TECHNOLOGY DATA

Renewable fuels

Technology descriptions and projections for long-term energy system planning.



Danish Energy Agency



Technology Data – Renewable fuels

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Amendment sheet

Publication date

Publication date for this catalogue "Technology Data for Renewable Fuels" is June 2017. The following amendment sheet has been added to record descriptions of amendments in the individual chapters if required. Hereby the catalogue can be updated continuously as technologies evolve, if the data changes significantly or if errors are found.

The newest version of the catalogue will always be available from the Danish Energy Agency's web site.

Amendments after publication date

All updates made after the publication date will be listed in the amendment sheet below. The newest version of the catalogue will always be available from the Danish Energy Agency's web site.

Table 1: Amendment Sheet

Date	Ref.	Description
08/2026	Chapter 80	Restructured and revised text, updates to technical and economic data to reflect the new data sheets and industry advancements. Revised architectural configurations, system boundaries, flexibility potentials and new section on offshore electrolysis.
02/2026	Chapter 86 and 87	Major update for Fast and Slow Pyrolysis with new qualitative text and revised technical and economic data including Catalytic Pyrolysis.
02/2026	Chapter 85	With the new Pyrolysis Overview chapter, the chapters on Catalytic Hydropyrolysis and Methane Pyrolysis were removed.
02/2026	Chapter 85	A new chapter added on Pyrolysis Overview
02/2025	All catalogue	Catalogue restructured
01/2025	Chapter 97	Methanol from Hydrogen and Biogas added
04/2024	Guideline/cover	Updated guideline in terms of scenario projection reference, price year, and further minor updates / new cover
01/2013	Guideline Chapter 80, 93 and 97	Guidelines have been updated to clarify use of contingency and business cycles. Major update with new qualitative text and revised technical and economic data. Biological methanation added. Methanol from hydrogen and carbon dioxide datasheet updated.
08/2023	Chapter 91	Parameter name correction in datasheets for Biogas Upgrading (no change of values).
06/2023	Chapter 81 and 91	Updated chapters and datasheets of Biogas Plants and Biogas Upgrading.

06/2023	Chapter 98	Unit adjustment in datasheet regarding fixed O&M per TPD.
03/2023	Chapter 96	Replaced Methanol from power with Methanol from hydrogen and carbon dioxide.
10/2022	Chapter 86	Corrections for figure 2 in chp. 105 and added note for Sankey diagrams.
04/2022	Chapter 87	Slow pyrolysis added.
04/2021	-	Methane pyrolysis added.
04/2021	Chapter 80	The three chapters on electrolysis (Solid Oxide Electrolyzer Cell, Low Temperature Proton Exchange Membrane Electrolyzer Cell (LT PEMEC), and Alkaline Electrolyzer Cell) have been replaced with one chapter (Chapter 1.1). Both text and data sheets have been updated.
02/2021	Chapter 98	Ammonia from Hydrogen and Air Capture added.
07/2020	Chapter 100	Minor adjustment in DH output for hydrogen to jet, note letters fixed and note added.
05/2020	Chapter 100	Power to Jet added.
02/2019	All catalogue	Minor corrections to text in all chapters. New figures and tables in introduction.
02/2019	-	Version number added to front page.
02/2019	-	Financial data added for configuration 1 (catalytic hydropyrolysis)
12/2018	Chapter 99	The two chapters on Biofuels from Gasification + Fischer Tropsch have been merged to one chapter that is now Chapter 5.5
12/2018	-	Catalytic Hydropyrolysis added.
03/2018	Chapter 83, 94, 99	Thermal gasification chapters added.

Preface

The *Danish Energy Agency* publishes catalogues containing data on technologies for energy carrier generation and conversion. This current catalogue includes technologies for energy carrier generation and conversion.

The catalogue will continuously be updated as technologies evolve, if data change significantly or if errors are found. All updates will be listed in the amendment sheet on the previous page and in connection with the relevant chapters, and it will always be possible to find the most recently updated version on the Danish Energy Agency's website, as well as all older versions.

The primary objective of publishing technology catalogues is to establish a uniform, commonly accepted and up-to-date basis for energy planning activities, such as future outlooks, evaluations of security of supply and environmental impacts, climate change evaluations, as well as technical and economic analyses, e.g. on the framework conditions for the development and deployment of certain classes of technologies.

