IEA Bioenergy
Technology Collaboration Programme

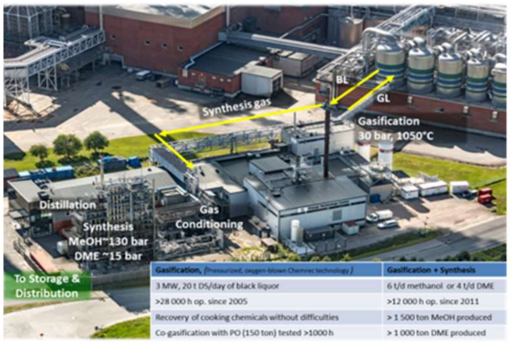
Synergies and Services from H_2 and Biobased Value Chains Deployment

Case studies

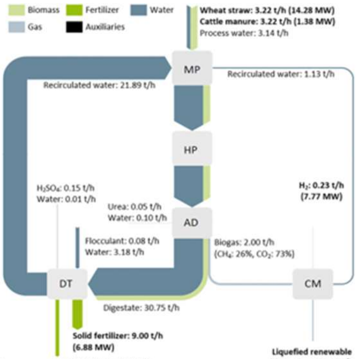
Contribution of IEA Bioenergy Tasks 34, 32, 33, 37, 39, 42 and 44 to Inter-Task Project (ITP) Synergies of green hydrogen and biobased value chains deployment

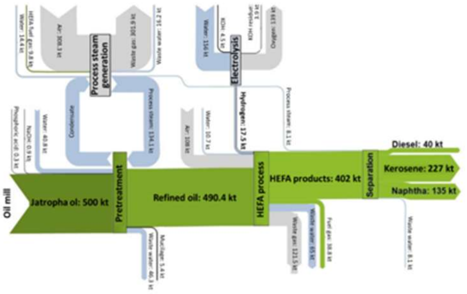
IEA Bioenergy: Task 34

September 2025



Gasification, (Pressurized, oxygen-blown Chemrec technology)	Gasification + Synthesis
3 MW, 20 t DS/day of black liquor	6 t/d methanol or 4 t/d DME
>28 000 h op. since 2005	>12 000 h op. since 2011
Recovery of cooking chemicals without difficulties	> 1 500 ton MeOH produced
Co-gasification with PO (150 ton) tested >1000 h	> 1 000 ton DME produced





Technology Collaboration Programme
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Preface

According to the International Energy Agency (IEA), fuels in the form of hydrogen, hydrogen-based fuels, and bioenergy will have to meet 24% of global final energy demand in 2070 in the Sustainable Development Scenario (SDS), particularly in the areas where direct electrification is difficult [1]. The statistics show that all these fuels need to ramp up quickly to meet the targets. Bioenergy is a limited but a very diverse energy carrier and required for various applications (industry, transport, high-temperature heat, negative carbon emissions etc.). The future scenarios typically see bioenergy in combination with carbon capture and storage/utilization (CCS/U).

The SDS describes that 20% of hydrogen use in 2070 will be in the production of synthetic fuels from hydrogen and CO_2 for the aviation and further 10% used for ammonia production [1]. While most of the hydrogen is produced from natural gas today, the demand for renewable hydrogen is increasing. The main interest has so far been in electrolytic hydrogen from wind and solar electricity [2].

In addition to electrolytic hydrogen, there are also great opportunities to convert biomass to renewable hydrogen, so-called biobased hydrogen or biohydrogen. This option is currently a rather overlooked opportunity for providing renewable hydrogen and there is a need to make information and data available on biohydrogen production and utilization options. Furthermore, there are many biobased processes either in demand for renewable hydrogen (e.g. synthetic renewable fuels, biorefining) or that could benefit from renewable hydrogen integration for improving the quality of products (e.g. boosting biomethane production). In addition to process level synergies between hydrogen and biobased value chains, system level synergies and services are expected to take place, such as increased flexibility, use of joint infrastructure and provision of long-term storage options. Different synergies could benefit the economic deployment of both bioenergy and renewable hydrogen-based fuels, and the overall energy system demands.

Biohydrogen and renewable hydrogen in biobased processes

Biomass-based hydrogen or biohydrogen pathways should be considered as an important complement to water electrolysis as many of the biogenic pathways may provide great benefits such as:

- Non-intermittent, fossil-free, large-scale hydrogen production, i.e. 24/7.
- Mitigation of the demand for fossil-free power.
- Process integration opportunities to reach more energy efficient production systems.
- Co-production of other value-added commodities such as biocarbon, biochar, biomethane etc.
- Carbon dioxide removal (negative CO_2-emissions) if CCS is applied or biochar produced.

Adding renewable hydrogen to biobased value chains represents another strong link between hydrogen and biomass/bioenergy. In principle, renewable hydrogen integration into biobased value chains can be done to 1) replace conventional, fossil hydrogen use, 2) upgrade the quality of products, or 3) produce (additional) products and by-products.

Within the IEA Bioenergy strategic Inter-task project “Synergies of green hydrogen and biobased value chains deployment” the focus is on the value chains directly linked to bioenergy, i.e., 1) biomass as a source of hydrogen production (biohydrogen) and 2) biobased processes utilizing renewable hydrogen. Representative examples are showcased to describe the potential role of biobased value chains linked to the hydrogen economy, and to create a clearer overall picture of the promising value chains and their potential for future applications.

In this report, we provide an overview of prominent projects (case studies) on the potential use of renewable hydrogen in biobased value chains from different parts in the world. We give a concise presentation of a variety of projects looking into synergies of hydrogen and biomass deployment and put the results into a broader context with regards to their technological and economic performance. Finally, we discuss further (research) needs for realizing projects.

In section 1, we present a general overview on concepts and describe the selection and basic characteristics of the concepts based on exemplary case studies. This includes an evaluation of these type of concepts concerning their characteristics, TRL and technical barriers for their implementation. Followed by conclusion with a forward-looking discussion into issues on the need of further research This is followed by individual presentations of each of the case studies in section 2. Thereby a uniform template has been used for presenting the different first-of-its-kind projects.

Summary

Over the duration of the IEA Bioenergy triennium 2022 to 2024, a consortium of IEA Bioenergy Tasks – 32, 33, 34, 36, 37, 39, 40, 42, 44 and 45 – collaborated on an inter-task project called Synergies of green hydrogen and biobased value chains deployment. Hydrogen is a very cross-cutting topic and the strategic inter-task project is a collaborative effort of the IEA Bioenergy TCP Tasks and also in collaboration with the Hydrogen TCP.

The objective of the project was to identify and assess technologies for producing hydrogen from biomass as well as synergies in the deployment of green hydrogen and biobased value chains that can enhance the use of biobased value chains in the energy system.

In the report on hand opportunities for adding renewable hydrogen to biobased value chains are presented and discussed. In principle, renewable hydrogen integration into biobased value chains can be done to 1) replace conventional, fossil hydrogen use, 2) upgrade the quality of biobased products, or 3) produce (additional) biobased products and by-products. The options differ in the degree of technological and process adaptation and development required. Adding renewable hydrogen vividly shows the potential synergies between renewable hydrogen and biobased value chains that can create various benefits.

The description and discussion of potential technologies and concepts - including 1) technology readiness and economic fundamentals and 2) climate effects and role in the energy system - are done through case studies. This serves to increase visibility and share state-of-the-art knowledge of promising applications.

From the various case studies the following insights can be gained. One obvious advantage for biobased value chains is the integration of methanation (or generally PtX) into the process without the need to separately capture CO_2, thus saving CAPEX and OPEX cost related to the provision of (relatively) clean CO_2. Further on, CO_2 capture could be beneficial either through efficient process integration and/or due to high CO_2 concentrations in the offgases when compared to CO_2 captured from air or sea water. While these are general benefits from the integration of CO_2 activation in biobased value chains, ultimate cost of available CO_2 will always be determined by additional factors such as e.g. cost of electricity, scale of realization, local infrastructure, contaminants in the offgas etc.

Introduction

A future H_2 based economy, or even substantial addition of H_2 as energy carrier in the global/ local economy, requires substantial additional infrastructure – not only for production but also for transport and storage. There are severable technologies available to realize transport and storage of H_2, which all have their own KPI (Key Performance Indicators) so that their suitability for future application can be assessed [3]. One possibility is to transform H_2 into other energy carriers such as e.g. methanol or NH_3.

In many biobased value chains, notably also in already established ones such as e.g. HVO/HEFA, H_2 is required to produce a high-quality transportation fuel. This is an obvious and very fundamental synergy between a H_2 economy and biobased value chains. Ultimately, existing or to be developed biobased value chains can act as energy carrier both for transportation and storage for H_2. At the same time there are fundamental differences to technologies that are typically considered for H_2 transportation and storage. In many cases, such as e.g. HVO/ HEFA, H_2 is integrated for the production of a final product rather than for the production of an efficient intermediate H_2 carrier which is transformed into H_2 to enable H_2 consumption in a different place and/or point of time. Apart from that, there are many examples where biobased carbon products are – or can be – used primarily as H_2 carrier (e.g. methane and methanol). This means that when integrating H_2 into biobased value chains, the properties of H_2 transport and storage always play a role, although the nature, benefits, and implementation of these properties vary greatly.

In addition to above mentioned benefits of realizing H_2 transportation and storage, there are several other synergies that can be created. These vary along the different cases, but also along the different technical barriers that are needed to achieve TRL 9 (see Figure 1). Exploring onsite synergies for the case of 'green' hydrogen represents a comparably low technical barrier. In other existing biobased value chains, hydrogen could be potentially used to adapt and improve processes, which goes beyond a 'simple' replacement of the H_2 source. There are also some new concepts of biobased value chains that rely on the use of (additional) hydrogen, which naturally represent a very high technical barrier for implementation. Irrespective the level of technical barrier, several benefits from integrating hydrogen in biobased value chains are anticipated. These benefits are categorized in the following four main effects for the purpose of this study:

A. **Integration of hydrogen in existing fuel infrastructure.**
   The use of hydrogen in carbon based value chains matches closely existing infrastructure for fuel logistics and use. Little to no adaptation would be required, including the end use. Some of these fuels are liquid in ambient conditions, which results in easier storage, transport, processing and use as compared to gaseous hydrogen.

B. **Onsite synergies through process integration of electrolysis and biofuel production.**
   This involves both synergies based on using the same infrastructure and administration, but also leads to reduced logistics/ transportation to achieve cost savings. Also, heat flows inside an integrated facility could be optimized. Moreover, integration of electrolysis promotes a distributed use of volatile renewable energies and provides a local chemical energy storage. The use of excess renewable energy has the potential to further reduce hydrogen cost.

C. **Increase of carbon efficiency of biobased value chains.**
   Conversion of biomass to fuels results in losses of carbon, quite often (but not always) as CO_2. The use of hydrogen in these conversion chains has the potential to significantly improve the carbon to fuel efficiency. In some cases, improved product quality is achieved.

D. **Required to enable BECCU liquids/ gases (over ‚simple' BECCS).**
   In the case of carbon capture and storage from biomass combustion flue gas (BECC), hydrogen is required to enable activation of the captured CO_2 (BECCU). In addition to the provision of heat (and power) as the primary objective of the value chain, liquid and gaseous carbon products can be co-produced.

The higher the technical barrier for implementation (i.e. for new process concepts), the more benefits are theoretically possible to achieve (see Figure 1).

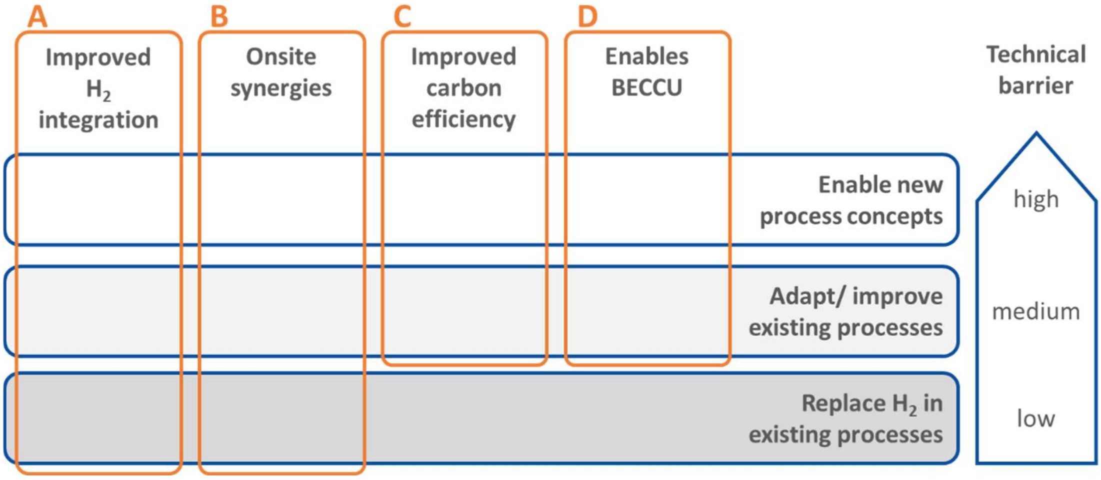

Figure 1: Matrix of potential benefits (A-D) and level of technical barrier for integration of hydrogen in biobased value chains.

Table 1: Overview of case studies and their link to investigated benefits (A-D) and technical barrier (dark grey: low, light grey: medium, white: high) (compare Figure 1) and type of hydrogen use (compare Figure 2)

Project/ company	Country	Technology	Feedstock	Products/ by-products (energy, materials)	specific comments related to the technology to outline advantages (e.g. hard-to-abate sectors)	other system services (integration of VRE, CCS, CCU, etc.)	Investigated benefits (potential benefits in ())	Type
Skærbæk	Denmark	Biomass combustion with CCUS	Wood chips	Heat, electricity	Allowing for substantial CCS/CCU	CCS, CCU	(A), B, D	II
LTU Green Fuels	Sweden	Hybrid biomass gasification and electrolysis process	Black liquor	Methanol/DME	-	-	(A), B, C, D	II
reFuels	Germany	Green refineries with focus on fuel production	Focus RED Annex IX A	SAF, Marine fuels, Road transport fuels	Refinery approach based on renewable intermediates	-	(A), B, C, D	I
Electrochaea	Denmark	Biological methanation		Heat	Renewable methane for a variety of uses in different sectors	-	(A), B, C, D	II
Pilot-SBG	Germany	Catalytic methanation	RED Annex IX A	BioLNG Fertilizer, HTC products like biochar	Renewable methane for LNG for heavy duty road transport and marine	Link to CCU via SynBioPTx	(A), B, C, D	II
HyFuelUp	Portugal	Dual Fluidised Bed - Sorption Enhanced Reforming Gasification + Fluidised bed methanation	Digestate and similar biomass	Bio-LNG	Bio-LNG for heavy transport	Flexible H2 addition according to its availability	(A), B, C, D	II
Lignin valorization	n/a	Upgrading/ hydrodeoxygenation of lignin	Lignocellulosic biomass/Lignin	Aromatic monomers/BTX/phenols/cycloalkanes	Obtaining aromatic and/or cyclic compounds unfeasible to be produced by other biobased processes. Valorization of waste biorefinery/pulp industry streams (I.e. lignin).	Opportunity of integration with Green Hydrogen.	(A), C	I
Steeper Energy, Aalborg University	Denmark, Canada	Upgrading of HTL oil	Woody biomass	Biofuels, methanol	-	CCS, CCU	(A), B, C, (D)	I
PTG-HEFA	Germany and other regions	PTG-HEFA-SPK	Different HEFA sources like veg oils, UCO	HEFA-SPK, HVO-Diesel, naphtha	Focus on aviation; study elaborated until 2017	Link to SynBioPTx	(A), B	I
Omega Green, Be8	Paraguay	electrolytical hydrogen in HVO plants		HVO, HEFA-SPK, green naphta	-	Link to SynBioPTx	(A), B	I

Case Study Overview

Ten different cases that relate to the integration of H_2 in biobased value chains have been collected and evaluated. All of these are presented in individual case studies attached to this report (Appendices: Detailed case studies). An overview is available in Table 1. The collected cases cover all categories of synergies identified in the introduction so that they will provide a good overview of the potential that lies in this integration. Concepts along the different technical barriers are represented, but one needs to have a closer look at the individual cases since the technology readiness level might vary a lot within the different technical barriers (see Table 2). The focus of the selected case studies was on technology, available information, and the aim to cover varying synergies/concepts. However, this aspect led to a rather poor geographic coverage and it has to be emphasized that the geographic focus that emerged does not indicate a certain priority or even level of activity in this field.

Table 2: Overview of case studies and their estimated technology readiness level (TRL)

Project/ company	Technology	Feedstock	Products/ by-products (energy, materials)	TRL (weakest link)	TRL (weighted average)
Skærbæk	Biomass combustion with CCUS	Wood chips	Heat, electricity	5	7
LTU Green Fuels	Hybrid biomass gasification and electrolysis process	Black liquor	Methanol/DME	4-5	6-7
reFuels	Green refineries with focus on fuel production	Focus RED Annex IX A	SAF, Marine fuels, Road transport fuels	5	6-7
Electrochaea	Biological methanation	-	Heat	8-9	8-9
Pilot-SBG	Catalytic methanation	RED Annex IX A	BioLNG Fertilizer, HTC products like biochar	4	7-8
HyFuelUp	Dual Fluidised Bed - Sorption Enhanced Reforming Gasification + Fluidised bed methanation	Digestate and similar biomass	Bio-LNG	5	6-7
Lignin valorization	Upgrading/ hydrodeoxygenation of lignin	Lignocellulosic biomass/Lignin	Aromatic monomers/BTX/ phenols/cycloalkanes	3-4	5-6
Steeper Energy, Aalborg University	Upgrading of HTL oil	Woody biomass		4-5	7-8
PTG-HEFA	PTG-HEFA-SPK	Different HEFA sources like veg oils, UCO	HEFA-SPK, HVO-Diesel, naphtha	9	9
Omega Green advanced biofuel plant	electrolytical hydrogen in HVO plants	-	HVO, HEFA-SPK, green naphta	9	9

A closer look at the case studies indicates that Figure 1 provides a general guidance through the varying possibilities to integrate H_2 in biobased value chains, but also that not all cases very clearly fit one or the other category. For the case of BECCU, it is very clear that carbon capture from flue gases of biomass combustion and subsequent CO_2 activation qualifies for this category. But from a fundamental point of view, activation of CO_2 as in biogas or gasification processes is very similar- the only difference is that the CO_2 is not necessarily captured/ recovered separately in this concept. It was decided to keep this structure to better distinguish in terms of business cases that need to be developed for the products. If H_2 is integrated in existing biobased value chains with the effect of improving carbon efficiency from feedstock to product (such as e.g. in the case of CO_2 methanation in a biogas process), the produced methane can be directly integrated in the existing biogas product logistics, including treatment, administration and sales. In contrast to this direct integration, producing carbon materials from CO_2 after biomass combustion (be it methane or methanol) demands for additional infrastructure both in terms of hardware and administration to trade this product (apart from the ambiguity that lies in defining a carbon to product efficiency).

Evaluation

It is important to distinguish between two fundamentally different roles H_2 might play in biobased value chains. The first is the use of H_2 to adjust the chemical structure of biobased molecules, typically through hydrogenation reactions (Type I use in Figure 2). In this case chemical features specific to biobased molecules are addressed. H_2 is required for these biobased value chains to be realized. The second is the use of H_2 to activate CO_2 generated during biomass conversion (Type II use in Figure 2), which is rather unspecific because the chemistry of CO_2 activation is independent from its origin (apart from feedstock specific impurities that might exist). Type II uses largely depend on external conditions that demand for CCU and ultimately its priority over e.g. CCS.

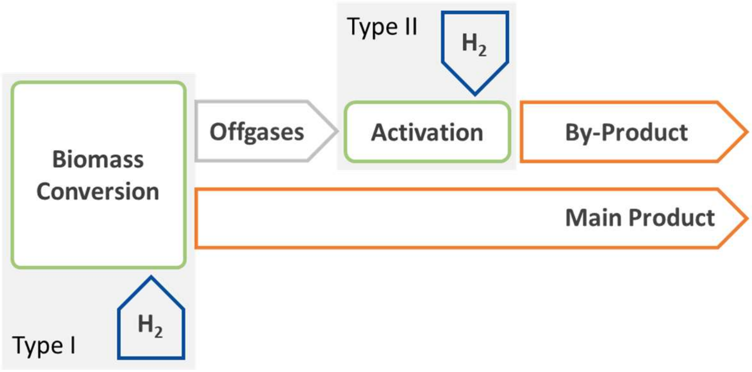

Figure 2: Point of H_2 addition in biobased value chains.

Whether CO_2 from biomass conversion processes is used for PtX value chains (and not, e.g. from direct air capture) depends on the cost of the CO_2 provided for activation. One obvious advantage for biobased value chains is the integration of methanation (or generally PtX) into the process without the need to separately capture CO_2, thus saving CAPEX and OPEX cost related to the provision of (relatively) clean CO_2. Such examples are shown in case studies Electrochaea, Pilot-SBG and LTU Green Fuels. In other cases, CO_2 capture could be beneficial either through efficient process integration (e.g. in the case HyFuelUp) and/or due to high CO_2 concentrations in the offgases (e.g. in the Skærbæk and Electrochaea cases) when compared to CO_2 captured from air or sea water. While these are general benefits from the integration of CO_2 activation in biobased value chains, ultimate cost of available CO_2 will always be determined by additional factors such as e.g. cost of electricity, scale of realization, local infrastructure, contaminants in the offgas etc.

The Clean Hydrogen JU has developed several KPIs to characterize H_2 technologies and pathways [3]. Some of the KPIs relevant for H_2 storage have been listed in Table 3 due to the relevance of storage aspects for H_2 integration on biobased value chains. Obviously, storage systems are benchmarked against H_2 liquefaction as comparably cost and energy efficient technology. This makes only sense for Type II H_2 integration.

If the production of methane integrated into biobased value chains is considered as H_2 storage concept, steam reforming of the produced methane to recover H_2 is an obvious benchmark. Most of the H_2 traded today is produced via steam reforming of natural gas and characterized by production cost in the same range as targeted H_2 liquefaction cost of 1.0-1.5 €/kg (compare Table 3), depending on price for natural gas and the scale of steam reforming plants [4]. The production costs of H_2 via steam reforming are governed by feedstock cost, which contribute around 75% for the case of natural gas [5]. For the production of methane from CO_2 offgases in biobased value chains it can therefore be concluded that methane production cost **excluding H_2 feedstock cost** will need to stay below 0.75-1.1 €/kg in order to be competitive against H_2 liquefaction. Additionally, benefits from producing methane (over H_2 liquefaction) are maximized if existing natural gas infrastructure (e.g. pipelines, steam reforming plants) can be utilized. H_2 production via methanol reforming has higher production cost than natural gas steam reforming at large scale, but benefits from a favourable cost development at lower reforming capacity, making it a competitive process at production capacity of around 100 m^3(STP)/h of H_2 or less [5].

