost effective “Power to Gas”.

Challenges for future R&D are in the field of load/demand following reactors and heat exchangers, proof/development of sour methanation catalyst that can operate under non-stoichiometric conditions without losing activity in the presence of aromatic hydrocarbons, and optimisation of the adsorbent/catalyst for the sorption-enhanced methanation.

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**ECN**

Westerduinweg 3  
1755 LE Petten  
The Netherlands

P.O. Box 1  
1755 LG Petten  
The Netherlands

T +31 88 515 4949  
F +31 88 515 8338  
info@ecn.nl  
www.ecn.nl