With this scope in mind, it is not the target of the technology data catalogues, to provide an exhaustive collection of specifications on all available incarnations of energy technologies. Only selected, representative, technologies are included, to enable generic comparisons of technologies with similar functions in the energy system e.g. thermal gasification versus combustion of biomass or electricity storage in batteries versus fly wheels.

Finally, the catalogue is meant for international as well as Danish audiences in an attempt to support and contribute to similar initiatives aimed at forming a public and concerted knowledge base for international analyses and negotiations.

Data sources and results

A guiding principle for developing the catalogue has been to rely primarily on well-documented and public information, secondarily on invited expert advice. Where unambiguous data could not be obtained, educated guesses or projections from experts are used. This is done to ensure consistency in estimates that would otherwise vary between users of the catalogue.

Cross-cutting comparisons between technologies will reveal inconsistencies which may have several causes:

- Technologies may be established under different conditions. As an example, the costs of off-shore wind farms might be established on the basis of data from ten projects. One of these might be an R&D project with floating turbines, some might be demonstration projects, and the cheapest may not include grid connections, etc. Such a situation will result in inconsistent cost estimates in cases where these differences might not be clear.
- Investors may have different views on economic attractiveness and different preferences. Some decisions may not be based on mere cost-benefit analyses, as some might tender for a good architect to design their building, while others will buy the cheapest building.
- Environmental regulations vary from between countries, and the environment-related parts of the investment costs, are often not reported separately.
- Expectations for the future economic trends, penetration of certain technologies, prices on energy and raw materials vary, which may cause differences in estimates.
- Reference documents are from different years.

The ambition of the present publication has been to reduce the level of inconsistency to a minimum without compromising the fact that the real world is ambiguous. So, when different publications have presented different data, the publication which appears most in compliance with other publications has been selected as reference.

In order to handle the above-mentoned uncertainties, each catalogue contains an introductory chapter, stating the guidelines for how data have been collected, estimated and presented. These guidelines are not perfect, but they represent the best balance between various considerations of data quality, availability and usability.

Danish preface

Energistyrelsen udgiver kataloger med data om teknologier til produktion og konvertering af energibærer.

Kataloget løbende opdateres i takt med at teknologierne udvikler sig, hvis data ændrer sig væsentligt eller hvis der findes fejl. Alle opdateringer vil registreres i rettelsesbladet først i kataloget, og det vil altid være muligt at finde den seneste opdaterede version på Energistyrelsens hjemmeside, sammen med alle tidligere udgaver.

Hovedformålet med teknologikataloget er at sikre et ensartet, alment accepteret og aktuelt grundlag for planlægningsarbejde og vurderinger af forsyningssikkerhed, beredskab, miljø og markedsudvikling hos bl.a. de systemansvarlige selskaber, universiteterne, rådgivere og Energistyrelsen. Dette omfatter for eksempel fremskrivninger, scenarieanalyser og teknisk-økonomiske analyser.

Desuden er teknologikataloget et nyttigt redskab til at vurdere udviklingsmulighederne for energisektorens mange teknologier til brug for tilrettelæggelsen af støtteprogrammer for energiforskning og -udvikling. Tilsvarende afspejler kataloget resultaterne af den energirelaterede forskning og udvikling. Også behovet for planlægning og vurdering af klima-projekter har aktualiseret nødvendigheden af et opdateret databeredskab.

Endeligt kan teknologikataloget anvendes i såvel nordisk som internationalt perspektiv. Det kan derudover bruges som et led i en systematisk international vidensopbygning og -udveksling, ligesom kataloget kan benyttes som dansk udspil til teknologiske forudsætninger for internationale analyser og forhandlinger. Af disse grunde er kataloget udarbejdet på engelsk.

Med dette omfang i tankerne er det ikke målet for teknologidatakatalogerne at give en udtømmende samling af specifikationer for alle tilgængelige inkarnationer af energiteknologier. Kun udvalgte, repræsentative, teknologier er inkluderet, for at muliggøre generiske sammenligninger af teknologier med lignende funktioner i energisystemet.

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Guideline/Introduction

This catalogue presents technologies for generation and conversion of energy carriers. In particular: production of hydrogen by means of electrolysis, biofuels from biomass and production/upgrade of biogas/syngas, including further conversion such as power-to-X technologies.

Most of the process are characterised by multiple inputs and multiple outputs, which include for example different fuels/feedstocks, electricity and process heat (recoverable or lost).