Another KPI for H_2 storage targets energy efficiency. This is more difficult to evaluate since natural gas steam reforming plants are optimized in various ways depending on local conditions (infrastructure, feedstock cost, energy cost etc). The minimum energy demand for steam reforming is almost 9 kWh/kg [4], which only leaves around 3-8 kWh/kg H_2 recovered for the production and transportation of methane, including energy intensity of CO_2 methanation. In one theoretic evaluation of an integrated methanation process including methane purification a total of around 4 MW for compression and pumps is required for 200 kmol/h methane production (which could be theoretically generated from available excess heat) [6]. This would translate in around 5 kWh/kg H_2 recovered. Excluding transportation, storage via CO_2 methanation would sum up to around 14 kWh/kg H_2 recovered, which renders this generally a reasonable option according to the KPI set by the Clean Hydrogen JU (compare Table 3).

Both estimates of storage cost and H_2 carrier specific energy consumption have been conducted excluding cost and effort to separate CO_2. It is one decisive advantage of integrating CO_2 methanation in biobased value chains that it can potentially be realized in gas mixtures, avoiding the need to separate CO_2 from offgases. From the case studies presented in this report, only the “HyFuelUp” and “Skærbæk” cases requires CO_2 recovery from a combustion flue gas.

The effect of integrating H_2 in biobased value chains on carbon conversion efficiency, defined as the amount of carbon in the produced fuel over the initial amount of carbon present in the biomass feedstock, is not possible to be evaluated based on the information available for the case studies in this report. Again, such an evaluation is only reasonable for Type II H_2 integration. All concepts in this report rely on the use of offgases, i.e. all unconverted liquid/ solid byproducts that lower carbon efficiency are not addressed. It follows those thermochemical processes, especially those at high temperatures such as e.g. gasification, could potentially reach a carbon efficiency close to 100% if H_2 is integrated in the process. Biological conversion can also significantly improve its carbon efficiency by H_2 integration but are limited by the cellulosic fraction of biomass feedstocks that they target.

Table 3: Selected KPIs for hydrogen carriers

Parameter	Unit	SoA	Target 2024	Target 2030
H_2 liquefaction energy intensity	kWh/kg	10-12	8-10	6-8
H_2 liquefaction cost	€/kg	1.5	<1.5	<1.0
Hydrogen carrier specific energy consumption	kWh input/ kg H_2 recovered	20	17	12

Conclusions

All these observations lead to two basic conclusions for the evaluation of targeting biobased value chains for H_2 storage:

1. Type I use where H_2 is needed to achieve a specific product quality does represent H_2 storage with several synergies, but is unlikely to be applied as dedicated means to store H_2 because such projects will most likely be realized for other reasons. This use of H_2 could rather represent an offtake that requires H_2 transportation and storage to the facility if H_2 production is not integrated in the biomass conversion process onsite. Following, such an integration onsite is highly favourable (see e.g. the case PTG-HEFA) and is already realized (case Omega Green).

2. Type II use of H_2 to valorise by-products from biomass conversion generally qualifies as being targeted for dedicated H_2 storage via CO_2 activation. These biomass conversion processes function well without CO_2 activation, and integrating it into the process can produce a variety of synergies that increase economic feasibility. However, they also require that the biomass conversion process is co-located with H_2 production and/ or ideally integrated. Whether biobased CO_2 is used for H_2 storage is largely dependent on its price compared to other CO_2 that is or can be provided in proximity (e.g. via direct air capture).

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Appendices: Detailed case studies

Appendix A - Biomass combustion with carbon capture for methanol production at large wood chip fuelled CHP plant in Denmark

BACKGROUND

Denmark's climate goals mandate a 70% reduction in CO_2 equivalent emissions by 2030 compared to 1990 and aims for climate neutrality by 2050. In this light, the imperative to explore innovative solutions has become paramount. In response, the company Ørsted A/S secured a Carbon Capture and Utilization (CCUS) tender in May 2023, committing to capture 430,000 t/y of CO_2 per year over a 20-year period from two biomass-fired power plants. This initiative is mandated to be operational by 2026.

The momentum towards sustainable practices in Denmark continued in August 2023 with the announcement of a new tender focused on negative emissions (NECCS). Anticipating an annual capture of 500,000 tonnes of CO_2 from 2026 onwards, this initiative represents an important step towards realising Denmark's climate objectives. Bolstering these efforts, a comprehensive CCS strategy was endorsed by a substantial majority in Parliament on 20th September 2023. The strategy earmarks a budget of 3.6 billion euros for two forthcoming tenders, collectively targeting a minimum annual capture and sequestration of 2.3 million tonnes of emissions over 15 years, commencing in 2029. This encompasses not only the objectives set by the recent tenders but also builds upon them, striving for a total capture of at least 3.2 million tonnes per year, including initial commitments.

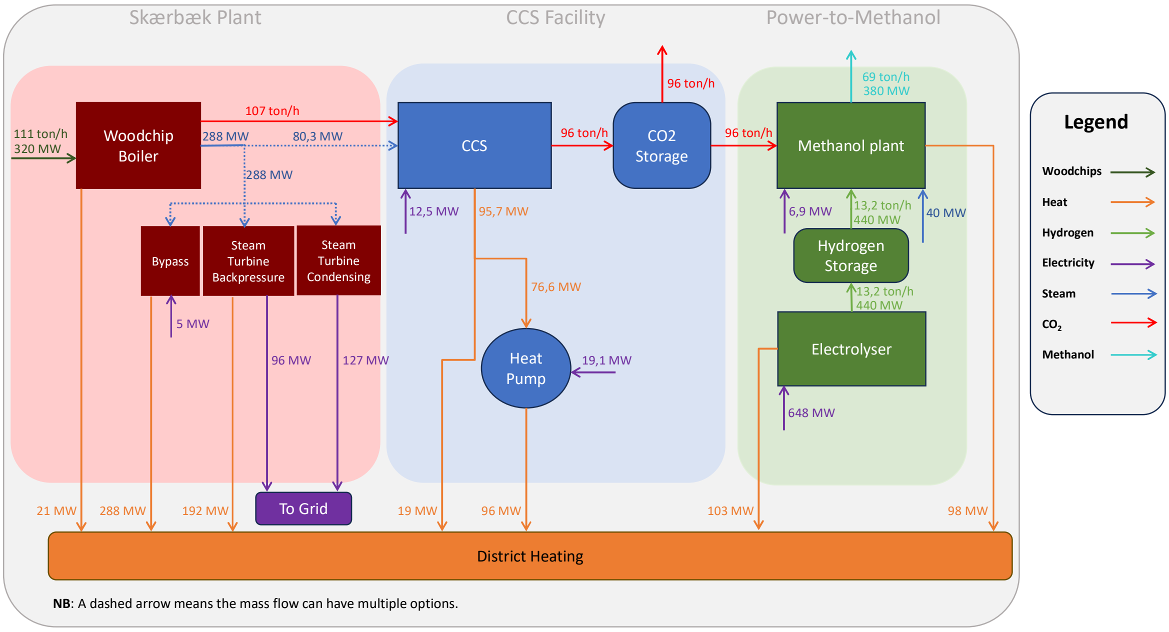
Against this backdrop, the case at hand involves the modelling of the interaction between a biomass Combined Heat and Power (CHP) plant and Carbon Capture and Storage (CCS) or Carbon Capture and Utilization (CCU). The Skærbæk Combined Heat and Power Plant has been identified as a pertinent case study for this investigation. The study aims to scrutinise the full-scale implementation of CCUS technologies. By delving into these technologies within the context of a biomass-fired CHP plant, the analysis seeks to elucidate the consequences of adding these technologies to the operation of the plant and the district heating system.

Table A.1: Technology profile for the Skærbæk wood chip CHP plant, please also refer to [1].

Designation	Skærbæk Biomass CHP Plant CCUS
Developer/Promotor	Ørsted A/S
Conversion technology	Combustion of wood chips on water cooled vibrating grate
Feeding System	Conveyors, day silo, gravity and air spout firing system
Principal Feedstock(s)	Moist wood chips chipped from forest residues
Principal Application(s)	Heat for district heating, power for the Nordic grid, carbon for methanol synthesis
Scale	2.5 TWh/y wood, 1.7 TWh/y heat, 750,000 t/y CO_2
Development Status	Combustion plant in operation since 2017, carbon capture project developed, realisation being considered. Case study models four scenarios of full-scale installation of CCS/CCU

PROCESS DESCRIPTION

A process scheme including a simplified energy balance of the Skærbæk CCUS case is shown in Figure A.1. The Skærbæk CHP plant is located in the Western part of Denmark where it delivers heat to one of Denmark’s largest district heating systems. Today, the Skærbæk CHP plant delivers app. 50% of the heat (3.3 PJ) and electricity for the grid, using approximately 550,000 tonnes of wood chips per year.



NB: A dashed arrow means the mass flow can have multiple options.

Figure A.1. Process scheme including energy balances across the system boundary for production of methanol.

The aim of the study is to analyse the consequences of investment in CCUS technology at the Skærbæk plant. The study analyses four scenarios, one of them being the CCUS scenario. The Balmorel energy system model serves as the foundational framework for the in-depth

simulations conducted in this study while an OptiFlow module, an extension of Balmorel, enables detailed modelling of different processes at the Skærbæk plant. Input data used are all publicly available. Please refer to [2] for more information on the models and assumptions.

This scenario entails the incorporation of carbon capture, storage, and utilisation at the Skærbæk Power Station where all or part of the CO_2 can be directed to an electricity-powered methanol production facility with hydrogen provided by a large electrolyser. The production capacity of the methanol process is dimensioned to match the CO_2-production from the biomass boilers at full capacity.

The amount of methanol produced over the year is assumed to be fixed and based on 4,500 full load hours operation of the methanol plant. However, it can be chosen to run the biomass boilers for more than 4,500 full load hours to provide more heat and power. Also, during this operation, CO_2 could be captured and sold at an assumed market price.

In the analysis of the CCUS scenario for 2035 a significant shift is observed in the dynamics of net electricity production, transitioning from positive to negative on average over the year. District heating generation is increased significantly compared to the business-as-usual scenario and in this scenario the heat production from the Skærbæk plant can cover almost all the district heating demand in the TVIS area.

Most of the heat-generation is delivered as surplus heat from the CC-plant, the electrolysers and the methanol plant, while the CHP plant delivers process heat for the CCUS-processes and increases power production in condensing mode. The negative net electricity generation underscores the energy-intensive nature of the CCUS process, necessitating a substantial power input of app. 3.2 TWh.

Hydrogen production is optimized for periods when the electricity value is low, while district heating demand proves significant only in winter months.

DEVELOPMENT STATUS, APPLICATIONS, AND PRODUCTION SCALE

While neither CCS nor CCU has yet been implemented in reality at the Skærbæk plant, the case is totally far-fetched. Ørsted is up to 2026 busy with partners developing CCS at the Avedøre (wood pellets) and Asnæs (wood chips) CHP plants and must capture and store significant amounts of CO2 from 2026.

Ørsted has been shortlisted in a public CCUS tender with a project at Skærbæk. Conditions for methanol production still have to mature – large scale H_2 production likely being the weakest link, and CCS currently seems to be a more viable pathway. The new tender focuses on CCS but enables an option to change for utilisation during the contract duration should conditions and technology option improve and a business case emerge.

ASSESSMENT OF THE TECHNOLOGY READINESS

An assessment of the technology readiness of the Skærbæk CCUS concept is provided in Table A.2.

Table A.2. Assessment of technology readiness for implementing CCUS at Skærbæk

Process steps	TRL	Weight [%]	Comments
Feedstock handling systems, combustion on water cooled vibrating grate in a 3-pass boiler	9	10	Proven for various feedstock in numerous plants around the World; however, the two 160 MW units at Skærbæk may be some of the World's largest grates of this type
Post combustion carbon capture - amine based	8	40	Proven in relevant size at relevant plant types. Relevant full-scale experiences will soon be available from Ørsted's CCS projects at the Avedøre and Asnæs plants that must capture and store 430,000 t/y CO_2 starting from 2026
Alkaline electrolyses	7	20	Yet to be proven in very large size at utility

Process steps	TRL	Weight [%]	Comments
Integrated full scale operation of methanol synthesis	5	30	Integrated operation in full scale/real life has yet to be proven
Overall “Weighted Average”	7		
Overall “Weakest Link”	5		Integrated full scale operation to be proven

ECONOMIC AND ENVIRONMENTAL ASSESSMENT

In a scenario where the Skærbæk CHP plant is equipped with CC-technology, electrolysers and methanol production, the surplus heat from the different processes increases the heat production from the plant significantly. As shown in the figure below, the heat generation of the plant increases to app. 6,100 TJ, effectively fulfilling almost the entire district heating demand.

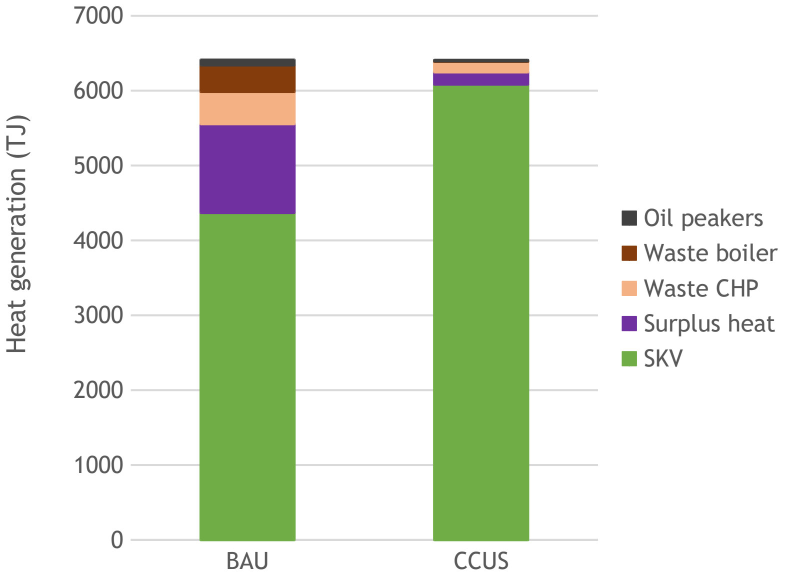


Figure A.2: Annual heat generation in the two scenarios for the TVIS system in the year 2035 given in TJ.

There is significant heat generation throughout the year as shown in the figure below.

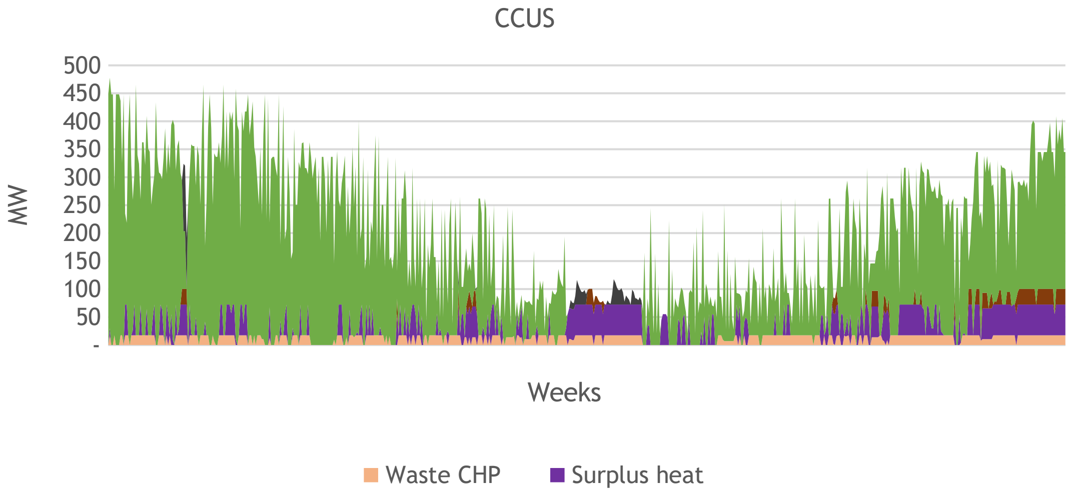


Figure A.3: Hourly heat generation in the CCUS scenario for the TVIS system in the year 2035 given in MW.

The duration curve (Figure A.4) shows that the Skærbæk plant covers most of the heat demand and that the heat storage is used to make the generation more flexible.

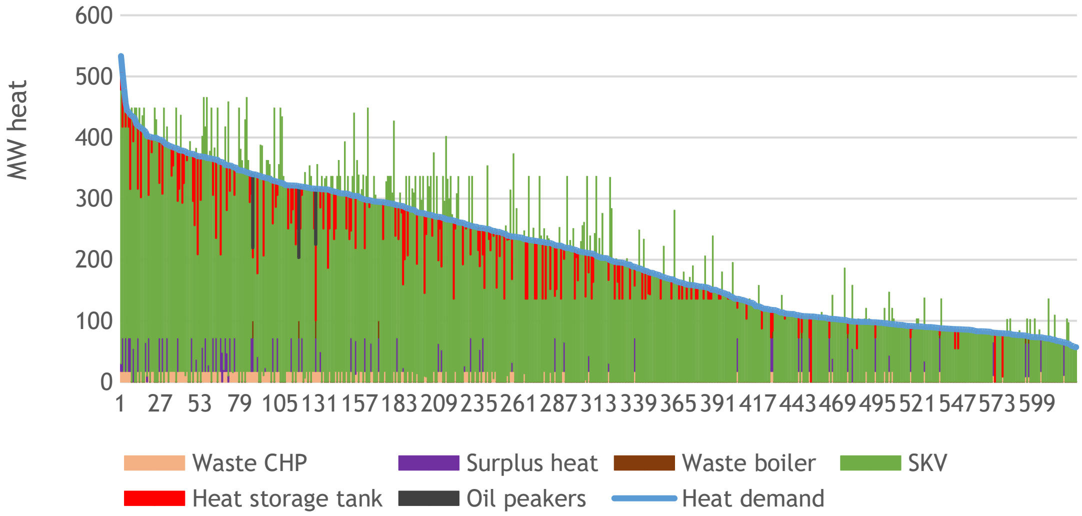


Figure A.4: Duration curve for heat demand and supply in the CCUS scenario

The figure below illustrates what part of the Skærbæk plant delivers the heat production in the BAU and CCUS scenarios.

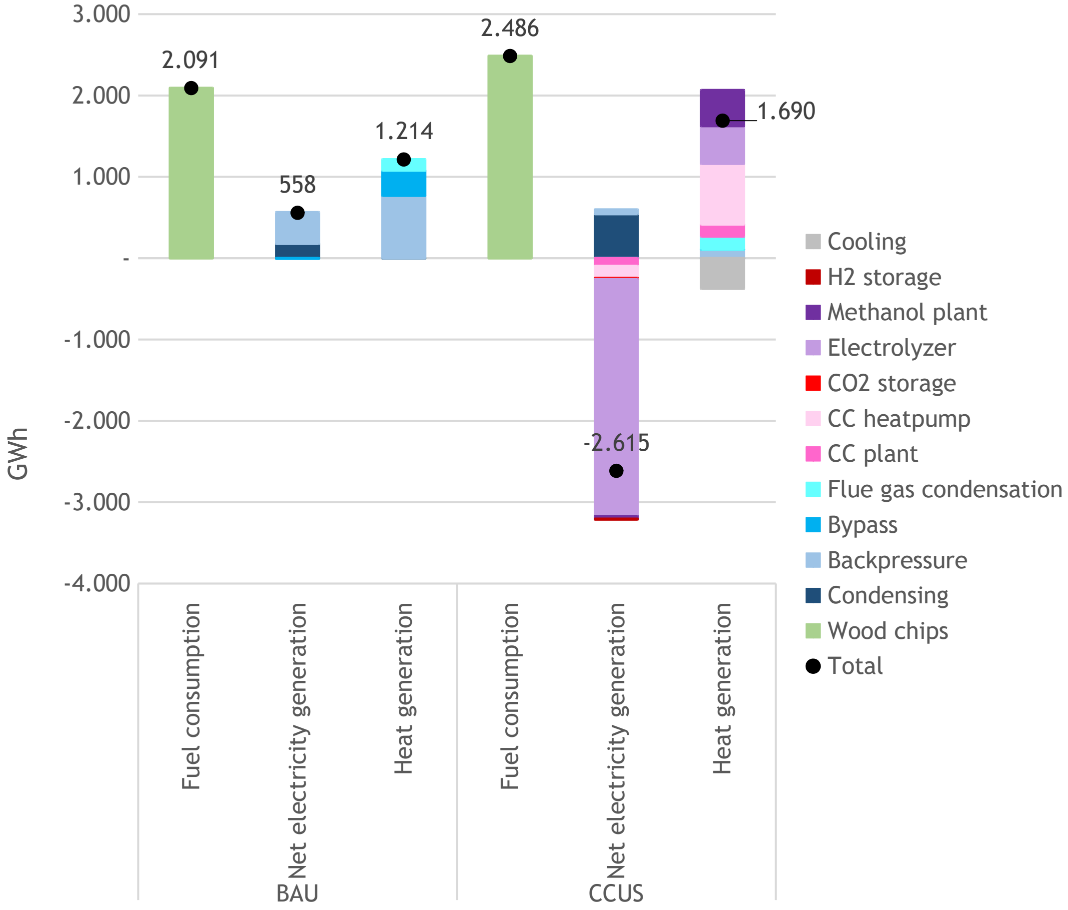


Figure A.5: Annual fuel consumption, heat and electricity generation on the Skærbæk plant in the two scenarios in the year 2035 given in TJ.

In the CCUS scenario, a significant part of the boiler's energy production is used for process heat in the CC-plant and in the methanol plant. This means that the direct heat generation from the boiler and turbine is reduced significantly. On the other hand, surplus heat from the CC-plant, electrolyser and methanol plant gives a large contribution to the heat production. Also, the large quantity of excess heat from the CCUS-process means that a significant part of the heat needs to be cooled away.

However, a consequential trade-off manifests in the net electricity generation, which experiences a substantial reduction from 647 GWh in the BAU to -2,593 GWh with CCUS. This negative net electricity generation underscores the energy-intensive nature of the CCUS process, necessitating a substantial power input of app. 3.2 TWh.