Upstream and downstream processes are not included - the datasheets do not provide information on prices for fuels, environmental impact from fuel procurement, or the economic consequences of the substitution of fossil fuels with liquid fuels produced from biomass.

The main purpose of the catalogue is to provide generalized data for analysis of energy systems, including economic scenario models and high-level energy planning.

These guidelines serve as an introduction to the presentations of the different technologies in the catalogue, and as instructions for the authors of the technology chapters. The general assumptions are described in section 1.1. The following sections (1.2 and 1.3) explain the formats of the technology chapters, how data were obtained, and which assumptions they are based on. Each technology is subsequently described in a separate technology chapter, making up the main part of this catalogue. The technology chapters contain both a description of the technologies and a quantitative part including a table with the most important technology data.

General assumptions

The data presented in this catalogue is based on some general assumptions, mainly with regards to the utilization and start-ups of plants and technologies.

On the one hand, plants for biofuel production and production/upgrade of biogas and syngas are assumed to be designed and operated on a continuous basis along the year, except for maintenance and outages. Therefore, they feature a high number of full load hours (around 8000 h/y) and a reduced number of start-ups (5 per year).

On the other hand, electrolysers are assumed to be designed and operated for approximately 4000 full load hours annually. In particular, use the advantage of lower power prices by producing e.g. in hours of high renewable energy production (similarly to heat pumps). The assumed number of start-ups and consequent shut-downs for electrolysers, unless otherwise stated, is 50 per year.

Any exception to these general assumptions is documented in the relative technology chapter with a specific note.

Qualitative description

The qualitative part describes the key characteristics of the technology as concise as possible. The following paragraphs are included where relevant for the technology.

Contact information

Containing the following information:

Contact information: Contact details in case the reader has clarifying questions to the technology chapters. This could be the Danish Energy Agency, Energinet.dk or the author of the technology chapters.
Author: Entity/person responsible for preparing the technology chapters

Reviewer: Entity/person responsible for reviewing the technology chapters.

Brief technology description

Brief description for non-engineers of how the technology works and for which purpose.

An illustration of the technology is included, showing the main components and working principles.

Input

The main raw materials and other forms of energy consumed (e.g. electricity, heat) by the technology or facility. Moisture content of the fuel and required temperature of the input heat is specified.

Auxiliary inputs, such as enzymes or chemicals assisting the process are mentioned and their contribution described, if considered relevant.

Output

The output energy carrier as well as co-product or by-products, for example process heat. Temperature of the output heat is specified as well. Non-energy outputs may be stated as well, if relevant.

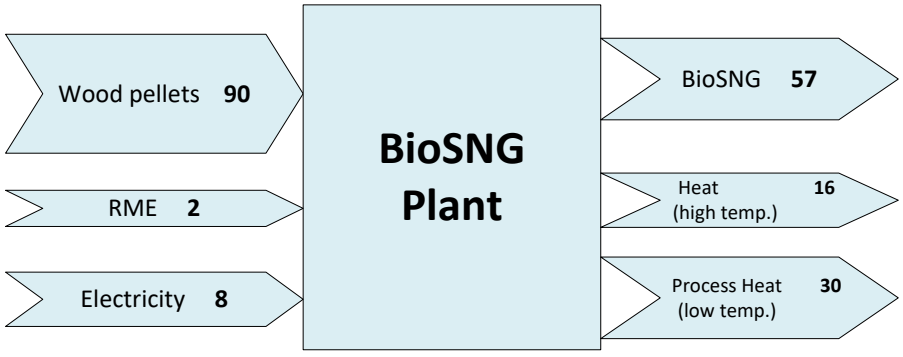
Energy balance

The energy balance shows the energy inputs and outputs for the technology. Here an illustrative diagram is shown based on data for the year 2015, thus currently available technology.

For process heat losses and produced energy carrier, it is important to specify information about temperature and pressure.

The first important assumption is that the energy content of all the fuels, both produced and consumed, is always expressed in terms of Lower Heating Value (LHV). As a consequence, because of the presence of some latent heat of vaporisation, the energy balance may result in a difference between total energy input and total energy output.

Figure 1: Example of Energy balance. All inputs sum up to 100 units.



Source: Danish Energy Agency

For comparison, 100 units of total input are used to standardize the diagrams. This choice allows the reader to easily calculate the efficiency for each of the output, which will be directly equal to the energy value in the balance.

Each of the inputs and outputs has to be accounted for in the diagram, including auxiliary electricity consumption in input and process heat losses in output.

Auxiliary products, as for example chemicals and enzymes, will in general only assist the process and are then not relevant for the energy balance. They should just be included as *auxiliary product input data*.

Typical capacities
The capacity, preferably a typical capacity (not maximum capacity), is stated for a single plant or generation facility. In case different sizes of plant are common, multiple technologies can be presented, e.g. Large, Medium and Small.

Regulation ability
Mainly relevant for hydrogen technologies where electricity is used as main input. Description of the part-load characteristics, how fast can they start up and how fast are they able to respond to supply changes and does part-load or fast regulation led to increased (or lower) wear and hence increased cost.

Space requirement
Space requirement is specified in 1,000 m^2 per MW of thermal (*Typical plant capacity*). The space requirements may for example be used to calculate the rent of land, which is not included in the financial cost, since this cost item depends on the specific location of the plant.

Advantages/disadvantages
A description of specific advantages and disadvantages relative to equivalent technologies. Generic advantages are ignored; e.g. renewable energy technologies mitigating climate risks and enhance security of supply.

Environment
Particular environmental and resource depletion impacts are mentioned, for example harmful emissions to air, soil or water; consumption of rare or toxic materials; issues with handling of waste and decommissioning etc.

The energy payback time or energy self-depreciation time may also be mentioned. This is the time required by the technology for the production of energy equal to the amount of energy that was consumed during the production of the technology.

Research and development perspectives
This section lists the most important challenges to further development of the technology. Also, the potential for technological development in terms of costs and efficiency is mentioned and quantified. Danish research and development perspectives are highlighted, where relevant.

Examples of market standard technology
Recent full-scale commercial projects, which can be considered market standard, are mentioned, preferably with links. A description of what is meant by “market standard” is given in the introduction to the quantitative description section (Section 1.3). For technologies where no market standard has yet been established, reference is made to best available technology in R&D projects.

Prediction of performance and costs
Cost reductions and improvements of performance can be expected for most technologies in the future. This section accounts for the assumptions underlying the cost and performance in the first technology year (base year) as well as the improvements assumed for furture years. For chapters published or updated after 2020, 2020 serves as base year for the technology instead of 2015, which has been the base year for several chapters.

The specific technology is identified and classified in one of four categories of technological maturity, indicating the commercial and technological progress, and the assumptions for the projections are described in detail.

In formulating the section, the following background information is considered:

Data for the base year

In case of technologies where market standards have been established, performance and cost data of recent installed versions of the technology in Denmark or the most similar countries in relation to the specific technology in Northern Europe are used for the base year estimates.

If consistent data are not available, or if no suitable market standard has yet emerged for new technologies, the base year costs may be estimated using an engineering-based approach applying a decomposition of manufacturing and installation costs into raw materials, labor costs, financial costs, etc. International references such as the IEA, NREL etc. are preferred for such estimates.

Assumptions for projecting costs into future years

According to the IEA:

"Innovation theory describes technological innovation through two approaches: the technology-push model, in which new technologies evolve and push themselves into the marketplace; and the market-pull model, in which a market opportunity leads to investment in R&D and, eventually, to an innovation" (ref. 6).

The level of "market-pull" is to a high degree dependent on the global climate and energy policies. Hence, in a future with strong climate policies, demand for e.g. renewable energy technologies will be higher, whereby innovation is expected to take place faster than in a situation with less ambitious policies. This is expected to lead to both more efficient technologies, as well as cost reductions due to economy of scale effects. Therefore, for technologies where large cost reductions are expected, it is important to account for assumptions about global future demand.

The IEA's Announced Pledges Scenario (APS) is used as a central estimate for projections in the Technology Catalogue, whenever possible. The IEA describes the Announced Pledges Scenario in their 2022 version as follows:

"The Announced Pledges Scenario introduced in 2021 aims to show to what extent the announced ambitions and targets, including the most recent ones, are on the path to deliver emissions reductions required to achieve net zero emissions by 2050. It includes all recent major national announcements as of September 2022 for 2030 targets and longer-term net zero and other pledges, regardless of whether these have been anchored in implementing legislation or in updated NDCs. In the APS, countries fully implement their national targets to 2030 and 2050, and the outlook for exporters of fossil fuels and low emissions fuels like hydrogen is shaped by what full implementation means for global demand. […] Non-policy assumptions, including population and economic growth, are the same as in the STEPS."