As shown below, the total generation of CO_2 from the boilers increases in the CCUS scenario because of increased operation hours. About half of the CO_2 is captured for use in the methanol plant and the remaining CO_2 is captured and stored.

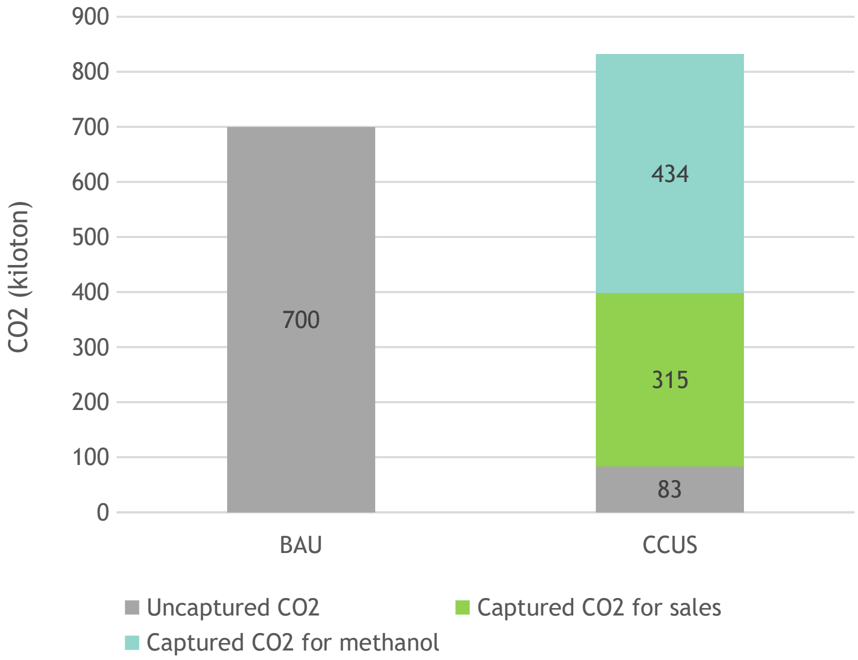


Figure A.6: Annual CO_2 flow on the Skærbæk plant in the two scenarios in the year 2035.

Analysing the duration curve for the electrolyser shown in Figure A.7 reveals that the electrolysers are operated when electricity prices are the lowest. The cost of electricity is the most important economic factor in the methanol production and electrolysers are therefore optimised to run during the hours with lowest electricity prices. However, the fact that the methanol plant must operate for 4,500 full load hours means that the plant also needs to operate during hours with moderate electricity prices.

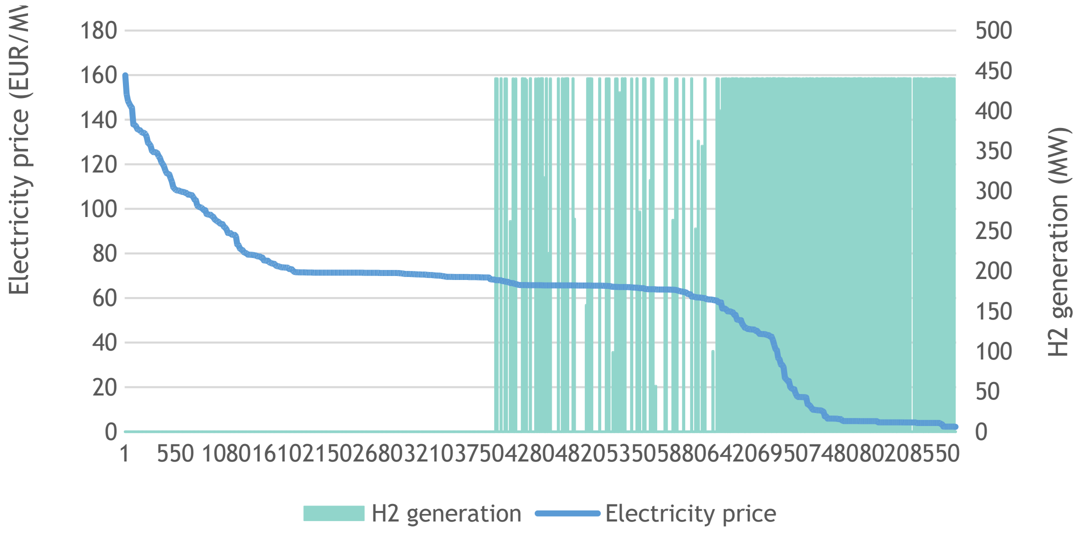


Figure A.7: Duration curve of the electricity price in Western Denmark and the corresponding hydrogen generation.

Finally, examining the trend depicted in Figure A.8 unveils the optimization mechanisms deployed within the plant in response to electricity price fluctuations.

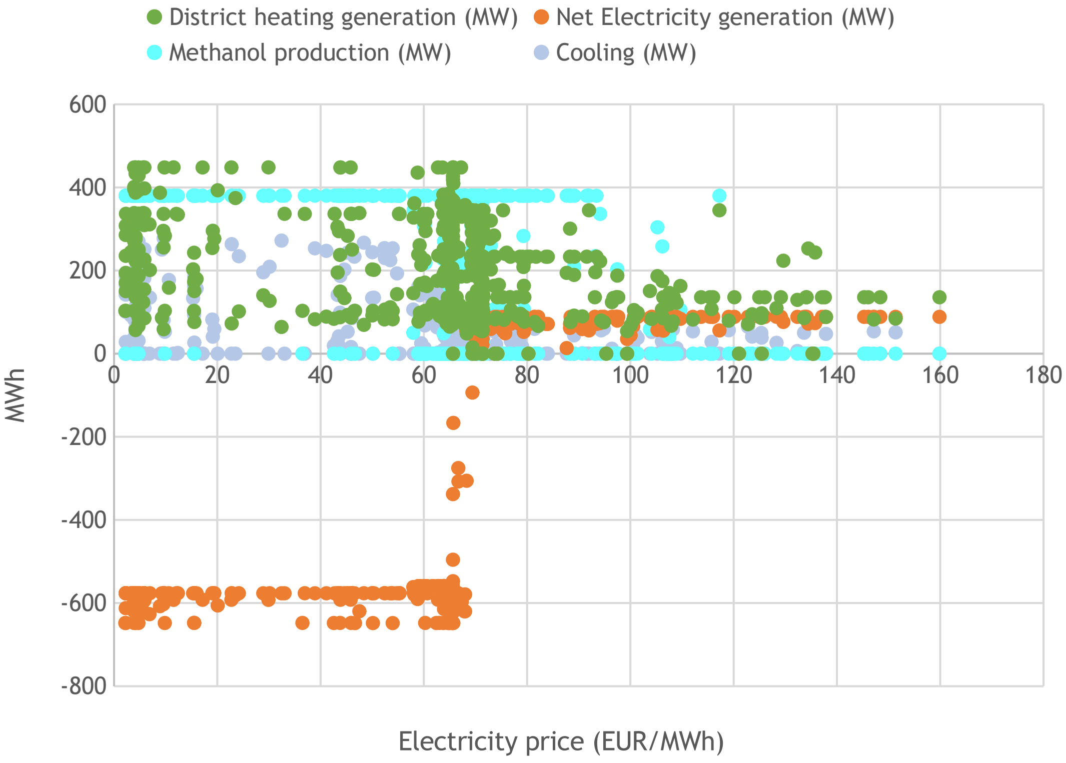


Figure A.8: XY-diagram of different generation modes on Skærbæk (Y-axis) at different electricity prices (X-axis)

When the prices are lower than roughly 70 EUR/MWh, corresponding to the lower half of the price spectrum, the system exhibits a peak methanol production of 380 MWh alongside a distinct negative net electricity generation.

As prices exceed 50% of the maximum threshold, methanol production becomes economically unviable, prompting cessation. This shift allows the CHP facility to have a positive net electricity generation and deliver power to the grid.

DEVELOPER FEEDBACK

While CCS would be a more probable pathway than CCUS at the Skærbæk CHP unit, recent geopolitical developments have changed the investment climate for Ørsted, and the company no longer intends to send an offer to meet the to meet the public 2025 CCS tender.

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Appendix B - Hybrid black liquor gasification and electrolysis for methanol production - The LTU Green Fuels case

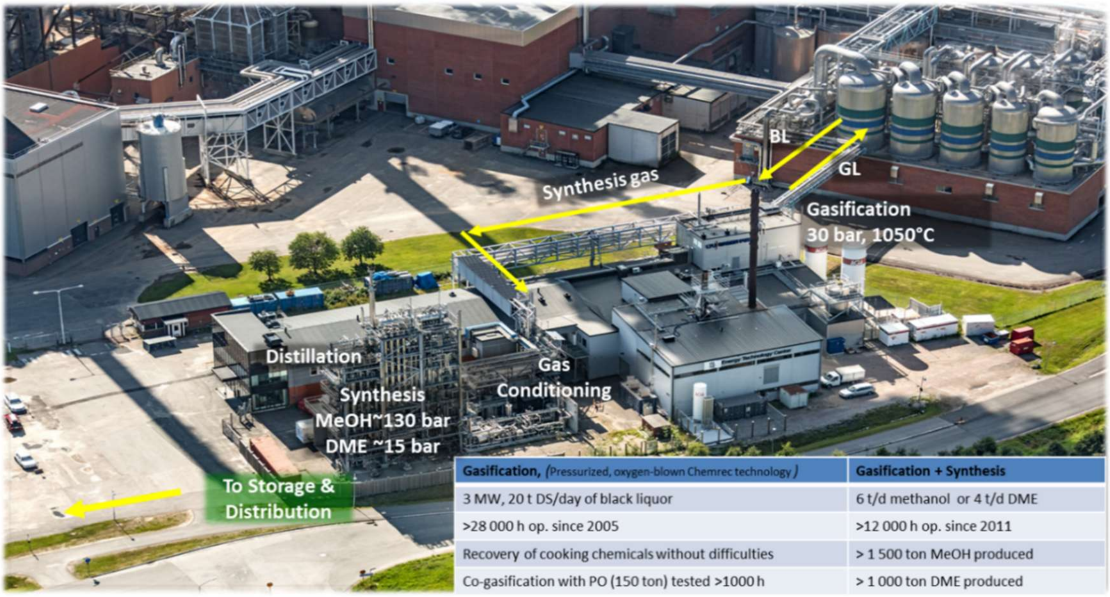
BACKGROUND

Luleå University of Technology is the owner of LTU Green Fuels AB (*fka* Chemrec black liquor gasification and Bio-DME pilot). The construction of the pressurized development plant, DP-1 of 20 tons dry substance per day capacity was done in the year 2005 at RISE ETC, close to the host mill Smurfit Kappa Kraftliner in Piteå from where the black liquor is obtained.

In Figure B.1, the LTU Green Fuels plant and some key operating data as well as data on the accumulated operating hours are shown. Maximizing the biogenic carbon utilization from feedstock to product via hydrogen addition is seen a promising pathway to maximize product volumes from a given amount of biomass. The use of hydrogen produced from fossil-free power or low carbon-intensity sources can additionally double or even triple the biofuel production in relation to the output of such biofuel plants. The is mainly due to that carbon in the form of CO is normally “sacrificed” in the water-gas-shift reactor (WGS) to make hydrogen and obtain a correct gas composition while the CO_2 is expelled from the process in the acid gas removal unit (AGR). This enhancement requires investment into an electrolyser to produce hydrogen that also, simultaneously, produces oxygen that could partially or completely substitute oxygen from the Air Separation Unit (ASU). In some cases, it would be possible to eliminate both the WGS and the ASU completely. This electrolyser oxygen contributes to lower production costs, eliminating the WGS and ASU will result in lower CAPEX.

This case study is based on the current LTU Green Fuels plant supplied with black liquor from Smurfit Kappa Kraftliner in Piteå, a Swedish integrated pulp and paper mill. Here, it is assumed that only a smaller fraction of the total black liquor flow is used, which means that the pulping capacity can be increased to the same fraction and thereby keep the thermal load on the recovery boiler the same as under normal operation.

More information about the concept and the data used in this case study (with minor modifications) can be found in Jafri et.al (2020) [1].



Gasification, (Pressurized, oxygen-blown Chemrec technology)	Gasification + Synthesis
3 MW, 20 t DS/day of black liquor	6 t/d methanol or 4 t/d DME
>28 000 h op. since 2005	>12 000 h op. since 2011
Recovery of cooking chemicals without difficulties	> 1 500 ton MeOH produced
Co-gasification with PO (150 ton) tested >1000 h	> 1 000 ton DME produced

Figure B.1. LTU GreenFuels plant process and data

Table B.1: Technology profile for LTU Green Fuels plant for methanol production.

Designation	Description
Developer/Promotor	Chemrec and LTU
Conversion technology	Entrained flow gasification process
Principal Feedstock(s)	Black liquor
Principal Application(s)	Methanol, DME, FT-products
Scale	Up to several hundreds of MW_{th} input
Development Status	Technical testing in pilot scale (3 MW gasifier)

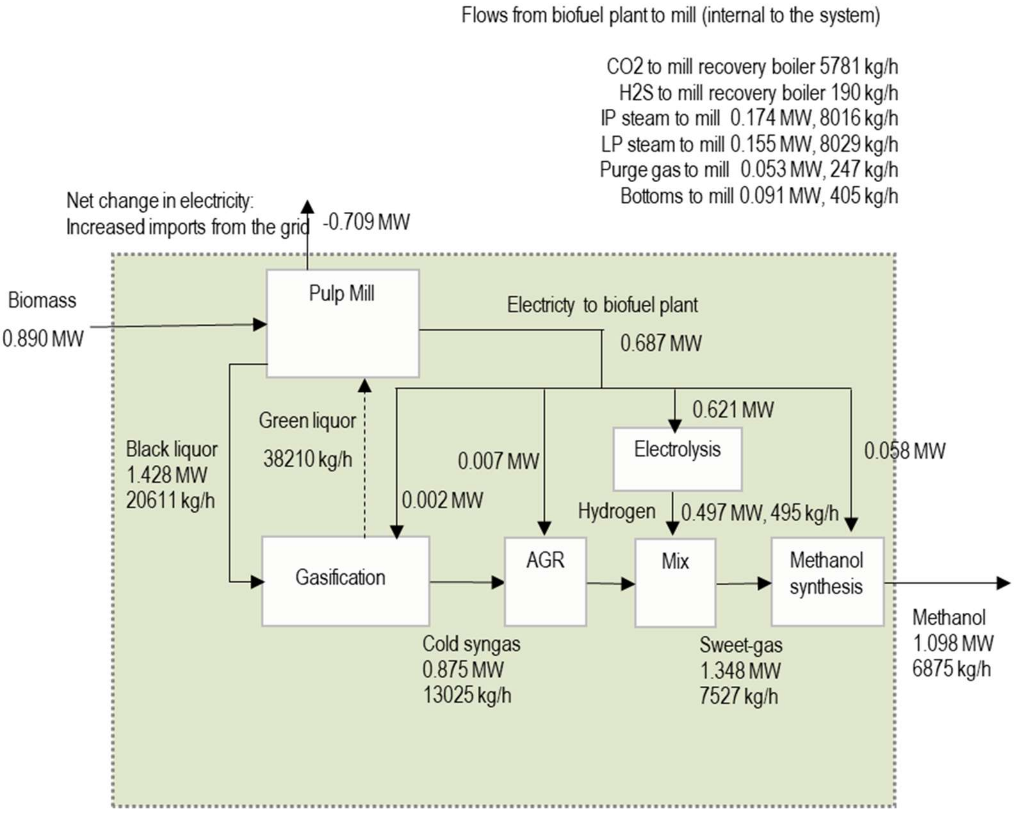


Figure B.2. Process scheme including material and energy balances across the system boundary for production of approximately 1 MW of methanol via black liquor gasification with added hydrogen. Adapted from Jafri et al. [1].

PROCESS DESCRIPTION

A process scheme including simplified material and energy balances (normalized to approx. 1 MW of methanol output) of the black liquor gasification concept with hydrogen addition for methanol production is shown in Figure B.3. The core of the concept is the Chemrec gasifier - a refractory-lined entrained bed reactor in which concentrated black liquor is gasified at around 1 000°C (*Figure B.3*) using oxygen as oxidizing and atomizing agent to produce a clean syngas made up primarily of carbon monoxide, hydrogen and carbon dioxide. Small quantities of hydrogen sulfide and methane are also present. The alkaline pulping chemicals in the black liquor form a molten slag, which is dissolved in water and returned to the pulp mill as green liquor for re-use in the kraft cycle.

The gas leaving the quench dissolver is cooled producing LP and IP steam. The cooling is done in counter current mode which means that the gas is efficiently washed of particulate matter. The gas is then free of melt droplets and can be scrubbed for H_2S removal and then used as a clean fuel or syngas.

The downstream processing contains acid gas removal (amine wash) and methanol synthesis. The hydrogen is assumed to be produced by PEM electrolysis of water with an electricity-to-product efficiency of 80% on a HHV basis. The oxygen produced as a by-product of electrolysis is sent to the gasifier.

In gasification-catalytic synthesis, the hydrogen sulphide separated in the amine wash is returned to the recovery boiler to avoid affecting the overall mill chemical balance. Purge gas from the methanol synthesis is fired in the mill lime kiln. Electricity is imported from the mill. LP and IP steam are exported from the biofuel units to the mill.

The electricity balance of the system is affected by the hydrogen production leading to increased import of electricity from the grid compared to normal mill operation. For this specific pulp mill, additional biomass needs to be supplied to the power boiler to meet the requirements of units such as the paper machines.

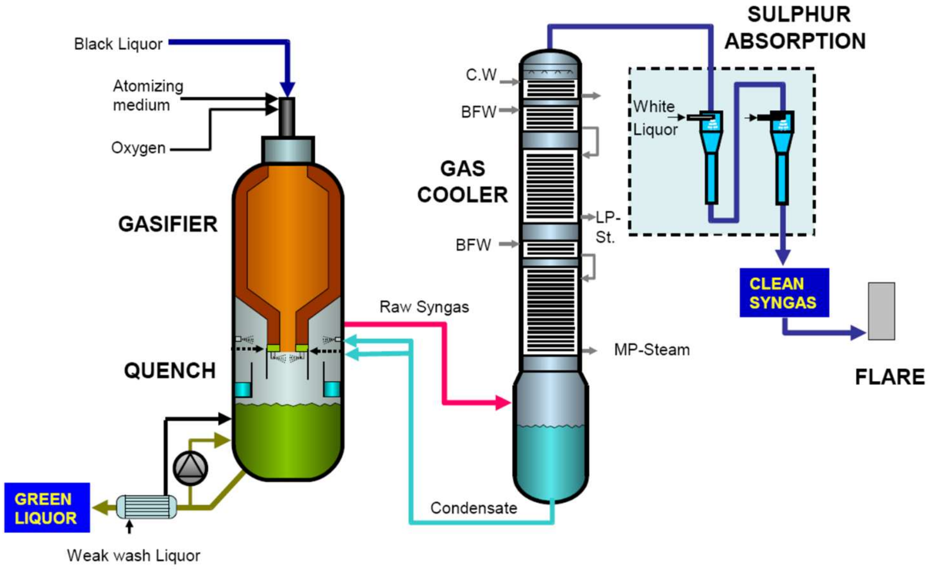


Figure B.3: The CHEMREC™ black liquor gasifier

Green liquor derived from gasification of black liquor has a higher concentration of carbonate ions compared with recovery boiler green liquor. The specific energy requirement of the causticizing process in which carbonate is calcined becomes 40-42% higher, which translates into an additional lime kiln energy requirement of 7.5-7.8%. In this system, this was met by purge gas from the methanol synthesis.

Regarding the effects on the steam system, the process delivers significant amounts of excess LP and IP steam to the pulp mill, which leads to less steam passing through the backpressure turbine and consequently lower electricity production. There are also relatively large amounts of excess hot water at around 90 °C, which has not been accounted for here, but which could, e.g., be used for make-up water preheating or be exported to a district heating network, if present.

DEVELOPMENT STATUS, APPLICATIONS, AND PRODUCTION SCALE

The pilot plant has since mid-2016 been in conservation state. Currently, there is however several research initiatives that aims to put the plant into operation again. One is, like described in this case study to boost the well-tested syngas production from black liquor, with addition of hydrogen produced via water electrolysis using fossil-free power.

In the autumn 2023, funding from the European Union and Swedish Agency for Economic and Regional Growth was granted to build a test bed for electrolysis systems (approximately 1 MW) at the LTU Green Fuels pilot plant site to carry out tests with electrolysis for producing hydrogen and oxygen in an arctic climate. This also open the opportunity to test and validate the concept of hybrid black liquor gasification and electrolysis process for methanol production.

ASSESSMENT OF THE TECHNOLOGY READINESS

An assessment of the technology readiness of the LTU Green Fuels process with hybrid black liquor gasification and electrolysis process for methanol production is provided in the table below.

Table B.2: Assessment of technology readiness of the hybrid black liquor gasification and electrolysis process for methanol production

Process steps	TRL	Weight [%]	Comments
Feedstock handling system	-	.	NA. Black liquor delivered from the pulp mill
Black liquor gasification system	7	35	The gasification system is proven on pilot scale (3 MW input) in Piteå, Sweden. Accumulated operation 26000 h and maximum continuous operation 500 h.

Process steps	TRL	Weight [%]	Comments
Hydrogen production	6	20	PEM Electrolysis, -1 point as integrated operation is not tested.
Gas cleaning & conditioning	8	10	Commercial technology (Rectisol), -1 point for operation with bio-syngas.
Methanol synthesis and purification	8	20	Commercial technology (BWR), -1 point for operation with bio-syngas.
Integration with pulp mill, green liquor and steam	4-5	15	Not demonstrated since GL flow from gasifier is negligible in pilot. GL analysed and upgraded to WL demonstrated in bench scale.
Overall “Weighted Average”	6-7		
Overall “Weakest Link”	4-5		Integrated operation to be proven

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Appendix C - REFUELS | Synthetic fuels from hydrogen-boosted biomass-to-liquids processes

BACKGROUND

At KIT, a biomass-to-liquids process was developed to convert residual lignocellulosic types of biomass into synthetic fuels. Essential feature of this process was its decentralized-central concept, by which biomass is first converted into an energy dense intermediate by fast pyrolysis in regionally distributed plants. Then, this liquid-like intermediate (biosyncrude) is transported to a large-scale industrial plant for further conversion to synthesis gas by entrained flow gasification and, subsequently, into synthetic fuels. The bioliq pilot plant erected on site of KIT demonstrates this concept in TRL6 producing synthetic gasoline via methanol and dimethyl ether (DME) synthesis.