According to the IEA, the less ambitious Stated Policies Scenario (STEPS)

"… provides a more conservative benchmark for the future, because it does not take it for granted that governments will reach all announced goals. Instead, it takes a more granular, sector-by-sector look at what has actually been put in place to reach these and other energy-related objectives, taking account not just of existing policies and measures but also of those that are under development. The STEPS explores where the energy system might go without a major additional steer from policy makers."

The STEPS Scenario may be used as an upper bound and to assess the expected development of technologies based on a frozen-policy approach. Previous versions of the Technology Catalogue before updating the guideline in april 2024 have used the outdated New Policies Scenario, relatively equivalent to the current STEPS, as a central framework for projections (and supplemented by other outdated scenarios of the IEA). This scenario corresponds to the frozen-policy approach that the Danish Energy Agency uses to project international fuel prices and CO_2-prices and technologies may be assessed in that regard when suitable.

Technologies updated before this cutoff date and which do not contain any explicit methodological description within the chapter regarding alternative supplementary scenarios have been updated based in this previous methodology.

As a more ambitious projection, the Net Zero Emissions by 2050 Scenario (NZE) may be used as a lower bound for the technology development. According to the IEA, the NZE

"… is a normative IEA scenario that shows a pathway for the global energy sector to achieve net zero CO2 emissions by 2050, with advanced economies reaching net zero emissions in advance of others. This scenario also meets key energy-related United Nations Sustainable Development Goals (SDGs), in particular by achieving universal energy access by 2030 and major improvements in air quality. It is consistent with limiting the global temperature rise to 1.5 °C with no or limited temperature overshoot (with a 50% probability), in line with reductions assessed in the IPCC in its Sixth Assessment Report."

By using this approach, the quantitative data in the Technology Catalogue provides a sample space that is consistent with the IEA's Global Energy and Climate Model, encompassing relevant outcomes for policy assessments of technologies as well as technology developments in compliance with national targets, and international treaties.

Learning curves and technological maturity

Predicting the future costs of technologies may be done by applying a cost decomposition strategy, as mentioned above, decomposing the costs of the technology into categories such as labor, materials, etc. for which predictions already exist. Alternatively, the development could be predicted using learning curves. Learning curves express the idea that each time a unit of a particular technology is produced, learning accumulates, which leads to cheaper production of the next unit of that technology. The learning rates also consider benefits from economy of scale and benefits related to using automated production processes at high production volumes.

The potential for improving technologies is linked to the level of technological maturity. The technologies are categorized within one of the following four levels of technological maturity.

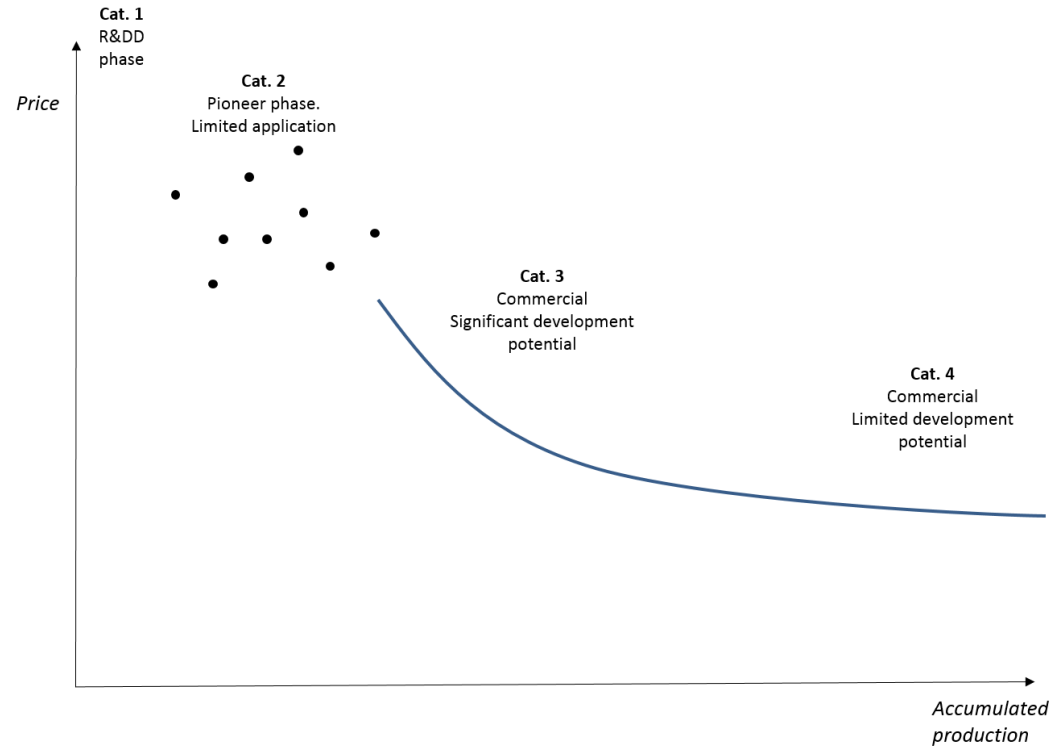
Category: Technologies that are still in the *research and development phase*. The uncertainty related to price and performance today and in the future is highly significant (e.g. wave energy converters, solid oxide fuel cells).

Category 2: Technologies in the *pioneer phase*. The technology has been proven to work through demonstration facilities or semi-commercial plants. Due to the limited application, the price and performance is still attached with high uncertainty, since development and customization is still needed. The technology still has a significant development potential (e.g. gasification of biomass).

Category 3: *Commercial technologies with moderate deployment*. The price and performance of the technology today is well known. These technologies are deemed to have a certain development potential and therefore there is a considerable level of uncertainty related to future price and performance (e.g. offshore wind turbines)

Category 4: *Commercial technologies, with large deployment*. The price and performance of the technology today is well known and normally only incremental improvements would be expected. Therefore, the future price and performance may also be projected with a relatively high level of certainty (e.g. coal power, gas turbine).

Figure 2: Technological development phases, correlation between accumulated production volume (MW) and price.



Source: Danish Energy Agency

Uncertainty

The catalogue covers both mature technologies and technologies under development. This implies that the price and performance of some technologies may be estimated with a relatively high level of certainty whereas in the case of others, both cost and performance today as well as in the future are associated with high levels of uncertainty.

This section of the technology chapters explains the main challenges to precision of the data and identifies the areas on which the uncertainty ranges in the quantitative description are based. This includes technological or market related issues of the specific technology as well as the level of experience and knowledge in the sector and possible limitations on raw materials. The issues should also relate to the technological development maturity as discussed above.

The level of uncertainty is illustrated by providing a lower and higher bound beside the central estimate, which shall be interpreted as representing probabilities corresponding to a 90% confidence interval. It should be noted, that projecting costs of technologies far into the future is a task associated with very large uncertainties. Thus, depending on the technological maturity expressed and the period considered, the confidence interval may be very large. It is the case, for example, of less developed technologies (category 1 and 2) and longtime horizons (2050).

Additional remarks

This section includes other information, for example links to web sites that describe the technology further or give key figures on it.

References

References are numbered in the text in squared brackets and bibliographical details are listed in this section.

Quantitative description

To enable comparative analyses between different technologies it is imperative that data are actually comparable: All cost data are stated in fixed 2015 prices excluding value added taxes (VAT) and other taxes. The information given in the tables relate to the development status of the technology at the point of final investment decision (FID) in the given year (2015, 2020, 2025, 2030, 2035, 2040 and 2050 where applicable). FID is assumed to be taken when financing of a project is secured and all permits are at hand. The year of commissioning will depend on the construction time of the individual technologies.

A typical table of quantitative data is shown below, containing all parameters used to describe the specific technologies. The datasheet consists of a generic part, which is identical for all technologies and a technology specific part, containing information, which is only relevant for the specific technology. The generic part is made to allow for easy comparison of technologies.

It has to be noted that, in case a technology has more than one input or output, rows will be added to the datasheet.

Each cell in the table contains only one number, which is the central estimate for the market standard technology, i.e. no range indications.

Uncertainties related to the figures are stated in the columns named *uncertainty*. To keep the table simple, the level of uncertainty is only specified for the base year and the final year.

The level of uncertainty is illustrated by providing a lower and higher bound. These are chosen to reflect the uncertainties of the best projections by the authors. The section on uncertainty in the qualitative description for each technology indicates the main issues influencing the uncertainty related to the specific technology. For technologies in the early stages of technological development or technologies especially prone to variations of cost and performance data, the bounds expressing the confidence interval could result in large intervals. The uncertainty only applies to the market standard technology. The uncertainty interval does not represent the product range (for example a product with lower efficiency at a lower price or vice versa).