During pyrolysis, gasification, and synthesis, carbon dioxide (CO_2) is by-produced, which to large extend is separated after gasification and during synthesis. So far, this CO_2 potentially recovered from the expected large industrial plants at relatively low cost is not intended for use today. CO_2 formed during fast pyrolysis, in perspective, could also be recovered and used, particularly when autothermal processes are applied. However, this is not considered in the exemplary study made below. Here, at the example of methanol and methanol-to-gasoline synthesis via biosyncrude gasification, the mass and fuel energy efficiency for the original bioliq process is compared to possible CCU cases when CO_2 from synthesis and from gasification are utilized. That way, maximization of the carbon efficiency is achieved, transferring practically all carbon from the biomass feedstock into the fuels produced.

PROCESS DESCRIPTION

A block flow diagram of the bioliq concept is shown in Figure C.1 together with the emission points for possible CO_2 recovery.

During bioliq fast pyrolysis, lignocellulosic biomass is converted into liquid bio-oil, solid char and ash as well as non-condensable gases, which is mainly composed of CO_2 (50 vol.%) and CO (30 – 40 vol.%) besides small amounts of hydrocarbon gases and hydrogen.

Entrained flow gasification within the bioliq process is conducted at elevated pressures up to 80 bar, utilizing pure oxygen and steam as gasification agents. A typical synthesis gas mixture obtained at the > 1000 °C reaction temperature contains nearly equimolar amounts of 0.35 and 0.36 vol.% for H_2 and CO, respectively, as well as around 18 vol.% CO_2 plus water vapour and few other gas components. Per kg of dry biomass around 1.4 kg of syngas are produced.

Gasoline synthesis within the bioliq pilot plant is carried out via methanol and DME as intermediates, which are produced in a reactor with mixed catalysts. Without additional make-up hydrogen, around half of the CO_2 contained in the syngas need to be converted to hydrogen via the water gas shift reaction also occurring in the methanol/DME reactor, by-producing CO_2 to significant extent of around 0.65 kg CO_2/kg hydrocarbon products.

Table C.1: Technology profile for the bioliq biosyncrude gasification concept

Designation	
Developer/Promotor	Karlsruhe Institute of Technology (KIT)
Conversion technology	Fast pyrolysis, entrained flow gasification, methanol, DME and MtG synthesis
Feeding System	n.a.
Principal Feedstock(s)	Lignocellulosic residues from forestry, agricultural and land management
Principal Application(s)	Biochar and tar and nitrogen free syngas for production of methanol, H_2, SNG
Scale	Pyrolysis 2 MW_{th}, gasification 5 MW_{th}, synthesis < 2 MW_{th}
Development Status	Technical testing in pilot scale

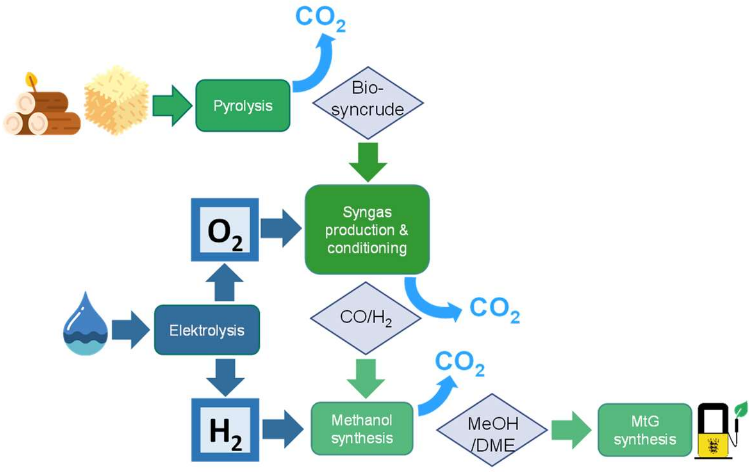

Figure C.1: Process steps of the bioliq process with CO_2 emission points for CCU.

Table C.2: Exemplary estimation of CO_2 recovery, specific hydrogen consumption and fuel yields for a hydrogen powered bioliq process layout related to a 100 MW_{th} syngas production unit.

Process (step)	Feedstock	CO_2 released	H_2 consumption	Fuel yield
bioliq unmodified	34.2 t/h		-	7.5 t/h
Fast pyrolysis	135.1 MW	5.4 t/h	1.9 t/h	1.7 t/h
Gasification	100 MW	10.2 t/h	3.6 t/h	3.2 t/h
Methanol synthesis	72.2 MW	4.8 t/h	7.1 t/h	1.5 t/h

The amounts of CO_2 released during the unmodified process on the hypothetic basis of a 100 MW_{th} gasification system, the specific hydrogen consumption for conversion to hydrocarbon fuels, and additional fuel yield are compiled in Table C.2. These data were obtained from simulations of the bioliq process. Possible yields were based on stoichiometric reaction equations and a specific hydrogen consumption for hydrocarbon production from CO_2 via methanol [1]. The rough estimated shows that essentially a doubling of the fuel yield by practical complete conversion of the biomass carbon can be achieved, in perspective. Other studies estimate similar effects or promise even higher yields [1].

Essentially, two options could be realised for hydrogen integration in the H_2-boosted bioliq process. CO_2 separation and subsequent use seems favourable for CO_2 recovery from gasification when a conventional methanol or DME synthesis is applied for syngas conversion, because the catalysts used only accept CO_2 contents of a few percent in volume. In this case the CO_2, by-produced during methanol/DME synthesis, would be converted directly in the synthesis reactor by adding make-up hydrogen to the required extent. Alternatively, CO_2 generated in the gasification process could be completely transferred to the synthesis after gasification, but requiring catalysts with high CO_2 tolerance then. In both cases, recirculation of unconverted syngas becomes necessary due to the limited single-pass conversion of the methanol synthesis. This will consequently result in a not complete conversion of the CO_2, for technical reasons (e.g. losses by the purge gas streams) and economic considerations along the trade-off of carbon yield maximization and increasing expenses for hydrogen.

DEVELOPMENT STATUS, APPLICATIONS, AND PRODUCTION SCALE

The bioliq project was concluded 2024 [3]. Construction of the fast pyrolysis and mixing plant for biosyncrude production began in 2006 and was completed in 2008. The summarized results and operating experience utilizing straw as feedstock can be found in [4] and for the last two campaigns with Miscanthus feedstock in [5]. Applications for funding and approval for consecutive construction followed in 2008 and 2009, with construction of the plant largely taking place in parallel. After completion, the individual plants were initially commissioned and tested independently of each other and then operated in combination [6]. Either the pyrolysis plant or the gasification with gas purification and synthesis was operated, occasionally the gasifier was also operated in stand-alone mode [7]. The highlight of the plant's operation was the production of several tons of fuel as part of the reFuels project at KIT, which was funded by the Baden-Württemberg Ministry of Transport. The so produced fuels, also used in blends with up to 86 % of CO_2 emission reduction, were successfully tested by various partners in test engines and in vehicles with real driving emission tests. Different own and other studies on costs for the production of synfuels by the bioliq process were compared in a meta-study, giving evidence of the economy-of-scale for this process and confirming the range of costs for synthetic biofuels [8]. Research and development on the processes will continue as part of KIT's Energy Lab in the submodule 'Carbon Cycle Lab' alongside the Power-to-Fuels Lab and the Hydrogen Integration Platform, among others [9].

ASSESSMENT OF THE TECHNOLOGY READINESS

An assessment of the technology readiness of the H_2 boosted bioliq concept is provided in the table below.

Table C.3: Assessment of technology readiness for the H_2 boosted bioliq concept

Process steps	TRL	Weight [%]	Comments
Fast pyrolysis	6-7	20	For wood as a feedstock, TRL is already 9 and plants are in commercial operation today. For biomass residues with high ash content and other complex composition, development is ongoing at TRL ≤ 6.
Entrained flow gasification	6-7	40	Gasification of biomass by other technologies (fixed bed, fluid bed, dual fluid bed) is in commercial operation from small scale up to several 10 MW. For high pressure entrained flow gasification development pilot plants are available but further scale-up is necessary.
Methanol and gasoline synthesis	9	20	MtG process technology is commercially available as is the DME synthesis. Both processes are licensed by various technology providers. The MtG process realized at the bioliq pilot plant proceeds via one stage methanol/DME synthesis to benefit from the CO/H_2 ratio after biomass gasification of around unity. Methanol plants increasingly are developed to make use of CO_2 rich feed gases.
Integrated operation	6-7	10	From the begin, there was emphasis on integrated operation along the complete bioliq process chain.
Hydrogen integration	6-7	10	Hydrogen integration is not yet implemented at the bioliq plant. Actually, CO_2 separation

Process steps	TRL	Weight [%]	Comments
			prior to synthesis to reduce its content to around 1 vol.%. For H_2 integration, i) the recovered CO2 is converted in a dedicated plant for hydrocarbon production via the reverse-water-gas-shift reaction or co-electrolysis (both already available at the EnergyLab 2.0 at KIT) or ii) the methanol/DME synthesis needs to be modified by CO2-tolerant catalysts.
Overall “Weighted Average”	7		
Overall “Weakest Link”	6-7		Integrated operation to be proven

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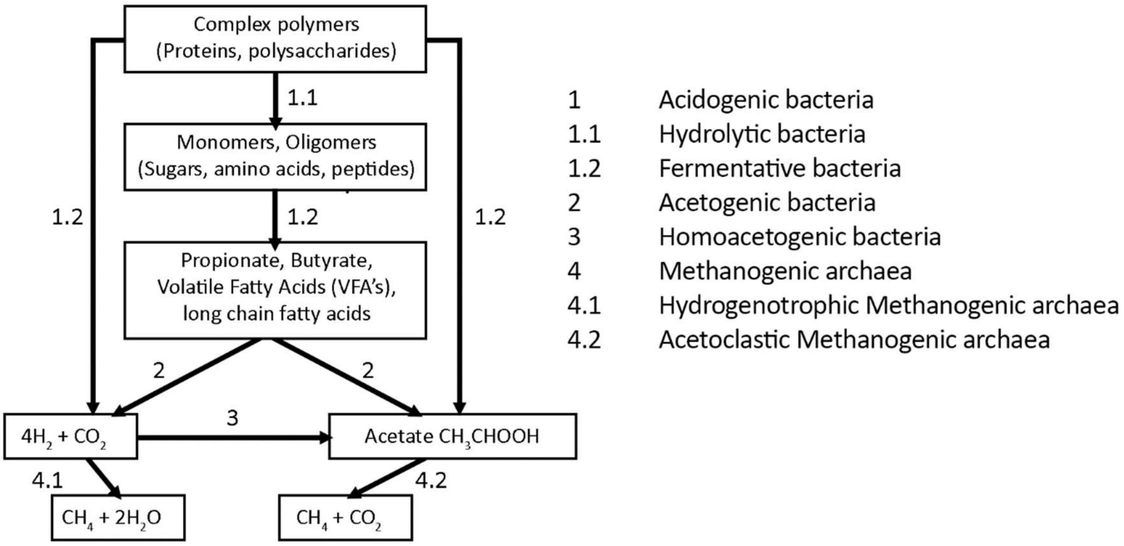
Appendix D – Biological Methanation: Electrochaea

BACKGROUND

The anaerobic process is well documented and may be greatly simplified as in Figure D.1. In brief wet organic feedstock is hydrolysed to sugars (by hydrolytic bacteria) which in turn are fermented to volatile fatty acids (VFAs) by fermentative bacteria. These volatile fatty acids may be further converted by acetogenic bacteria to acetate (CH_3COOH), hydrogen, and carbon dioxide. If for example, the breakdown of ethanol (CH_3CH_2OH) as a representative volatile fatty acid is assessed, species 2 the acetogenic bacteria produce hydrogen and carbon dioxide while the hydrogenotrophic methanogenic archaea (species 4.1) use hydrogen and carbon dioxide as ingredients to produce methane within a favourable exothermic reaction. For biomethanation we are most interested in species 4.1 (the hydrogenotrophic methanogenic archaea).

PROCESS CONFIGURATIONS

Several configurations may be used to employ a power to methane process at a biogas facility. In Figure D.2-A hydrogen from an electrolyser is added to an existing anaerobic reactor. This would be considered ***an in-situ biological methanation*** process. With reference to figure 1, acetoclastic methanogenic archaea would continue to produce CO_2 and minimise potential to reach a biomethane standard (with methane content more than 97%). Another issue is the rise in the hydrogen partial pressure within the digester. In assessing the stoichiometry associated with species 2 (in Figure D.1) it may be noted that the Gibbs Free Energy is positive and as such the reaction is not favourable. This may be explained with



			ΔG (kJ/reaction)
Species 2	$CH_3CH_2OH + H_2O$	$= CH_3COO^- + H^+ + 2H_2$	+5.95
Species 4.1	$2H_2 + 0.5CO_2$	$= 0.5CH_4 + H_2O$	−65.45
Species 4.2	$CH_3COO^- + H^+$	$= CH_4 + CO_2$	−28.35
Net	CH_3CH_2OH	$= 1.5CH_4 + 0.5CO_2$	−87.85

Figure D.1: Four trophic groups involved in anaerobic digestion (adapted from Persson et al., 2014) reference to the hydrogen partial pressure. The reason the anaerobic process works, is the syntrophic mutualism that exists between acetogenic bacteria (species 2) and hydrogenotrophic methanogenic archaea (species 4.1). Hydrogenotrophic methanogenic archaea have a high affinity for hydrogen as shown by the negative Gibbs Free Energy and as such the archaea efficiently remove the hydrogen produced by the acetogenic bacteria (Figure D.1) which reduces hydrogen partial pressure and facilitates the reaction. As such this configuration would need to be followed by a traditional biogas upgrading step to achieve a biomethane standard.

In Figure D.2-B the biogas and the hydrogen are reacted together in an ***ex-situ biological methanation*** process. In theory this configuration can replace traditional biogas upgrading processes and as such act as a biogas upgrading process. This is a huge benefit as the CAPEX and OPEX of traditional physio-chemical biogas upgrading can be avoided thus reducing the overall costs substantially. Of potential issue, however, is that the methane that accompanies the CO_2 within the biogas could impact negatively on the efficiency of the process due to the impact of the presence of methane on the partial pressure of the other gases.

In Figure D.2-C biogas is upgraded to biomethane, and the concentrated form of CO_2 is reacted with the hydrogen to produce a second stream of synthetic methane. This can be a more efficient CO_2 conversion process, however a negative aspect of this, is that the developer must also invest in traditional biogas upgrading, raising the costs of the process.

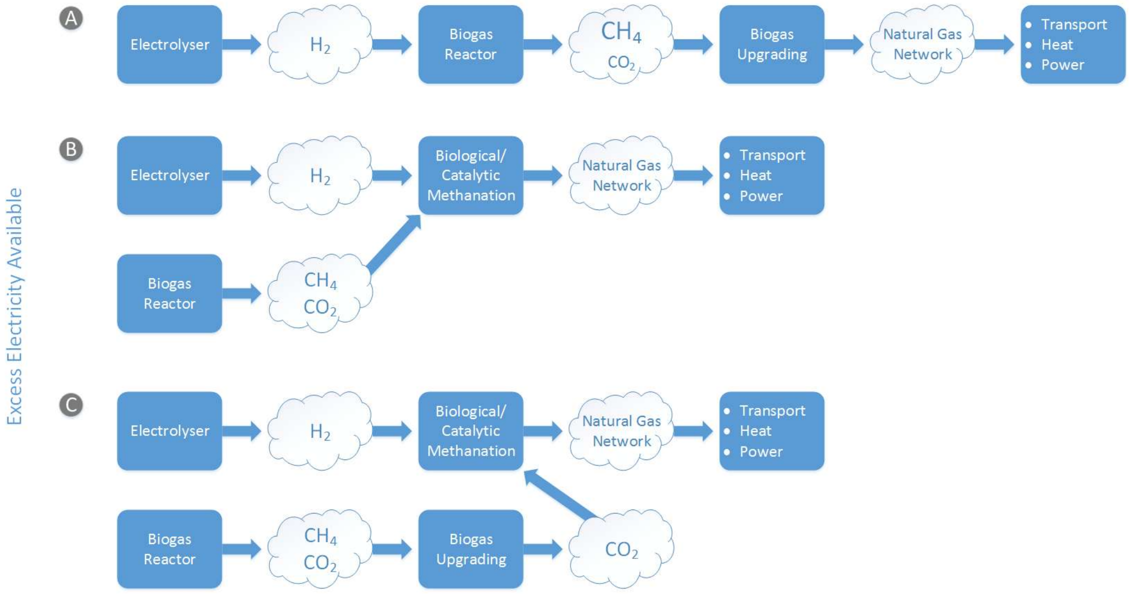


Figure D.2: Different pathways that can be used to integrate an electrolyser at a biogas facility (from Persson et al., 2014)

DESCRIPTION OF ELECTROCHAEA-BIOCAT

The Electrochaea-BioCat demonstrator (Table D.1 and Figure D.3) began operation at a wastewater treatment plant in Avedøre, Denmark in April 2016; it operated in excess of 4500 hours, until October 2019. This demonstration plant included a 1 MW electrolyser (Hydrogenics' S1000 alkaline electrolyser) integrated with a biological methanation facility (operating at 10 bar and 62ºC) with CO_2 sourced from a local biogas plant and included the use of raw biogas as well as purified CO_2. The design point of the plant is the utilisation of biogas at a rate of 125 m^3/hour (75m^3/h CH_4 & 50 m^3/h CO_2).

Table D.1: Technology profile for the Electrochaea's BioCat biological methanation facilities

Designation	Torrgas
Developer/Promotor	Electrochaea
Conversion technology	Ex-situ biological methanation
Feeding System	Biogas from wastewater treatment facility Carbon dioxide from biogas upgrading system Hydrogen from electrolyser
Principal Feedstocks	Biogas, carbon dioxide and hydrogen
Principal Applications	Grid quality renewable methane
Scale	1MW electrolyser & 125 m^3/h biogas input 125 m^3/h methane output
Development Status	Technical testing in demonstration scale; pending commercialization

Figure D.3: Photograph of BioCat Plant on BIOFOS site in Avedøre (from Power-to-Gas via Biological Catalysis (P2G-Biocat)

Electrochaea demonstrated the process in two configurations:

- As per Figure 2.C pure streams of CO_2 were added post upgrading of biogas to the hydrogen to produce methane.
- As per Figure 2.B the technology was applied as a method of upgrading biogas, converting the CO_2 within the biogas stream to methane. This, as mentioned previously, is more challenging due to the effect of methane in the biogas influent on the partial pressure of the hydrogen and carbon dioxide gases. It also requires a higher throughput of biogas, as 63% of the biogas (CH_4) is not involved in the process.

Operation under both configurations was shown to offer CO_2 conversion efficiencies of more than 97% and both configurations produced a similar gas output, which was injected into the local gas distribution grid.

With reference to Table D.2, a stoichiometric ratio of 4:1 (H_2:CO_2) led to a design with a 1MW electrolyser producing 200 m^3/h of hydrogen to react with 50 m^3/h of CO_2 to produce 50 m^3 of CH_4. The methane end product increased in yield by 67%. Implementing a circular bioeconomy approach, the heat from the exothermic reaction, both from biomethanation and from the electrolyser, was recycled to be used in the wastewater treatment plant.

Table D.2: Estimate of input and output of Electrochaea-BioCat power to methane facility at Avedøre, Denmark

	Alkaline Electrolysis Cell Electrolyser	Methanation reactor
Theory	$2H_2O = 2H_2 + O_2$ H_2 LHV of 3kWh/m^3 or 33.33kWh/kg	$4H_2 + CO_2 = CH_4 + 2H_2O$
Input	1 MW electrolyser using 1000 kWh of electricity per hour	125 m^3 raw biogas/h 75 m^3 CH_4/h & 50 m^3 CO_2/h
Output	1000 kW$_e$h at 60% capacity = 600 kWh = 200 m^3 H_2 /h	75 m^3 CH_4/h from biogas plus 50 m^3 CH_4/h from biological methanation facility

DEVELOPMENT STATUS, APPLICATIONS, AND PRODUCTION SCALE

Electrochaea facilities such as in Avedøre, Denmark and in Switzerland have demonstrated flexibility with load factor tests demonstrating operation between 0 and 100 % capacity. The technology readiness level is assessed in the range 8 to 9 (Table D.3). These facilities have shown rapid recovery after periods of shutdown; such intermittent operation is of great benefit in matching operation of an electrolyser with electricity sourced from variable renewable electricity. Electrochaea see development of facilities in scale as per Table D.4.

DEVELOPER FEEDBACK

Maria Beschid responded to a request for feedback on the assessments with corrections and clarifications on various aspects…

Table D.3: Assessment of technology readiness for the Electrochaea's BioCat biological methanation facility

Process steps	TRL	Weight [%]	Comments
Electrolyser	8-9	50	1 MW electrolyser (Hydrogenics' S1000 alkaline electrolyser) It is expected with the growth of the hydrogen economy that efficiencies and start-stop operations of electrolysers will improve and the costs will reduce
Biological methanation	8-9	50	The biological methanation is a young technology that has been demonstrated at a number of facilities. It can operate as an upgrading mechanism for biogas (with CH_4, CO_2 and H_2 mixed in the same reactor) or simply as a means of reacting CO_2 (from biogas upgrading) with hydrogen to make methane. Operation under both configurations was shown to offer CO_2 conversion efficiencies of more than 97% and both configurations produced a similar gas output, which was injected into the local gas distribution grid.

Table D.4: Future scale of Electrochaea's BioCat biological methanation facilities (from Electrochaea GmbH data sheets)

	BioCat 1	BioCat 10	BioCat 50
Electrolyser	1MW$_e$	10 MW$_e$	50 MW$_e$
Reactor Temperature and Pressure of 63°C at 10 bar with CO_2 conversion efficiency of greater than 97%			
Gas Input	200 m^3/h H_2 50 m^3/h CO_2	2000 m^3/h H_2 500 m^3/h CO_2	10,000 m^3/h H_2 2500 m^3/h CO_2
Installed power at methanation	45 kW	370 kW	1600 kW
Outputs Grid quality gas Thermal energy Metabolic water	 50 m^3/h CH_4 130 kW$_{th}$ 80 L/h	 500 m^3/h CH_4 1275 kWth 800 L/h	 2500 m^3/h CH_4 6400 kWth 4000 L/h
Footprint	150 m^2	480 m^2	1070 m^2

Appendix E - Pilot-SBG | Methanation of electrolytic hydrogen in biogas plants for renewable methane

BACKGROUND

The Pilot-SBG project (SBG = synthetic biogas) is a research and demonstration project for the production of renewable methane (e.g. as Bio-LNG) from previously unused biogenic residues, by-products and waste. Key market is the use as a climate-friendly fuel in the transport sector. To this end, the DBFZ planned a full-scale pilot plant, which is now under commissioning for previously unused biogenic residues, by-products and waste into renewable methane. The core processes of the plant are anaerobic digestion followed by catalytic methanation. The anaerobic digestion is carried out in two lines to compare the performance of different process parameters. The main objective is to increase the methane yield using carbon dioxide from the biogas process and externally supplied electrolytic hydrogen. For the pre- and post-treatment of substrates and digestates, hydrothermal processes (HTP) and a separation cascade will be used to increase the product portfolio of the whole plant by

Table E.1: Technology profile for the Pilot-SBG

Designation	Renewable methane
Developer/Promotor	On R&D scale: DBFZ Deutsches Biomasseforschungszentrum
Conversion technology	Anaerobic digestion with subsequent catalytic methanation
Feeding System	Feeding system: System consisting of various pumps (e.g. hopper pump, eccentric screw pump, screw conveyor) at several positions in the pilot plant
Principal Feedstock(s)	Urban and agricultural biogenic residues, by-products and waste related to REDII Annex IX A
Principal Application(s)	Methane, Fermentation residues (hydrochar and fertiliser)
Scale	Pilot scale: 0,1 m³ methane per hour (Concepts for commercial scale: 1,500 m³ methane per hour)
Development Status	Technical testing in pilot scale

separating valuable by-products. Accompanying studies include a digital twin of the pilot plant to create commercial scale plant concepts.

This plant addresses important issues such as the cascade use of biomass, the circular economy, sector coupling, CO_2 reduction through the power-to-X process of biogas methanation, and decentralised and regional energy generation.

PROCESS DESCRIPTION

A process scheme of the Pilot-SBG concept is shown in Figure E.1.

The pilot plant combines the anaerobic fermentation of selected residual and waste materials with a subsequent methanisation of the biogas from the fermentation. Innovative upstream and downstream processes increase resource efficiency and provide valuable by-products.

The process chain essentially consists of the specific preparation of the feedstock, hydrothermal processing as an optional second stage of feedstock pre-treatment, anaerobic fermentation to biogas in two parallel fermenter lines of different designs, catalytic methanation of the carbon dioxide contained in the biogas with hydrogen and treatment of the fermentation residues. As a result, renewable methane is produced as the main product at the pilot plant. At least, the methane will then be purified and compressed.

The process steps described were each integrated into the pilot plant as a separate module in order to increase the process options on the one hand and to be able to integrate new technical findings into the plant at short notice on the other. Accordingly, the pilot plant is a technology platform for R&D with the following modules:

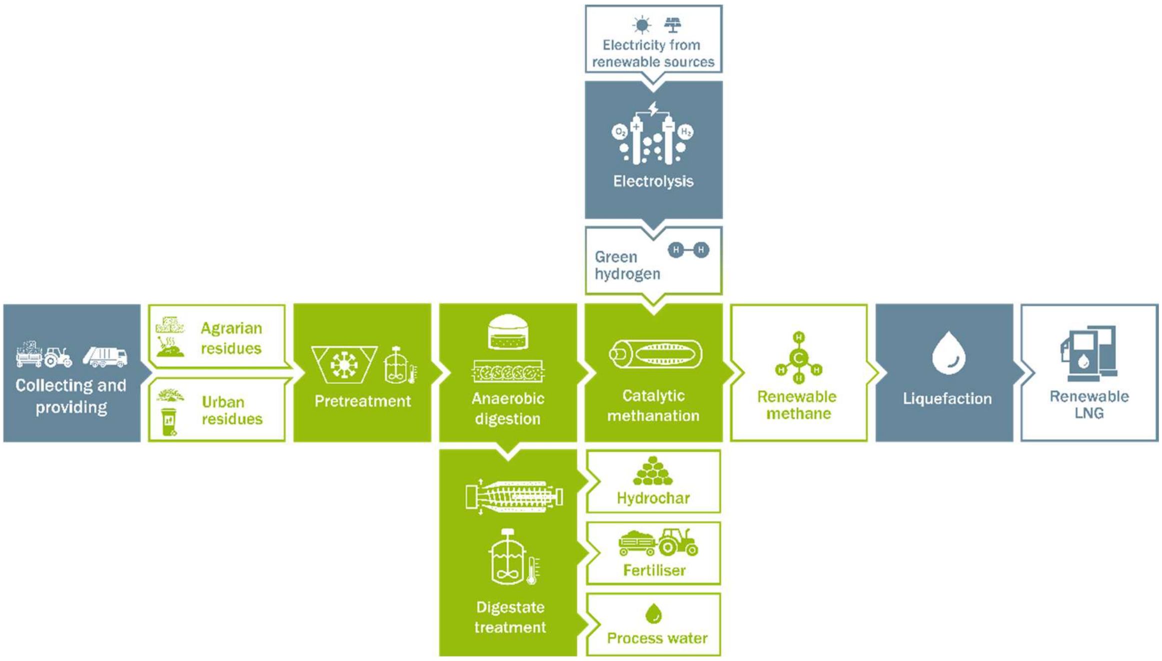


Figure E.1: Process scheme of the Pilot-SBG concept for production of renewable methane.

Substrate preparation module

The biomass is prepared in the substrate pre-treatment module. The primary objective is to adjust the particle size and fibre length of the substrates (less than 20 mm), as well as the dry matter content of the substrate mixes, so that the plant operation is as unobstructed as possible. The different characteristics of different biomasses - such as cereal straw, green cuttings or biowaste - require different processing steps for coarse, fine, dry and wet shredding. This module does not consist of a single unit, but of several aggregates operated in batches, a central weighing station and areas for filling, transferring and mixing the substrates. The aim is to provide storable and transportable batches for subsequent process steps.

Anaerobic fermentation module

In the anaerobic fermentation module, the processed biomass is broken down by microorganisms in a controlled process. Part of the organic matter is converted into biogas, a mixture of mainly methane and carbon dioxide. Two separate reactor lines are available for this purpose, which differ in the first stage of the process. One line has a continuously stirred tank reactor, such as is commonly used in agricultural applications. The second reactor line uses a plug flow reactor at this stage, which is widely used in the treatment of organic waste. Common to both process lines is that the second and third process stages are each designed as continuous stirred tank reactors. This allows a direct comparison of the most common reactor systems. The methane yield across the process stages can be assessed as a function of varying process parameters.

Methanation and gas purification module

In the methanation module, the biogas from the two digesters is mixed with externally supplied hydrogen without separating the methane. In the temperature-controlled tubular reactor, carbon dioxide from the biogas reacts with hydrogen over a catalyst to form methane and water. After the water is separated in a condensate trap, a product gas is produced with a methane content almost twice that of the biogas. On approx. 500 ml catalyst and at a process pressure of 20 bar(a), 2.5 m^3/d biogas with a methane content of approx. 50 vol% is converted into approx. 2.47 m^3/d synthesis product with more than 95 vol% methane content. To ensure that the product gas is suitable as a fuel, unreacted residual carbon dioxide can be removed by gas scrubbing.

Digestate treatment module

In the digestate treatment module, the part of the biomass that has not been fermented into biogas is treated to produce dischargeable water, a solid organic farm fertiliser rich in nitrogen and phosphorus, and a liquid inorganic farm fertiliser rich in nitrogen and potassium. In addition, a recirculation of the liquid phase of the digestate can be provided for mixing into the biogas process. The first step in digestate treatment is to separate the solid phase from the liquid phase using a press screw separator and decanter centrifuge or chamber filter press with the addition of flocculants. The membrane technologies of ultrafiltration and reverse osmosis are available for further treatment or concentration of the liquid phase. The separation cascade can process an average of 0.1 to 0.5 m^3 of digestate per day with the aim of achieving a volume reduction of at least 50 %. The quantity and quality of the recovered nutrients or farm fertiliser depends on the biomass used and the process conditions in the anaerobic digestion. During processing, all relevant process parameters are recorded to monitor the separation processes and to determine mass and energy balances.

Hydrothermal processing module

In the pilot plant context, the hydrothermal processing module covers processes for the digestion, treatment or upgrading of biomass and residual materials, particularly those with a high water content, at elevated pressure and temperature. The processes typically run at temperatures of 150 to 280 °C and pressures of about 5 to 65 bar for 30 minutes to several hours. Part of the module is a 0.5 m^3 reactor that can operate up to 240 °C and 39 bar. This will be used in the lower temperature and residence time range to pre-treat difficult to access biomass for fermentation. For digestate treatment, the same technology can be used at higher temperatures and pressures to convert aqueous digestate to hydrocarbons and to move nutrients between the aqueous and solid phases. Taking phosphorus as an example, the portion bound in the solid phase can be transferred to the liquid phase, making it accessible for recovery in downstream processes via a precipitation reaction.

DEVELOPMENT STATUS, APPLICATIONS, AND PRODUCTION SCALE

On pilot scale, the entire process chain is represented in one plant, which is unique in Germany. The pilot plant is used, for example, to gain insights into the interaction of components and equipment or the influence of process parameters on stability and yield. Its capacity has been selected to provide sufficient flexibility for research and development and to be cost-justifiable. After the operational campaigns planned from 2024 onwards, the plant will be available as a technology platform for research and development to players from science and industry alike, and is designed to allow the flexible use of different technology modules. To a limited extent, it can already be used for these purposes during the Pilot-SBG project.

Production scale of the pilot plant

An integrated and highly automated pilot plant, which produces 2.5 m^3 of renewable methane per day, is operated on a test area of around 1000 m^2. The process is fed with 20 to 30 kg of biogenic residues or waste and 7.3 m^3 of green hydrogen every day.

Production scale of a possible commercial plant

The biorefinery was assumed to operate for 8,000 hours per year. The electricity required for the individual process steps is provided by solar and wind energy from the grid. A woodchip-fuelled steam generator generates process heat of up to 500 °C. The processing capacity of the upscaled biorefinery is 51,520 tonnes feedstock per year, which is provided by a 1:1 (w/w) ratio of straw and liquid manure. This capacity corresponds to a biogas plant with an electrical output of around 2.5 MW. The dimensioning is based on the spatial availability and distribution of the raw materials in the selected region of Central Germany and an economic assessment of the reasonable transport routes. Based on the fact that liquid manure has a lower energy content per weight than straw, an average transport distance of up to 50 km for straw and up to 20 km for liquid manure was assumed. Within this availability radius, there is a shortage of available slurry. However, to maximise the potential of straw, a real mixing ratio of 1:1 is assumed.

ASSESSMENT OF THE TECHNOLOGY READINESS

An assessment of the technology readiness of the Pilot-SBG concept is provided in Table E.2.

The technical assessment of a commercial scale plant is described in [1].

Table E.2: Assessment of technology readiness for the Pilot-SBG concept

Process steps	TRL	Weight [%]	Comments
Substrate preparation module	9	-	- Optimisation of the mechanical substrate preparation to the plant concept
Anaerobic fermentation module	9	10	- Increase of the organic volume load - Analysis of initial, intermediate and final products - Treatment of the biomass stream after the main fermenter with ultrasonic disintegration to increase methane yields in the post-digester
Methanation and gas purification module	8 to 9	30	- Parameter screening of the different catalyst types - Variation of the recirculation rate - Determination of the dependencies of methane yield, throughput and residual hydrogen content - Identification of economies of scale - Adaptation of product gas purification to the resulting process requirements - Lifetime prediction and support for catalyst selection for continuous commercial operation
Digestate treatment	4 to 9	30	- Needs-based recycling of partially processed fermentation residues to

Process steps	TRL	Weight [%]	Comments
module			minimise external nutrient requirements • Increase in methane yield and reduction of fermentation residues • Minimising the water content of fermentation residues to be spread and producing nutrient products that are as defined as possible • Eliminate the use of polymers as flocculants by using alternative substances
Hydrothermal processing module	5 to 8	30	• Test of the hydrothermal reactor with high dry matter content (25% dry matter) • Quantification and evaluation of economies of scale • Expansion of the substrate base through optimized hydrothermal pre-treatment • Establishing added value from urban fermentation residues
Overall “Weighted Average”	6 to 8		
Overall “Weakest Link”	4		• Integrated operation to be proven

ECONOMIC AND ENVIRONMENTAL ASSESSMENT

Economic and environmental assessment of the Pilot-SBG-concept is described for the use of agricultural feedstock in [1]. The resulting mass and energy balance for the concept is shown in Figure E.2 and the resulting GHG balance for the concept is shown in Figure E.3. The cost distribution and sensitivity analysis for the concept is shown in Figure E.4.

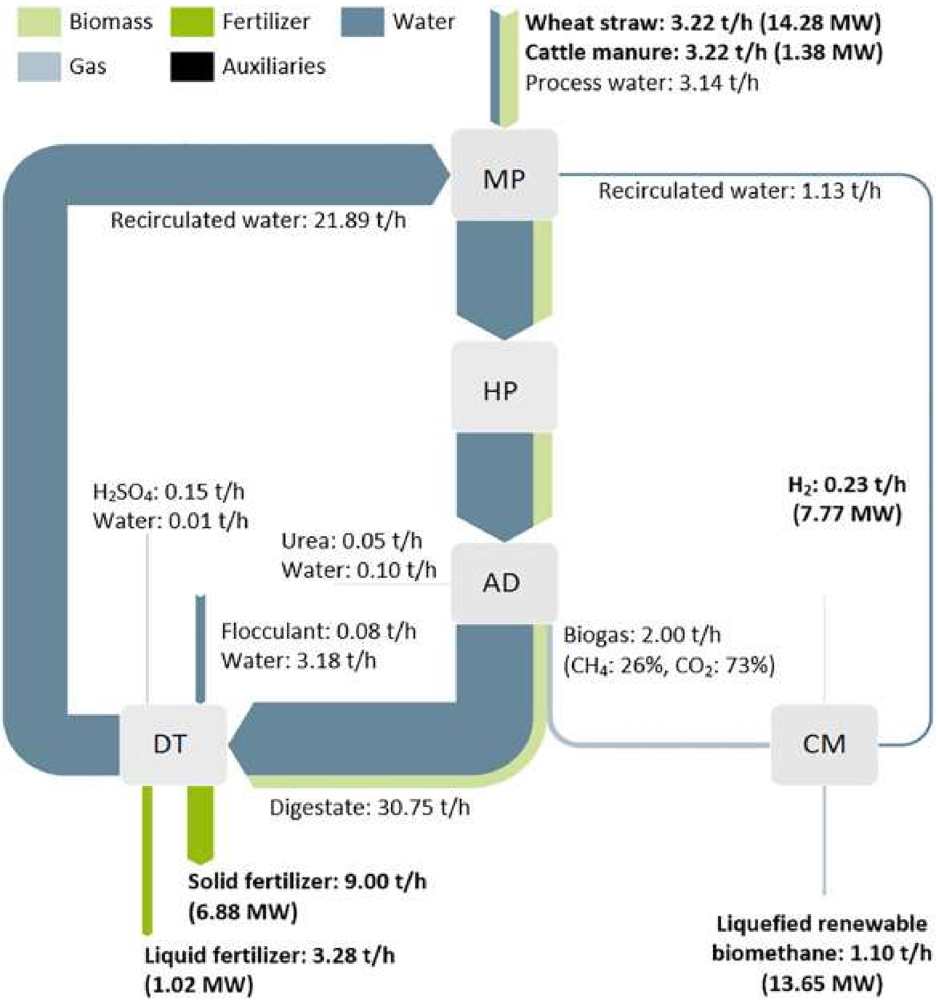


Figure E.2: Mass and energy balance of the commercial size agricultural Pilot-SBG concept (MP: mechanical substrate preparation, HP: hydrothermal pre-treatment, AD: anaerobic digestion, CM: catalytic methanation, ML: methane liquefaction, DT: digestate treatment); source: [1]

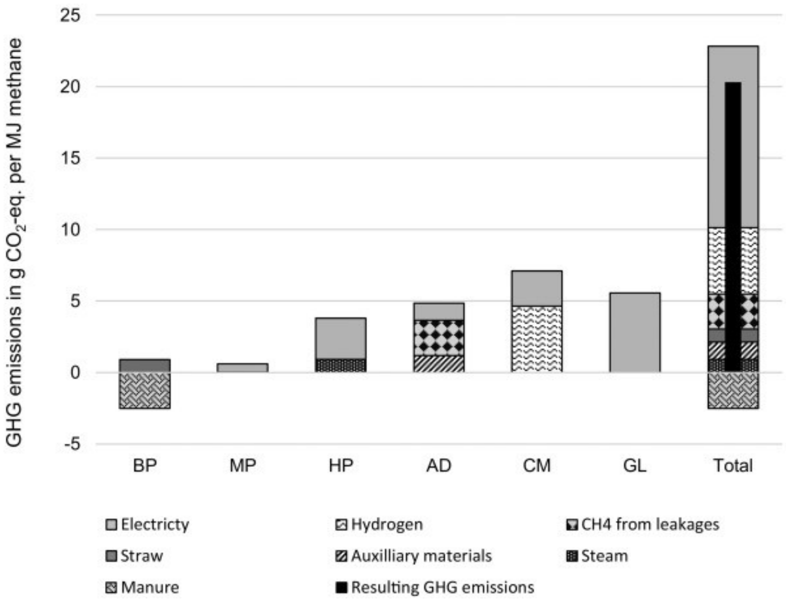


Figure E.3: Specific GHG emissions of the different processes (BP: biomass provision, MP: mechanical substrate preparation, HP: hydrothermal pre-treatment, AD: anaerobic digestion, CM: catalytic methanation, DT: digestate treatment, GL: gas liquefaction) and overall GHG emissions for the commercial size agricultural Pilot-SBG concept; source:[1]

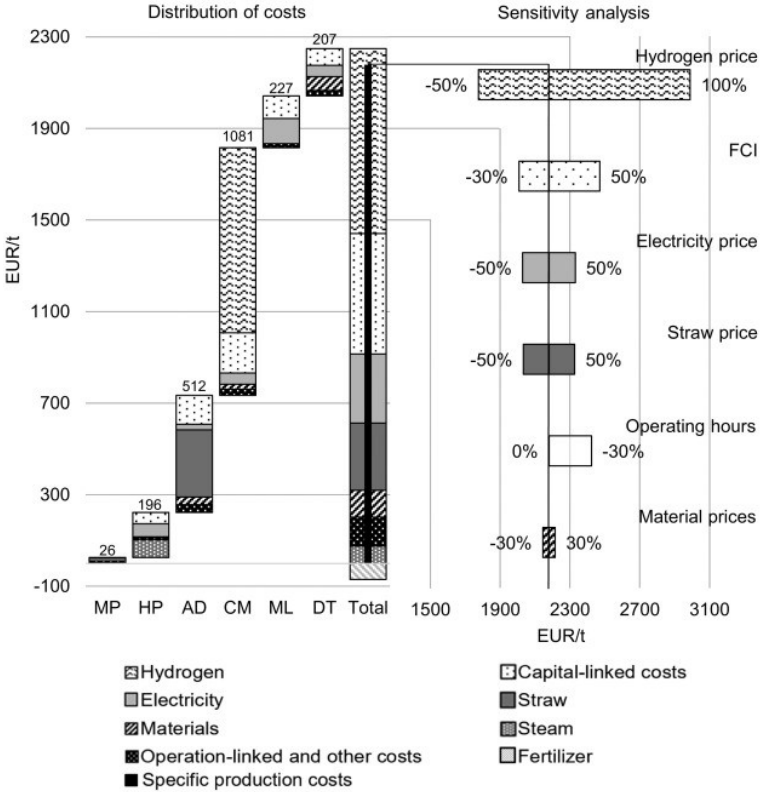


Figure E.4: Waterfall chart for the cost distribution and tornado chart for the sensitivity analysis of specific production costs of the commercial size agricultural Pilot-SBG concept (MP: mechanical substrate preparation, HP: hydrothermal pre-treatment, AD: anaerobic digestion, CM: catalytic methanation, DT: digestate treatment, GL: gas liquefaction); source: [1]

DEVELOPER FEEDBACK

Further information can be found on the website https://www.dbfz.de/en/projects/pilot-sbg/start.

REFERENCES

[1] H. Etzold et al. (2023) Technical design, economic and environmental assessment of a biorefinery concept for the integration of biomethane and hydrogen into the transport sector. doi.org/10.1016/j.biteb.2023.101476

Appendix F - HYFUELUP (Hybrid Biomethane Production from Integrated Biomass Conversion)

BACKGROUND

Within the European Union project HyFuelUp (co-financed by the Swiss SERI), an existing biomass gasifier with CHP in Tondela/Portugal will be refurbished and augmented by a methane synthesis to produce around 500 kW renewable Bio-LNG for the use of vehicle fleets. Start of operation is planned in 2026.

Low grade biomass will be gasified at temperatures between 600 and 850°C whereby the addition of limestone will allow the *in situ* sorption of CO_2. The product gas from the gasification is rich in hydrogen, CO, and methane. After gas cleaning, a fluidised bed methanation that is able to convert also unsaturated hydrocarbons will produce a methane-rich gas, whose methane content can even be increased by the flexible addition of hydrogen from electrolysis. Remaining CO_2 will be separated and unreacted hydrogen will be recycled to the process with the aid of membranes. A final liquefaction step will enable the production of Bio-LNG (also referred to as LBG).

PROCESS DESCRIPTION

Table F.1: Technology profile for the HyFuelUp project

Designation	HyFuelUp
Developer/Promotor	Circle Molecule, LNEG, IPP Portalegre
Conversion Technology	Sorption enhanced gasification process (Host Eng., BIOREF, Uni Stuttgart)
Synthesis Technology	Catalytic fluidised bed methanation (AlphaSYNT, PSI)
Principal Feedstock(s)	Digestates and biomass wastes; flexible H2 addition to methanation
Principal Application(s)	Production of bio-LNG (also referred to as LBG)
Scale	2 MW gasifier, slipstream for 500 kW methane output (LHV basis)
Development Status	Gasifier with sorption enhancement TRL 7, methane synthesis TRL 7

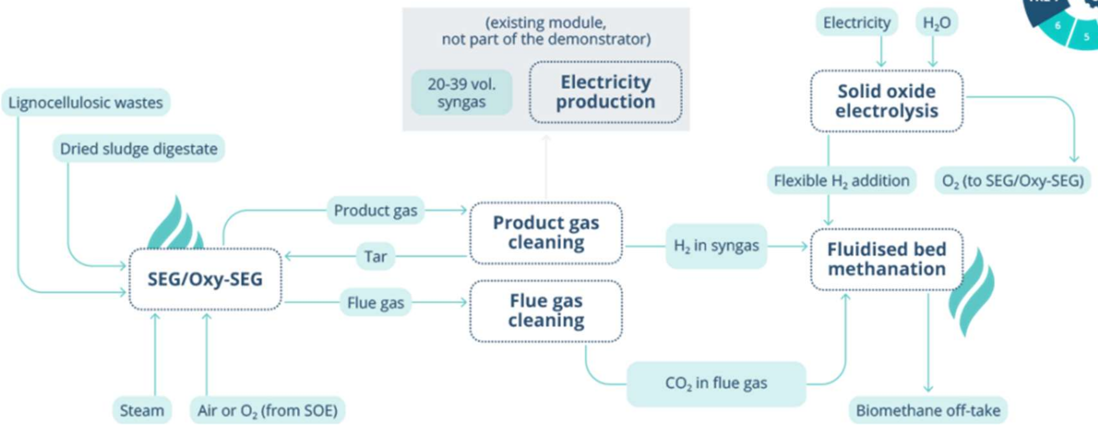


Figure F.1 Process scheme for production of 500 kW of renewable methane [1].

A process scheme of the HyFuelUp process concept is shown in Figure F.1. Biomass wastes and calcined limestone are fed to a circulating bed gasifier in which the sorption enhanced gasification takes place at temperatures between 600°C and 700°C. For operation without limestone, 850°C are possible. The gas cleaning comprises a tar cracker, condensation, active carbon based sorption and metal oxides as final gas cleaning step.

After compression to a few bar, the gas is fed to a fluidised bed methanation unit, in which nickel-based catalyst particles support synthesis of a methane-rich gas stream at temperatures between 300°C and 400°C. The fluidisation of the catalyst bed allows for nearly isothermal operation, high yields and conversion of unsaturated hydrocarbons to methane without catalyst deactivation. When available from electrolysis, hydrogen can be added in flexible Power-to-Gas option to increase the methane yield (and to store the renewable electricity otherwise not usable). Added hydrogen is converted with CO_2 in the product gas to additional methane with an efficiency of around 80%. As this efficiency is given by thermodynamics, it is independent from the amount of added hydrogen, i.e. ”partial load PtG” is possible.

After a particle filter, condensation, CO_2 removal and recycle of residual hydrogen (membrane based), nearly pure methane is obtained that can be injected into a gas linear liquified as fuel for vehicle fleets.

DEVELOPMENT STATUS, APPLICATIONS, AND PRODUCTION SCALE

While the gasification plant as such is commercial, the application of the sorption enhanced gasification by calcite addition will be tested first time in MW scale (besides a short test within the EU project AER II in 2007/2008).

The catalytic fluidized bed methanation of product gas from a Dual Fluidized bed gasifier has already been demonstrated at 1 MW scale within the European Union project BioSNG in Güssing/Austria in the year 2009. The work of the last years including pilot scale tests expanded the reactor performance by flexible hydrogen addition to integrate an easy option for Power-to-Gas for times when renewable electricity is available but cannot be used otherwise.

ASSESSMENT OF THE TECHNOLOGY READINESS

An assessment of the technology readiness of the HyFuelUp concept is provided in Table F.2.

Table F.2: Assessment of technology readiness for the HyFuelUp process concept

Process steps	TRL	Weight [%]	Comments
Gasification reactor with calcite addition	7	30	The gasification system is proven at pilot scale (several 100 kW input) at University of Stuttgart.
Catalytic fluidised bed methanation	7	30	Catalytic fluidized bed methanation was demonstrated within the EU project BioSNG. The flexible addition of hydrogen to allow for optional PtG is tested in pilot scale TRL 6 (200 kW)
Product gas separation cleaning	8	20	This concerns off-the shelf technology bought from third parties.
Integrated operation	7	20	Integrated operation of gasification and fluidized bed methanation was demonstrated within the EU project BioSNG
Overall “Weighted Average”	7-8		Low-cost biomass shall be converted to a versatile fuel with flexible hydrogen addition from highly efficient solid oxide electrolysis (SOE)
Overall “Weakest Link”	7		Integrated operation to be proven, no heat integration between SOE and methanation

ECONOMIC AND ENVIRONMENTAL ASSESSMENT

Within the European Union project HyFuelUp (co-financed by the Swiss SERI), a thorough techno-economic and life-cycle assessment will be conducted.

REFERENCES

[1] hyfuelup.eu; last accessed October 10th, 2025

Appendix G - Lignin hydrodeoxygenation for alkylphenols and BTX production

BACKGROUND

Lignin hydrodeoxygenation (HDO) process aims to convert a lignin rich stream into high value-added chemicals, such as BTX and alkylphenols. A pure hydrogen stream, which flow depends on the feedstock and on the degree of hydrodeoxygenation desired, has to be mixed with the input lignin stream. The process requires fairly mild temperatures, usually 350°C, and moderately high pressures, ranging from 50 to 100 bar.

The products that yield the highest market value are alkylphenols, the most important of which is cresol (methylphenol); nonetheless byproducts like benzene, toluene, xylene or other cyclic aromatics have a high added value too. Furthermore, by correctly using waste streams, like exploited solid and off-gases, electricity generation is made possible thanks to gas and steam turbines; this feature increases the profitability of the plant, since the electric power produced can be used in an electrolyzer in order to lower the need for makeup hydrogen.

PROCESS DESCRIPTION

Table G.1: Technology profile for the lignin HDO process

Designation	Lignin HDO
Developer/Promotor	still on an experimental stage
Conversion technology	Lignin catalytic hydrodeoxygenation
Principal Feedstock(s)	Lignin from lignin-first lignocellulosic biomass pretreatment
Principal Application(s)	High added-value chemicals production
Scale	≈58 MW_{th} input (LHV basis), ≈9350 kg/h lignin feed
Overall process TRL	5

A simplified process scheme, as a block flow diagram, is shown in Figure G.1. Lignin or a lignin rich stream is used as feedstock in the process: it undergoes, alongside with pure hydrogen gas, a pretreatment step, which mainly consist of heating the streams to the reactor specified temperature (350°C). After said step, solid and gas are mixed and sent in a HDO reactor, which is of the continuous stirred tank type, containing the reaction catalyst (sulfided-NiMo/MgO-La_2O_3).

The main process stream has to be fed to a separation section in order to obtain the desired products and to recover as much hydrogen as possible. The first step of the section is represented by a series of flash tanks, that have the purpose to isolate non condensable gases, and by a solid separator, which removes the unreacted lignin; hence three streams with different states (gas, liquid and solid) are obtained. The gas stream, mainly composed by CH_4 and H_2, is sent to a pression swing adsorption (PSA): this unit operation is able to recover 95% of the hydrogen as an essentially pure gas, so it can be easily recycled after recompression. The other fraction of the gas stream is fed to a gas+steam turbine combination, in order to produce electricity. As far as the liquid stream is concerned, it undergoes a series of two distillation columns: in the first one alkylphenols, the product with the highest added value, are obtained from the bottom, whereas in the second column BTX are separated as byproduct. The unreacted solid phase is sent to a combustion+steam turbine system, hence obtaining electricity and the hot utilities needed for correct plant operation.

The resulting mass balance is illustrated in Figure G.2. A significant lignin fraction is not converted in the hydrodeoxygenation step, whereas the two main products are gas and an aqueous stream containing monomers, further separated in BTX and alkylphenols. It is noteworthy to mention that, regarding the energy balance, both the unreacted solid and the gas residue stream are sent to an energy recovery section, capable of producing 13.2 MW of electric energy, also considering the operating machines' demands. It has been observed that a significant quantity

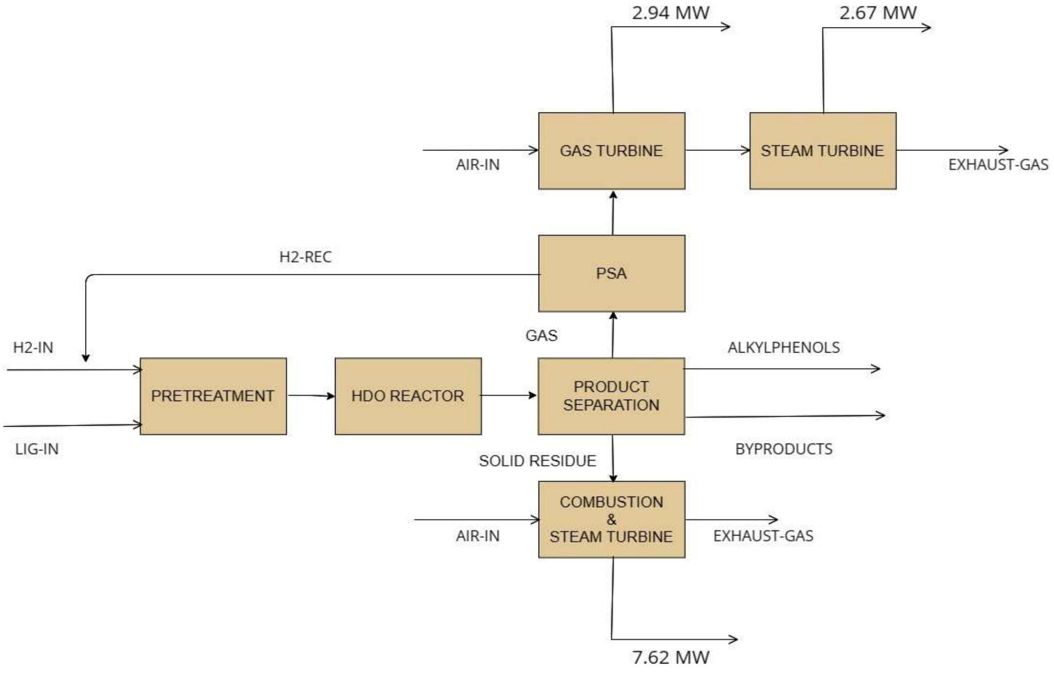


Figure G.1: Lignin HDO block flow diagram

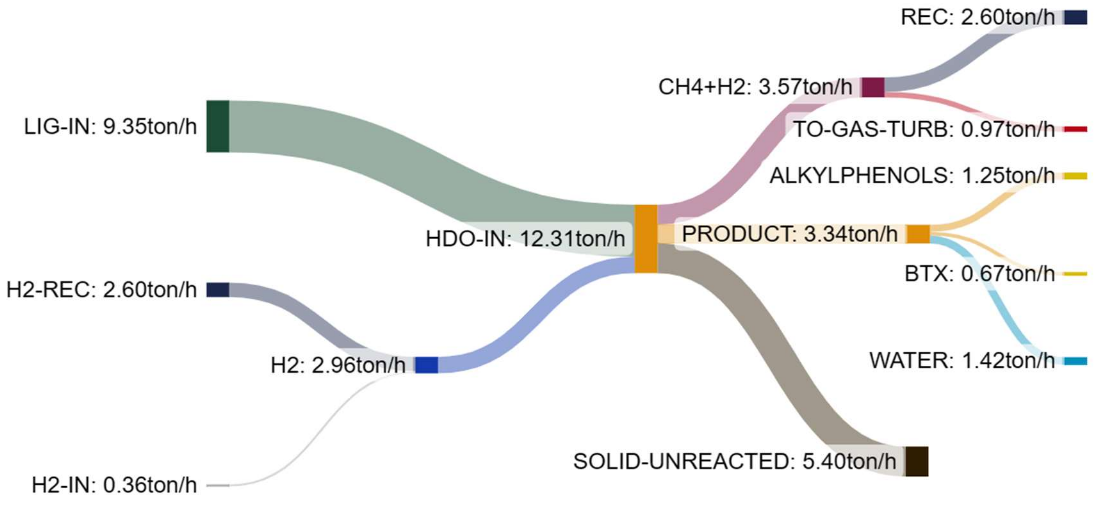


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Figure G.2 Mass balance of the lignin HDO process

of surplus electric energy is produced: it can be used to operate an electrolyzer in order to obtain pure H_2 gas and to reduce the need of fresh makeup hydrogen.

MAKE UP HYDROGEN NEED IN THREE DIFFERENT SCENARIOS

Two different scenarios were considered in order to assess the quantity of make up hydrogen needed for the process:

- HDO: hydrodeoxygenation is performed and waste streams are used to obtain electrical energy, by means of gas and steam turbines.
- HDO+electrolyzer: similarly to “HDO” scenario, both off-gas and unreacted solid streams are used for electric energy generation. The electric power obtained is fed to an electrolyzer, which is able to convert water into pure hydrogen gas.

Table G.2: Technology profile for the lignin HDO process

Scenario	H_2 produced	Make up H_2 needed	Electric energy to the grid
HDO	-	362 kg/h	13.2 MW
HDO+electrolyzer	274 kg/h	88 kg/h	-

ECONOMIC ASSESSMENT OF THE LIGNIN HDO

An economic assessment of Lignin HDO process is provided in Table G.3 (CAPEX), Figure G.3 (OPEX) and Figure 5 (Minimum Selling Price of AP). Capital costs computed for the optimal economic case can easily be seen that most of the expenses are due to the HDO reactor and to the combined heat and power production system (turbines and steam generators).

Table G.3: CAPEX for the economic optimal HDO case.

Equipment	Bare modulus cost
HEAT EXCHANGING NETWORK	526,000 €
HDO reactor	9,887,000 €
CATALYST	187,000 €
FLASH1	348,000 €
FLASH2	358,000 €
FLASH3	334,000 €
SOLID SEPARATION	3,112,000 €
PSA	321,000 €
DIST. COL. 1	1,266,000 €
DIST. COL. 2	420,000 €
WATER PUMP 1	504,000 €
WATER PUMP 2	847,000 €
H2 MAKEUP COMPRESSOR	2,507,000 €
H2 RECYCLE COMPRESSOR	1,373,000 €
HRSG-SOLID	12,339,000 €
HRSG-GAS	6,457,000 €
STEAM TURBINES	23,275,000 €
GAS TURBINE	13,341,000 €
CTBM	77,402,000 €
CTDC	108,363,000 €
CTCI	112,698,000 €

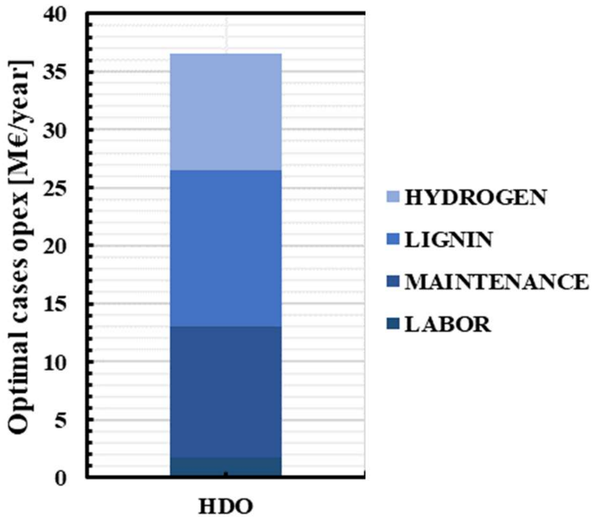


Figure G.3: OPEX for direct HDO.

OPEX results are schematically represented in Figure G.4. Lignin represents the main OPEX, considering a Lignin supply cost of 200 €/t.

The lowest obtainable, hence optimal, alkylphenols minimum selling price from both processes can be compared (Figure G.4). On the whole, the MSP equal to 3.47 €/kg$_{AP}$, which is coherent with literature values (Bbosa et al., 2018). The purification and reaction sections have a very limited impact on the final alkylphenols payback selling price. Conversely, the raw materials (variable operative costs) contribution is extremely significant for both processes: namely, lignin has a consistent impact (around 35%), while hydrogen influence is higher for the direct hydrodeoxygenation configuration, since it requires an overall higher amount of H_2. The cost of the combined energy and heat production section importance is remarkable in the direct HDO process: nearly 17% of the MSP (around 0.60€/kg$_{AP}$) is due to that. It is also noteworthy to mention that, for the direct hydrodeoxygenation process, the revenue due to energy and byproducts production has a non-negligible effect: in fact, because of the consistent electric power yield, the obtained MSP is reduced of around 0.60 €/kg$_{AP}$, which is very similar to the energy recovery section cost contribution.

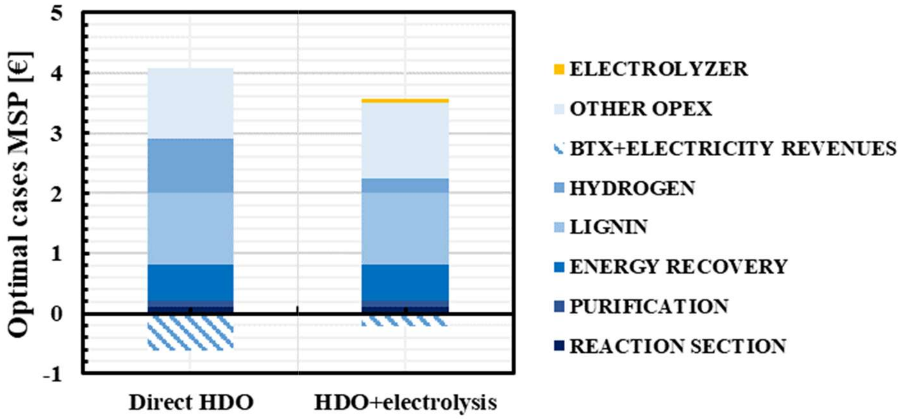


Figure G.4: Alkylphenols MSP for both H_2 supply mode cases.

As explained in the previous section, the electric power generated in the direct HDO process can alternatively be used to produce pure H_2 in situ by means of an electrolyzer. The considered cases have been summarized in Table G.2: only the economic optimum configuration (which is the one able to yield the lowest MSP_{AP}) of each process has been analyzed.

The case “HDO+electrolysis” has been added to the already conducted analysis: it consists in a process which is basically identical to the direct HDO one, with the only difference that the electric energy produced is not sold back to the grid but is used to power an electrolyzer. Obviously, the overall H_2 need for both HDO processes is the same: nevertheless, in the one in which the electrolysis reaction is conducted, 73.7% of the hydrogen required is obtained in situ. This important contribution significantly lowers the need to acquire pure H_2 from external sources, since the necessary make-up stream is equal to 92.16 kg/h (instead of the 350.92 kg/h of the direct HDO case). Anyhow, the need for a lower fresh hydrogen inlet stream is not enough to state that the configuration which includes the electrolyzer is more convenient that the original one: for this reason, an economic analysis has been conducted, which results are reported in Figure G.4.

Overall, the alkylphenols minimum selling price obtained for the HDO+electrolysis optimal case is lower than the direct HDO one: in fact, the first is equal to 3.14€/kg_{AP}, while the second stands at 3.28€/kg_{AP}. Since the two processes are the same from a technical point of view, meaning that they yield the same quantity of alkylphenols and of byproducts and they generate equivalent amounts of electric power, the H_2 in situ production has improved the economic performance of the direct hydrodeoxygenation layout, by reducing the MSP of 0.14 €/kg_{AP}. In fact, while this configuration implies higher investment costs due to the presence of the electrolyzer, it greatly reduces the hydrogen requirement and, subsequently, the MSP contribution linked to its cost. For this reason, H_2 has an impact of nearly 0.90€/kg_{AP} in the direct HDO case, while it’s equal to 0.23€/kg_{AP} in the HDO+electrolysis process: hence a reduction of nearly 4 times is recorded. The electrolyzer contribution on the final minimum selling price is around 0.07€/kg_{AP}; overall it can be concluded that its implementation is completely absorbed by the H_2 impact reduction from the cost point of view and, for that matter, it’s outright favourable in the economic scenario considered (H_2 cost equal to 4€/kg).

ASSESSMENT OF THE TECHNOLOGY READINESS

An assessment of the technology readiness of the lignin HDO concept is provided in Table G.4.

Table G.4: Assessment of technology readiness for the lignin HDO concept

Process steps	TRL	Weight [%]	Comments
Lignin pretreatment	7	15	Covers the lignin drying, particle sizing preparation, solid transport by a gaseous phase, heating until to the reactor temperature.

Process steps	TRL	Weight [%]	Comments
HDO reactor	3-4	50	The lignin HDO reactor is currently under study. Its TRL is low, further specific European projects have to fill this challenge (i.e. https://www.cbe.europa.eu/open-calls-proposals).
Product separation	5-6	15	Due to quantity of possible products that can be generated by the HDO process, a first liquefaction is necessary to recover the H_2-rich stream.
PSA	7-8	5	PSA to purify a hydrogen-rich stream is a conventional process, about the specific utilization considering all possible products can represent a new obstacle to overcome.
Gas turbine and steam turbines - off-gas side	7-8	5	The gas turbine can be considered the same as a natural gas turbine, on the other hand, some products (e.g. BTX) can be found also in the methane-rich stream.
Boiler and steam turbines - residual solid side	7-8	10	Further, the combustion of the residual solid is not a commercial process yet.

Appendix H - Hydrothermal liquefaction in combination with methanol synthesis: Coupling of HTL with carbon capture and methanol synthesis integrating hydrogen production.

BACKGROUND

In this study the synergies between Biofuel and E-fuel production are explored. A constant production of biofuels via hydrothermal liquefaction (HTL) is combined with a dynamic production of methanol by carbon capture from the HTL gas and dynamic hydrogen production from water electrolysis. A techno-economic analysis of the system is performed using the energy planning software energyPRO.

PROCESS DESCRIPTION

The system evaluated consists of an HTL plant with integrated carbon capture and methanol synthesis. The size of the plant corresponds to the production of 100 kt/y of drop-in biofuels, equivalent to approximately 2,000 barrels-per-day (BPD) or 3.5% of the capacity of a medium size refinery of 68,000 BPD capacity. Forestry residues is selected as feedstock for the process as one of the resources listed in Annex IX of RED II (unchanged in RED III) that

Table H.1: Technology profile for the BECCUS-HTL concept.

Designation	BECCUS-HTL
Developer/Promotor	Steeper Energy (HTL) and Aalborg University (integration study)
Conversion technology	HTL+H_2T integrated with Methanol synthesis
Feeding System	High pressure pump for biomass + water slurry
Principal Feedstock(s)	Woody biomass
Principal Application(s)	Biofuels and E-fuels (Methanol)
Scale	production of 100 kt/y of drop-in biofuels
Development Status	HTL - Demonstration scale. Integration with methanol

Designation	BECCUS-HTL
	synthesis - Lab scale

complies with sustainability requirements for advanced biofuels production. A process scheme of the system showing the inputs/outputs and the main process units is seen in Figure H.1.

The model consists of an HTL to drop-in biofuels process, a methanol plant as well as an electrolyzer producing hydrogen for the processes.

In the HTL unit biomass/wood is converted to biocrude, which is used to produce hydrotreated biofuels in the hydrotreater unit in combination with hydrogen. Finally, the hydrotreated biofuels are distilled into Gasoline, Jet fuel, Diesel, and Heavy fuel.

The CO_2 from the HTL gas is captured, after which it is sold for underground storage or used in combination with hydrogen to produce methanol in the methanol synthesis. Whether it is sold or utilized depends on the optimum hourly operation that maximises the yearly net cash flow of the system.

The system is operated as "semi-islanded" with an installed capacity of Renewable Energy Sources (RES) of 150 MW, consisting of 100 MW of solar PV and 50 MW of wind turbines. Weather data and hourly spot market electricity prices for the western part of Denmark in 2019 are used.
Given the relatively high carbon intensity of current electricity grid, the supply is restricted to hours where the carbon intensity is below or equal to 90 g CO_{2eq}/kWh, in line with estimates in literature to guarantee the sustainability of electro fuels. When the carbon intensity is above or equal to 90 g CO_{2eq}/kWh the load of the methanol plant is decreased to the minimum load of 50% capacity.

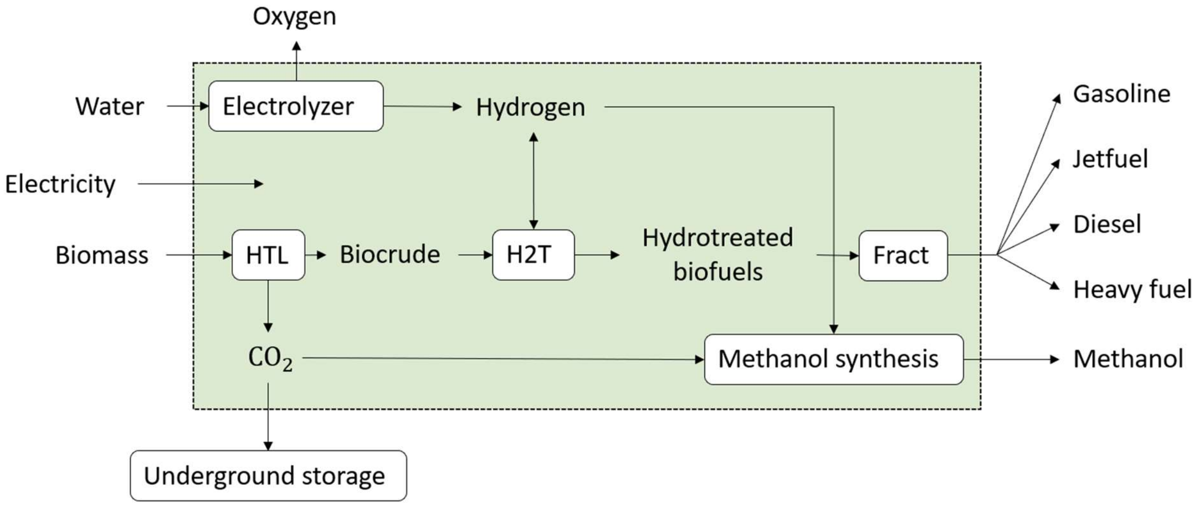


Figure H.1: Process scheme of the system.

ASSESSMENT OF THE TECHNOLOGY READINESS

An assessment of the technology readiness of the BECCUS-HTL concept is provided in Table H.2. The aspect with lowest TRL 4-5 (weakest links) is hydrotreatment of the HTL biocrude. The weighted average TRL of the whole process chain is 7-8.

Table H.2: Assessment of technology readiness for the BECCUS-HTL concept.

Process steps	TRL	Weight [%]	Comments
Electrolyzer	9	10	Equipment available commercially. System integration needs to be tested.
HTL	7-8	50	-
Hydrotreating of HTL biocrude	4-5	20	-
Carbon capture from HTL	5-7	10	Available carbon capture solutions need to be tested for this specific case/ matrix, including system integration. High CO_2 concentration in off-gas.
Methanol synthesis	7-8	20	Any methanol synthesis from CO_2 can be theoretically applied. System integration needs to be tested.

ECONOMIC AND ENVIRONMENTAL ASSESSMENT

The mass balance of the system is illustrated in Figure H.2. With the restrictions on the methanol synthesis based on the carbon intensity in the grid, 68% of the CO_2 is utilised in the methanol synthesis, while the rest is sold for underground storage. When removing the restrictions on the methanol synthesis the share of utilised CO_2 increases to 98%. With the selling prices of methanol and CO_2 used in this study of 1000 €/ton and 100 €/ton respectively, it is more profitable to utilise the CO_2 to produce methanol rather than selling it for underground storage. In Figure H.3 the CO_2 emissions per GJ of fuel produced and the yearly cash flow is seen. The emissions per GJ of fuel produced based on the total emissions associated with the yearly operation of the system and the mass ratio of the produced fuels are in the range 8-12 kg CO_2/GJ.

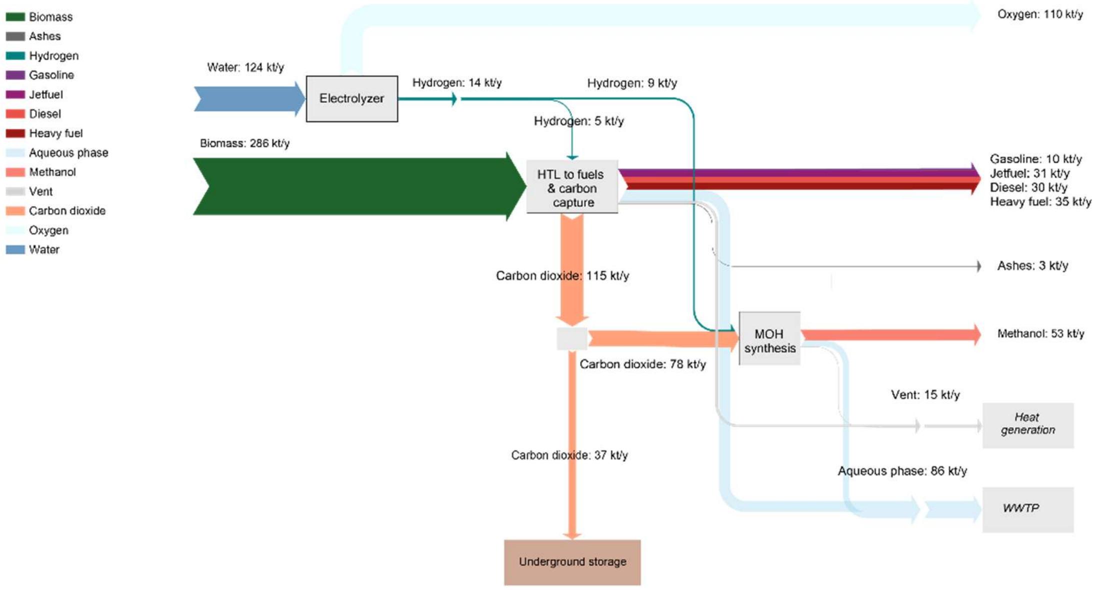
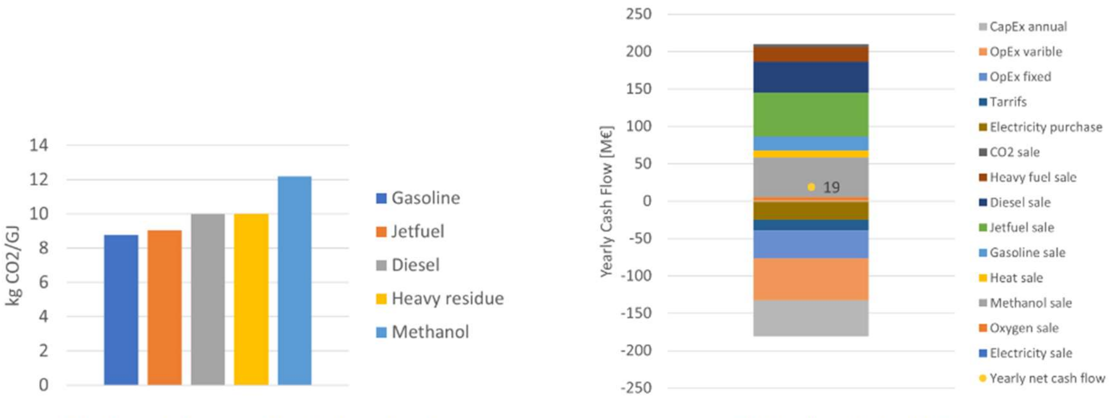


Figure H.2: Sankey diagram showing the mass balance.

The total emissions consist of scope 3 emissions, imported electricity based on the carbon intensity in the grid, and negative emissions by underground carbon storage.

With the economic data used in this study, the optimum operation of the system results in a yearly net cash flow of 19 M€, consisting of the yearly revenues and expenses shown in the figure. The purchase of electricity including tariffs results in expenses of 39 M€, while the revenue from the sale of electricity is <1 M€. This indicates that the installed capacity of 150 MW of RES is insufficient for this system. Operating the system as “islanded” (only with RES and no connection to the grid) would require intermediate storage of hydrogen and/or electricity, as the HTL to drop-in biofuels should be operates continuously even in the hours of no RES production. Covering the total yearly electricity demand for the given location and year can be achieved with various combinations of solar PV and wind turbines, amongst the combinations are 0 MW solar PV + 300 MW wind turbines and 250 MW solar PV + 200 MW wind turbines.



(a) CO_2 emissions per GJ of fuel produced.

(b) Yearly cash flow [M€].

Figure H.3: Environmental and economic results.

Removing the restrictions regarding the carbon intensity in the grid increases the yearly net cash flow to 24.4 M€ as it enables a higher methanol production. Increasing the production of methanol in hours with high carbon intensity in the grid, however, results in higher CO_2 emissions per GJ of fuel produced. For the biofuels the specific emissions are doubled, while for the methanol the emissions per GJ of fuel produced are tripled.

Appendix I - PTG-HEFA | HEFA refining with electrolytic hydrogen

BACKGROUND

The German government's climate protection targets and the aviation industry's international voluntary commitment to reduce greenhouse gas (GHG) emissions represent an enormous challenge for the constantly growing aviation sector. Sustainable aviation fuel (SAF) with correspondingly high specific GHG reductions plays a key role in achieving these targets. Of the SAF authorised by the relevant standards, only synthetic paraffinic kerosene (SPK) of hydroprocessed esters and fatty acids (HEFA) is available in significant quantities in the short to medium term. In 2021, global production capacities for HEFA products (mainly HVO diesel) amounted to 8 million tonnes per year (of which 4.8 million tonnes in the EU) and a further 16.5 million are under construction or in the planning phase. HEFA-SPK may be blended with conventional Jet A1 fuel up to 50 % by volume. A rapidly growing market for this fuel is currently emerging.

The underlying feasibility analysis for a PTG-HEFA hybrid refinery started at this point. Jatropha was selected as the raw material for the basic project. The subject of the study is the provision of electrolytic hydrogen. This power-to-gas process is an essential auxiliary process in the production of HEFA-SPK.

PROCESS DESCRIPTION

Table I.1: Technology profile for the PTG-HEFA hybrid refinery concept

Designation	PTG-HEFA (SAF)
Developer/Promotor	Various
Conversion technology	Hydrotreatment of esters and fatty acids (HEFA) with electrolytic hydrogen
Feeding System	n/a
Principal Feedstock(s)	Oil-rich feedstock like oilseeds and fruits, algae, animal fats, used cooking oil (UCO), tall oil and other biocrudes (case study: different vegetable oils and UCO)
Principal Application(s)	Naphtha, synthetic paraffinic kerosene, diesel, waxes and other by-products

Designation	PTG-HEFA (SAF)
Scale	0,5 million tonnes vegetable oils
Development Status	Here results of a feasibility study finalised in 2017. Example for realisation of such a concept is the OMEGA GREEN in Paraguay and DSL-O1 in the Netherlands (status open). Theoretical description of the technology with high specific technology readiness level

A process scheme including a simplified energy balance of the PTG-HEFA concept is shown in Figure I.1. The technical feasibility of a PTG-HEFA hybrid refinery with a feedstock input of 500,000 tonnes per year of jatropha oil to HEFA-SPK was set as reference case to be realized in 2024/25. In addition to that, nine alternative scenarios have been investigated with regard to different approaches for electricity supply, hydrogen production, feedstocks, and refinery products.

The specific assumptions for the different scenarios are summarized in Figure I.2. The different scenarios are grouped with special regard to the following:

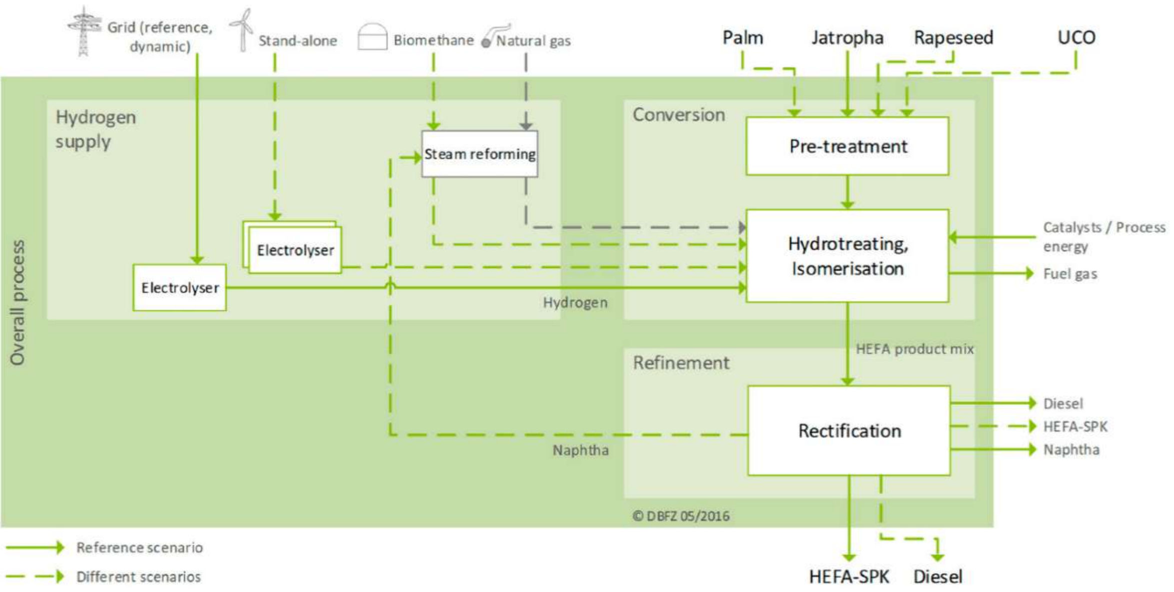
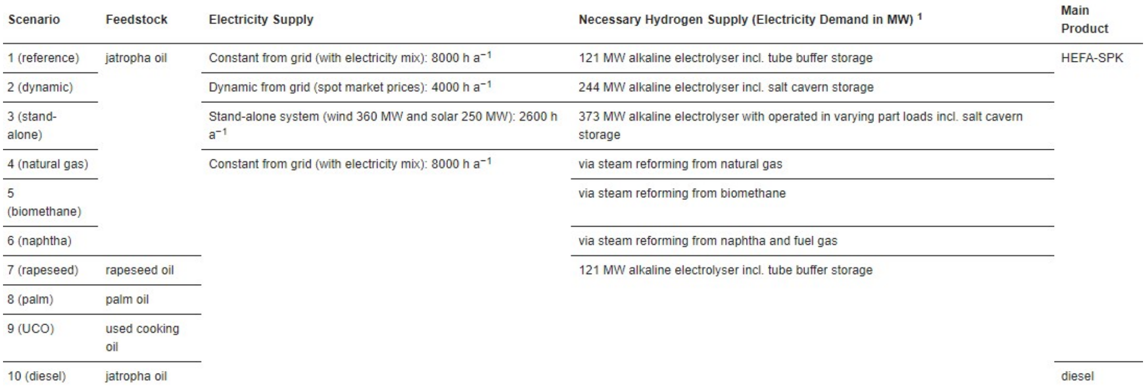


Figure I.1: Process scheme PTG-HEFA concept.

(a) Electricity supply
constantly from the electricity grid with the specific electricity mix based on different primary energy sources, in a dynamic mode with based best cost-effectivity (in case of cheap spot market prices) whilst using electricity from the grid or as stand-alone power system based on fluctuating renewable energies (here, onshore wind park and solar photovoltaic park);
(b) Hydrogen supply
based on alkaline electrolysers with different storage systems depending on the electricity supply and—more conventionally—based on steam reforming from either natural gas, biomethane, or internal use of by-products of the HEFA plant such as naphtha and fuel gases (as is done today in most of the HVO/HEFA plants).
(c) Feedstocks
as alternatives to jatropha oil; rapeseed oil with regard to a domestic option, palm oil with regard to cost and specific hydrogen demand, and used cooking oils (UCO) with regard to lower GHG emissions (according the current regulations in the EU) .
(d) Main product
referring to the different operational modes of a HEFA plant with regard to products, compared to the HEFA-SPK case, the alternative operation mode is focused on producing diesel.



Scenario	Feedstock	Electricity Supply	Necessary Hydrogen Supply (Electricity Demand in MW) [1]	Main Product
1 (reference)	jatropha oil	Constant from grid (with electricity mix): 8000 h a^{-1}	121 MW alkaline electrolyser incl. tube buffer storage	HEFA-SPK
2 (dynamic)		Dynamic from grid (spot market prices): 4000 h a^{-1}	244 MW alkaline electrolyser incl. salt cavern storage	
3 (stand-alone)		Stand-alone system (wind 360 MW and solar 250 MW): 2600 h a^{-1}	373 MW alkaline electrolyser with operated in varying part loads incl. salt cavern storage	
4 (natural gas)		Constant from grid (with electricity mix): 8000 h a^{-1}	via steam reforming from natural gas	
5 (biomethane)			via steam reforming from biomethane	
6 (naphtha)			via steam reforming from naphtha and fuel gas	
7 (rapeseed)	rapeseed oil		121 MW alkaline electrolyser incl. tube buffer storage	
8 (palm)	palm oil			
9 (UCO)	used cooking oil			
10 (diesel)	jatropha oil			diesel

Figure I.2: Scenarios and their specific assumptions.

The entire PTG-HEFA plant is operated with an annual full utilisation period of 8,000 hours. This applies to both the HEFA plant and the PTG plant for the continuous supply of hydrogen to the HEFA plant.

A more detailed description of the assumptions for the design of a PTG-HEFA plant are described in [1].

PTG process of hydrogen

The power-to-gas part for the hydrogen supply is realized as alkaline electrolysis of water. The decision to include alkaline electrolysers into the described concept was based on the results of a comprehensive market assessment (including concrete requests for proposal from international electrolyser manufacturers) that shows that only this technology can be provided by 2024/25 in the required capacity. Different electrolysis sizes are based on different electricity supply scenarios that influence the annual operating hours of the electrolysers. The shorter the equivalent full load period of the electrolyser, the larger the capacity of the electrolyser to cover the hydrogen demand of the HEFA process (121 to 373 MW hydrogen including electricity for hydrogen compression). The hydrogen demand of the HEFA plant amounts to about 17,500 tonnes per year. At an equivalent full load period of 8,000 hours per year, the electrolysis plant has to generate about 2.19 tonnes hydrogen per hour, leading to an electricity input of 121 MW. At an equivalent full load period of 2,600 hours per year, the electrolysis plant has to generate about 6.66 tonnes hydrogen per hour, leading to an electricity input of 373 MW.

The same is true for the required type and size of the hydrogen storage, and thus the additional electricity demand for storage loading and unloading. An underground tube buffer of 6,860 m^3 for short-term storage is required for constant and permanent electricity supply from the power grid (grid supply: 8,000 h/a). If even only a proportion of the hydrogen from a stand-alone operation is to be utilised, salt caverns must be used as buffer storage. For a dynamic change between stand-alone operation power and grid power (grid supply: 4,000 h/a), caverns with a volume of 104,000 m^3 (net hydrogen storage capacity of 775 tonnes) must be utilised. With a constant power supply from stand-alone power generation and without grid power, caverns with a volume of or213,000 m^3 (net hydrogen storage capacity of 1585 tonnes) must be used.

In order to produce the required electricity in the stand-alone scenario using 100 % wind and solar power (scenario 3), an appropriate combination would be a wind farm with 360 MW installed power and a solar power park with an area of 625 hectares.

HEFA production and product preparation

The HEFA plant includes a pre-treatment of the feedstock (degumming, bleaching, and neutralisation), two main processing steps (hydrotreating, subsequent hydrocracking, and isomerisation stages), and a product separation (distillation).

Detailed information about the design of HEFA plants are described elsewhere [2].

As a state-of-the-art technology within HEFA facilities, in scenarios 4-6 the hydrogen production via steam reforming is considered instead of the PTG part. In the case of steam reforming, the total electricity demand of the plant is considerably reduced.

Figure I.3 shows the annual mass balance of the reference concept. Starting from the raw material jatropha oil, the individual process steps leading to the target product HEFA-SPK and the by-products diesel and naphtha are shown.

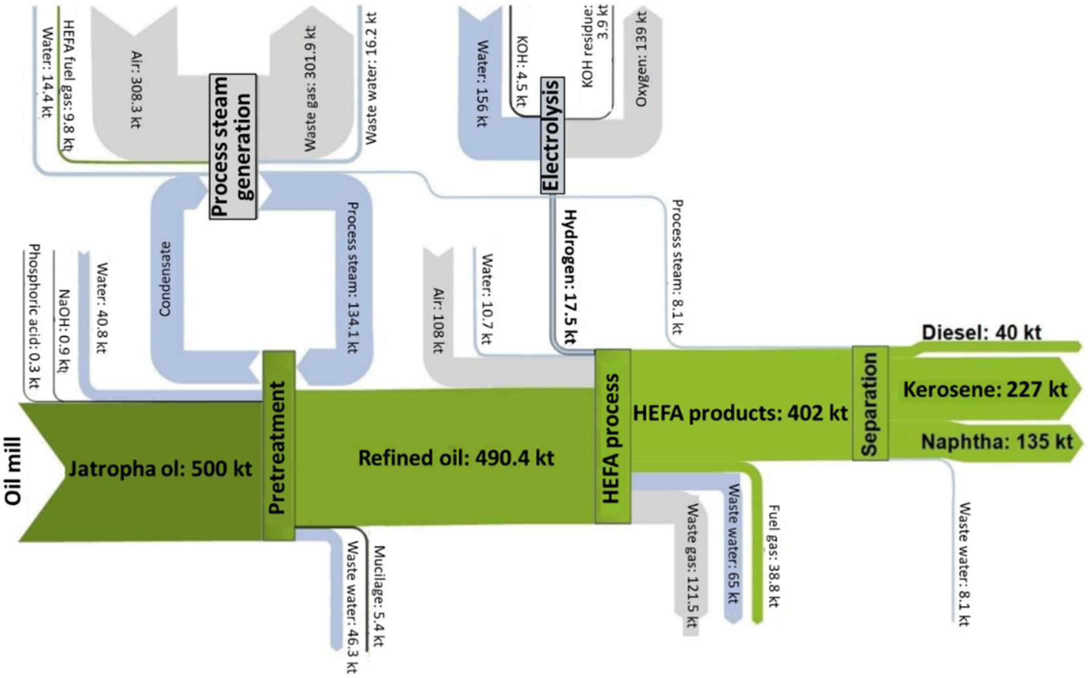


Figure I.3: Mass balance of the PTG-HEFA reference concept

DEVELOPMENT STATUS, APPLICATIONS, AND PRODUCTION SCALE

In 2021, there were twenty-nine diesel focused HEFA production facilities worldwide with an annual production capacity of 8.0 million tonnes. These plants are almost focused on the production of HVO/HEFA diesel fuel with only small amounts of HEFA-SPK. A large number of new HEFA plants will be built in the coming years. The production of HEFA-SPK will increasingly become the focus of these. With a view to advanced biofuels, HVO/HEFA technology is well suited for processing challenging feedstocks. Most plants are so-called multi-feedstock plants, which can utilize a range of different feedstocks – apart from a few plants in Scandinavia that process only tall oil.

Eni (Italy), Neste (Finland), The Renewable Energy Group (USA), TotalEnergies (France) and UPM Biofuels (Finland) are currently the main actors in HEFA development. Total fuel capacities of stand-alone HEFA plants currently range from 130,000 tonnes per year (UPM Biofuels, Lappeenranta refinery) to 1.3 million tonnes per year (Neste, Singapore refinery).

SkyNRG and Shell announced in 2019 to develop the first dedicated SAF production plant [3]: The DSL-01 production plant in Delfzijl [...] is on schedule for commissioning in 2024, representing the earliest dedicated commercial supply of SAF to the aviation market in Europe. The plant will annually produce 100,000 tonnes of SAF, corresponding to a reduction in lifecycle CO_2 equivalent emissions of approximately 270,000 tonnes. The plant will also produce naphtha, and 15,000 tonnes of bioLPG annually as a by-product. [...] The feedstocks used for production will be waste and residue streams, such as used cooking oil, sourced predominantly from regional industries. The facility will run on sustainable hydrogen, produced local to the site in the Groningen Seaport. The combined benefits of the feedstocks, sustainable hydrogen, and use of low carbon energy to power production, will contribute to the production of SAF with lifecycle carbon emissions approximately 85 % lower than conventional jet fuels, as estimated by the Roundtable on Sustainable Biomaterials.

ASSESSMENT OF THE TECHNOLOGY READINESS

An assessment of the technology readiness of the PTG-HEFA concept is provided in the table below.

Table I.2: Assessment of technology readiness for the PTG-HEFA concept

Process steps	TRL	Weight [%]	Comments
HEFA process	9	50	
Alkaline electrolysis	9	50	
Overall “Weighted Average”	9		
Overall “Weakest Link”	9		

ECONOMIC AND ENVIRONMENTAL ASSESSMENT

The GHG balances for the scenarios have been conducted according to the European Renewable Energy Directive methodology from 2009 (EU Directive 2009/28/EC). Referring to Figure I.4, the provision of jatropha-based HEFA-SPK can achieve a GHG reduction of 76 % compared to the fossil reference (83.8 g CO_2-equivalent per Megajoule). Prerequisite is the exclusive use of electricity from wind and solar systems to cover the process energy demand. The provision of hydrogen via electrolysis is electricity-intensive and the use of grid electricity under the forecasted conditions leads to significantly higher emissions. In particular, the relatively high share of fossil fuels in the German electricity mix prevents higher GHG reductions. Other ways to reduce emissions from the supply of HEFA-SPK, in addition to the use of renewable electricity or a significantly higher share of the same in the electricity mix, is the use of in-process provided naphtha for hydrogen production via steam reforming. Reductions of more than 70 % can be achieved here. On the other hand, the use of sustainably grown jatropha, which means jatropha grown mainly from extensive cultivation, can reduce overall GHG emissions.

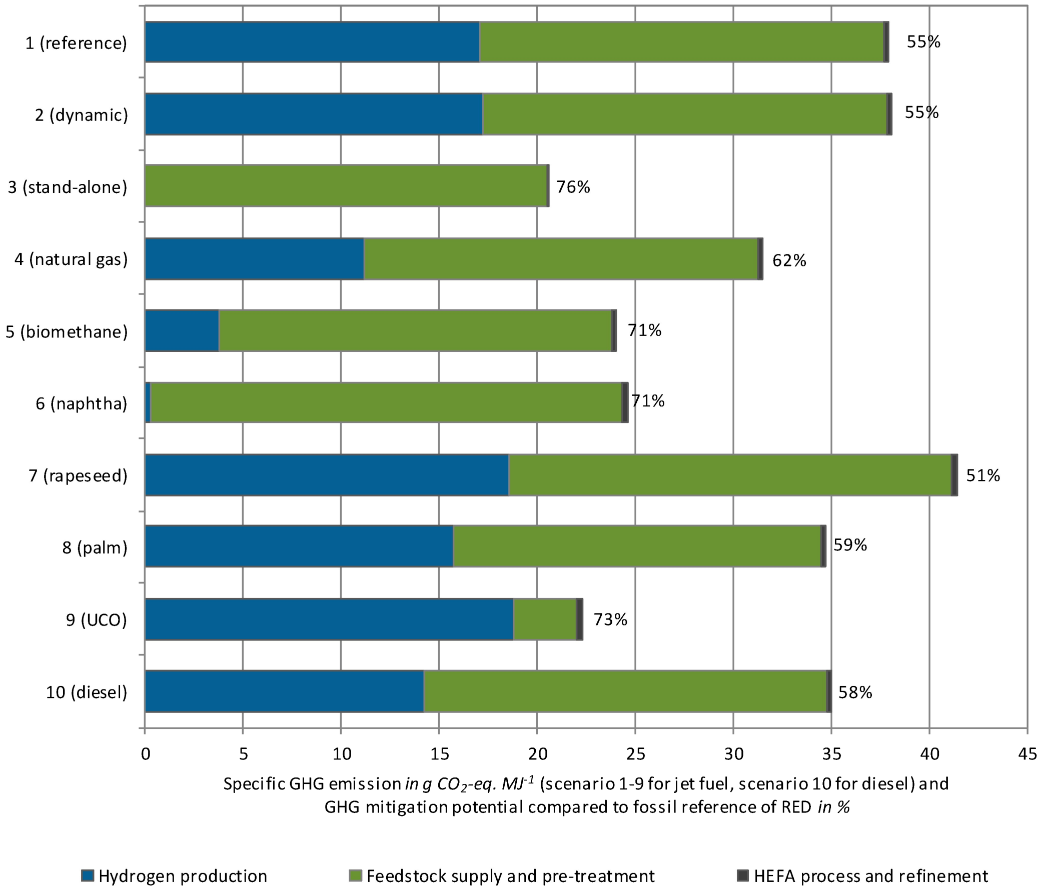


Figure I.4: Specific GHG emission per scenario for HEFA-SPK and HEFA-diesel, and GHG mitigation potential

Based on the results of the technical assessment, specific production costs were calculated. Referring to Figure I.5, the bandwidth of total production costs is from 1,295 to 1,800 EUR per tonne for HEFA-SPK, which is well above the current market price of fossil jet fuel of around 425 EUR per tonne. Costs can be lowered to 1,210 EUR per tonne for shifting the product values to diesel; this is due to the smaller production of short-chained, low-value naphtha and fuel gases. (Note: Economic assessments are from 2017. An adjustment for inflation must be taken into account accordingly.)

When analysing the impact of the specific cost components to the total costs, feedstock costs dominate and the contribution from the PTG part is significant. Costs of capital and operation seem negligible in that context. The scenario of steam reforming with naphtha and fuel seems to be more favourable than PTG in terms of costs. This is due to the fact that only small amounts of revenue are lost but large investments in electrolysers and hydrogen storage and therefore also electricity costs are saved.

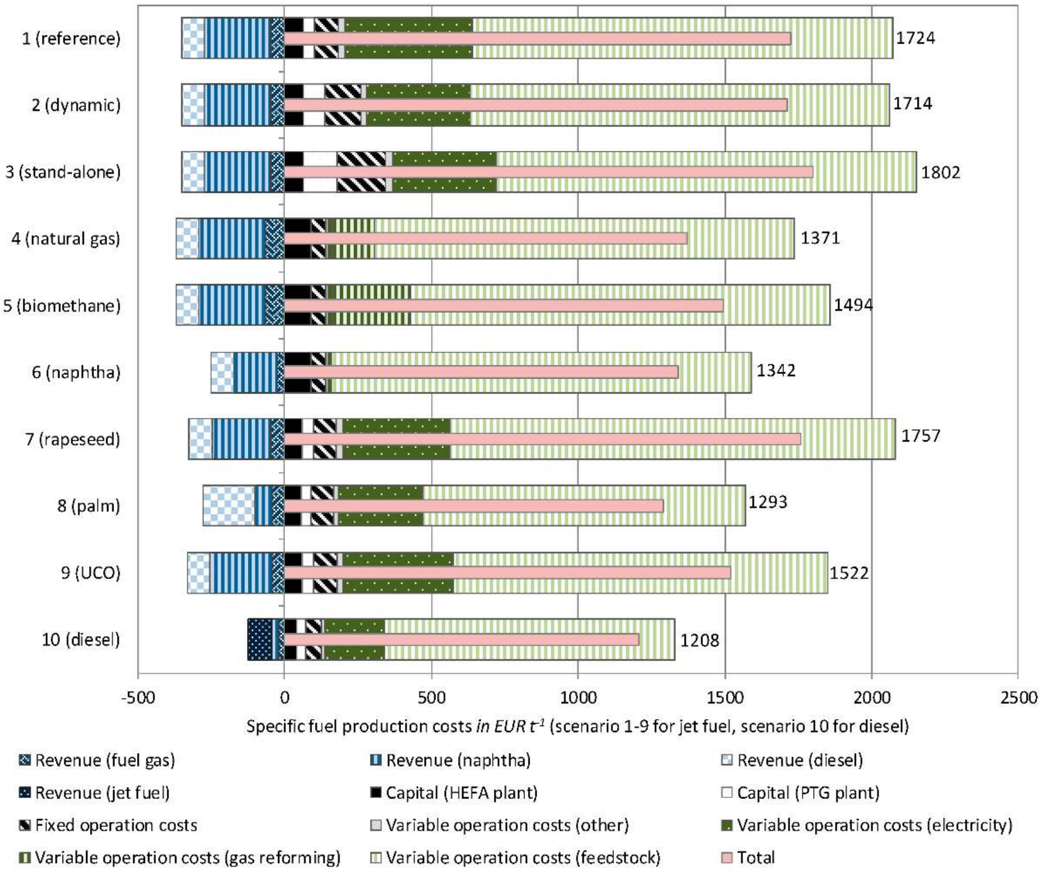


Figure I.5: Specific fuel production costs per scenario.

A more detailed discussion of the results, an exemplarily sensitivity analysis and comparisons with results of similar investigations for HEFA fuels are presented in Zech et al. [4].

DEVELOPER FEEDBACK

Please note: as here results of a feasibility study are reflected there is no dedicated feedback of developer. For this please refer to the OMEGA GREEN project.

Excursion on Favourable Regions

After discussing the interim results, three further concepts in favoured regions (exemplary for Sweden, Spain and Namibia) were included in the study. This further concept development was aimed at a more favourable supply of renewable electricity for hydrogen production compared to Germany and thus has a direct impact on the cost and greenhouse gas balance.

Sweden

The windiest locations in Sweden are mountainous regions far from the coast. However, as a corresponding location must have a harbour, a selection must be made from the harbour locations. The cargo harbour, the airport and a comparatively high wind speed speak in favour of Trelleborg on the south coast as a potential location. There are as yet no refineries here that could be converted for PTG-HEFA production. If this argument is important for economic viability, Gothenburg would be the location that fulfils all the requirements and also has only slightly lower wind speeds.

Spain

The Cartagena region - in combination with solar energy supply - can be recommended for a profitability analysis of the planned plant. Within Spain's sunniest region, this location has a seaport, refinery infrastructure and an airport. If the PTG-HEFA refinery were to be powered by wind energy, however, a site in La Coruña would be ideal, as it has the highest wind speeds and is also home to a seaport, an airport and a refinery.

Namibia

Of the country's two seaports, Lüderitz has the highest global radiation. Since Lüderitz also has an airport, it is recommended that the profitability calculation for the PTG-HEFA refinery be calculated on the basis of the two solar power generation technologies for this location.

Comparison with boundary conditions in Germany

For Spain and Sweden, a location with comparatively high wind speeds was selected for the installation of an onshore wind farm, but where there is also a seaport, an airport and, if possible, a refinery nearby. In addition, a location with comparatively high solar radiation was selected for the installation of a ground-mounted PV system in Spain and Namibia, for which the same applies. In Spain, La Coruña was identified as a location on the Atlantic coast with a higher wind potential than the Gothenburg location selected in Sweden, meaning that a wind farm in La Coruña could achieve more full-load hours than in Gothenburg. However, Gothenburg has an advantage over La Coruña in that there are three refinery sites nearby. This increases the possibility of converting at least one of the three refineries to PTG-HEFA products.

With regard to the installation of solar parks (PV, CSP), Spain and Namibia offer much more favourable conditions than Germany and Sweden. Spain and Namibia are exposed to significantly higher global radiation and solar parks can therefore achieve a much higher number of full load hours and therefore do not need to be as large in order to generate the same amount of electricity per year. In terms of solar radiation, Namibia is significantly more favourable than Spain, but there are also significant disadvantages: There is no refinery in this country (all oil products are imported) and there is no experience in the field of solar thermal power plants or in the field of generating hydrogen. In addition, the infrastructure required (electricity grid, rail and road connections) for the realisation of such an ambitious project is not available to the same extent as in Spain.

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Appendix J - HVO and SAF - Plant for HVO and SAF production using hydrogen from hydroelectricity (Omega Green) concept

BACKGROUND

The Omega Green Project is an industrial complex being built by Be8 S/A to produce advanced biofuels using vegetable oil, used cooking oil (UCO), and animal fat as feedstock. The products, hydrotreated vegetable oil (HVO, also known as renewable diesel), sustainable aviation fuel (SAF, also known as synthetic paraffinic kerosene), and green naphtha, are produced via hydrotreatment processes using green hydrogen. The project uses proprietary technology from Honeywell UOP (also known as Ecofining™). The project is currently under development in Paraguay, and it intends to use feedstock produced in Paraguay, Brazil, Argentina, and surrounding countries.

PROCESS DESCRIPTION

Table J.1: Technology profile for HVO and SAF production in the Omega Green Project

Designation	OMEGA GREEN
Developer/Promotor	Be8 S/A
Conversion technology	Hydrotreatment, isomerization, and hydrocracking of vegetable oils and animal fats
Feeding System	n/a
Principal Feedstock(s)	Vegetable oil, such as soybean oil, or animal fat, such as tallow, and other advanced feedstock such as UCO, carinata oil, pongamia oil, fatty acids, or acids oils, etc
Principal Application(s)	Production of HVO, SAF, green naphtha, green liquid pressurized gas, and renewable plant off-gas (for plant heating)
Scale	20.000 barrels per day
Development Status	Under Development

A process scheme including a simplified energy balance of the conversion of vegetable oils to HVO, SAF, and green naphtha is shown in Figure J.1.

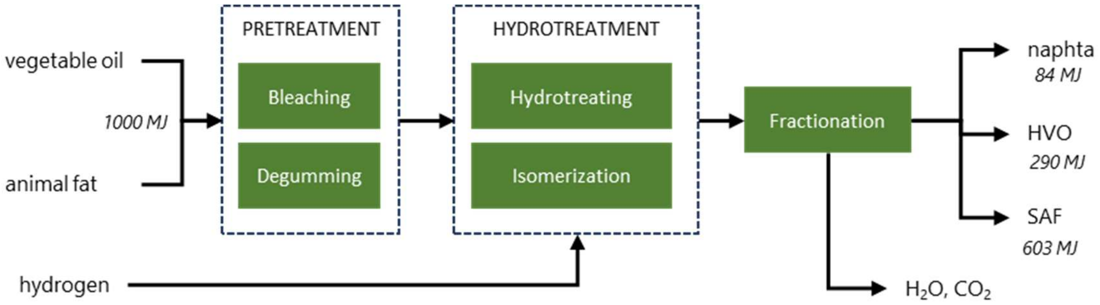

Figure J.1: Process scheme of the conversion of vegetable oil and animal fat to HVO and SAF [1].*For reference only.*

In this process, feedstocks are pre-treated to avoid contamination of catalysts during the hydrotreatment steps. Technology providers of HVO processes have very stringent quality requirements for feedstock, which makes this process step very important. Due to the unique characteristics of vegetable oils and animal fats, they need to be treated separately. In the case of vegetable oils, bleaching, degumming, and neutralization are necessary in a basic configuration. Additionally, pre-filtration, heat treatment, and a second bleaching process might be required depending on feedstock quality. Acid oils might require fat splitting and fatty acid distillation depending on the HVO process requirements.

In hydrotreatment, pre-treated vegetable oils and fats (50/50 %wt) are hydrogenated and hydrocracked in the presence of catalyst and hydrogen at high temperature and pressure. This process removes heteroatoms such as oxygen attached to the carbon chains of the feedstock (mainly triglycerides and fatty acids), resulting in paraffins of different chain lengths. The following chemical reactions represent a general view of the chemical process of conversion of fatty acids and triglycerides:

$$R\text{-}COOH + 2H_2 \rightarrow R\text{-}H + 2\ H_2O$$

$$C_3H_5(R\text{-}COO)_3 + 12\ H_2 \rightarrow C_3H_8 + 3R\text{-}CH_3 + 6H_2O$$

In these reactions, R-COOH represents fatty acids and $C_3H_5(R\text{-}COO)_3$ represents triglycerides, in which R corresponds to generic structures of carbon and hydrogen, thus yielding hydrocarbons. These hydrocarbons can be further cracked and isomerized so the final product meets end-used criteria according to fuel specifications (HVO for use in diesel engines and SAF for use in jet engines). After reactions, the products are separated into different fractions using conventional distillation processes.

DEVELOPMENT STATUS, APPLICATIONS, AND PRODUCTION SCALE

The Omega Green Project was launched in 2019. Stakeholders and the government of Paraguay signed a free trade zone agreement for the production facility in 2023. The construction of the facility was scheduled to start in 2023, and they expect to start deliveries in 2026. The expected capacity of the facility is 20.000 barrels per day (HVO and SAF capacities combined).

ASSESSMENT OF THE TECHNOLOGY READINESS LEVEL

A technology readiness assessment of the Omega Green project is provided in Table J.2.

Table J.2: Assessment of technology readiness for the HVO+SAF concept [2].

Process steps	TRL	Weight [%]	Comments
Feedstock handling system	9	10	Includes storage and logistics of feedstock
Feedstock pre-treatment	9	40	Feedstock pre-treatment is key for hydrotreating steps. Processes must be reliable to homogenize different feedstocks and prepare them according to the quality requirements of further process steps. Pre-treatment processes need to be developed to work with many different feedstocks to increase process reliability.
Hydrotreatment	9	20	The hydrotreatment process is well developed, with many technology vendors.
Cracking and isomerization	9	20	Cracking and isomerization processes are well developed, with many technology vendors.
Fractionation	9	10	The fractionation of paraffins is well-developed by the oil industry.
Overall “Weighted Average”	9		Based on grades for all steps
Overall “Weakest Link”	9		Integrated operation to be proven when working with different feedstocks on large-scale.

MASS BALANCE OF THE PROCESS

The resulting mass balance of the process is illustrated in Figure J.2 for a general case of hydrotreatment of vegetable oils to produce HVO and SAF.

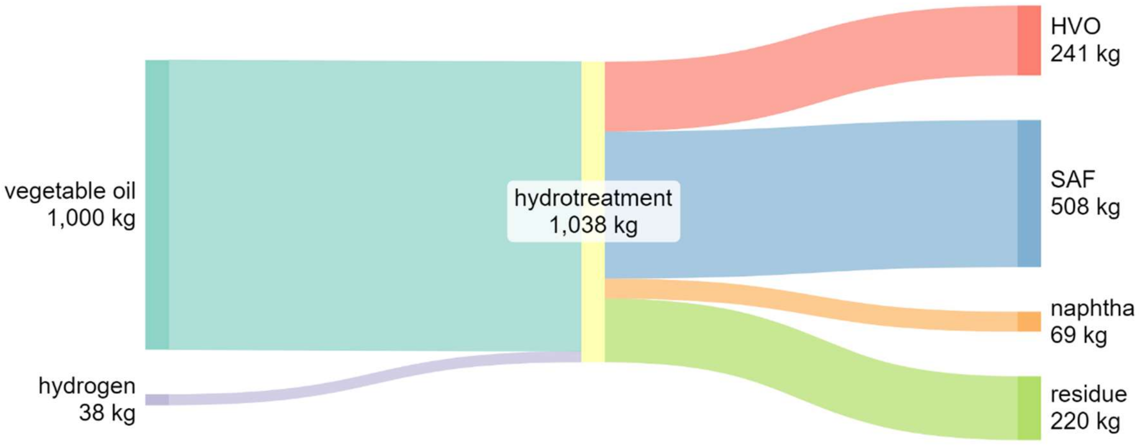

Figure J.2: Mass balance of an example of hydrotreating process to produce HVO and SAF [1]*. For reference only.*

KPI

Yield of HVO+SAF: 0.75 t per t of t of feedstock (potentially, [1])

Global warming potential: 16-22 g/MJ of biofuel (potentially, [1,3])

Biofuel price: US$ -1.00/L-1.30/L of biofuel (potentially, [1])

DEVELOPER FEEDBACK

Omega Green responded to a request for feedback on the assessments with corrections and clarifications on various aspects.

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[3] Johnson E. A carbon footprint of HVO biopropane. Biofuels, Bioprod Biorefining 2017;11:887–96. doi:10.1002/BBB.1796.

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