The level of uncertainty is stated for the most critical figures such as investment cost and specific output shares. Other figures are considered if relevant.

All data in the tables are referenced by a number in the utmost right column (Ref), referring to source specifics below the table. The following separators are used:

; (semicolon) separation between the time horizons (2015, 2020, etc.)

/ (forward slash) separation between sources with different data

+ (plus) agreement between sources on same data

Notes include additional information on how the data are obtained, as well as assumptions and potential calculations behind the figures presented. Before using the data, please be aware that essential information may be found in the notes below the table.

The generic parts of the datasheets for energy carrier generation and conversion technologies are presented below:

Table 2: Generic parts of datasheet for energy carrier generation and conversion technologies

Technology	Name/Description									
	2015[1]	2020[1]	2030[1]	2050[1]	Uncertainty (2020[1])		Uncertainty (2050[1])		Note	Ref
Energy/technical data					Lower	Upper	Lower	Upper		
Typical total plant size (MW output)										
- Inputs										
A) Energy input share (% total input(MWh/MWh))										
B) Energy input share (% total input(MWh/MWh))										
C) Energy input share (% total input(MWh/MWh))										
X) Auxiliary products inputs (kg/MWh)										
Y) Auxiliary products inputs (kg/MWh)										
- Outputs										
A) Output share (% total input (MWh/MWh))										
B) Output share (% total input (MWh/MWh))										
C) Output share (% total input (MWh/MWh))										
X) Non-energy outputs (kg/MWh)										
Y) Non-energy outputs (kg/MWh)										
Forced outage (%)										
Planned outage (weeks per year)										
Technical lifetime (years)										
Construction time (years)										
Financial data										
Specific investment (€ /MW of total input)										
- hereof equipment (%)										
- hereof installation (%)										
Fixed O&M (€ /MW of total input)										
Variable O&M (€/MWh of total input)										
Startup cost (€ /MW of total input per startup)										

Technology specific data										

Source: Danish Energy Agency. Ann.: [1]Technology years may be updated from this shown example.

Energy/technical data

Typical total plant size

The total thermal capacity, preferably a typical capacity, is stated for a single plant or facility. It represents the sum of all input and is expressed in MW thermal.

Input

All inputs that contribute to the energy balance are included as *main energy input* and are expressed as percentage in relation to the total energy input, or equivalently as MWh/MWh of total input.

The energy inputs (and outputs) are always expressed in lower heating value (LHV) and moisture content considered is specified if relevant.

Auxiliary inputs, such as enzymes or chemicals that are assisting the process but do not contribute to the energy balance are included as *auxiliary products* (under *input*) and are expressed in kg/MWh of total energy input.

Output

Similarly, to the energy inputs, energy outputs are expressed as percentage value in relation to the total energy input, or equivalently as MWh/MWh of total input.

Any energy co-product or by-product of the reaction has to be specified within the outputs, including process heat loss. Since fuel inputs are measured at lower heating value, in some cases the total efficiency may exceed or be lower than 100%.

The output shares represent the partial efficiencies in producing the different outputs.

The process heat (output) is, if possible, separated in recoverable (for example for district heating purposes) and unrecoverable heat and the temperatures are specified.

Forced and planned outage

Forced outage is defined as the number of weighted forced outage hours divided by the sum of forced outage hours and operation hours. The weighted forced outage hours are the sum of hours of reduced production caused by unplanned outages, weighted according to how much capacity was out.

Forced outage is given in percent, while planned outage (for example due to renovations) is given in days per year.

Technical lifetime

The technical lifetime is the expected time for which an energy plant can be operated within, or acceptably close to, its original performance specifications, provided that normal operation and maintenance takes place. During this lifetime, some performance parameters may degrade gradually but still stay within acceptable limits. For instance, power plant efficiencies often decrease slightly (few percent) over the years, and O&M costs increase due to wear and degradation of components and systems. At the end of the technical lifetime, the frequency of unforeseen operational problems and risk of breakdowns is expected to lead to unacceptably low availability and/or high O&M costs. At this time, the plant is decommissioned or undergoes a

lifetime extension, which implies a major renovation of components and systems as required making the plant suitable for a new period of continued operation.

The technical lifetime stated in this catalogue is a theoretical value inherent to each technology, based on experience. As explained in the *General Assumptions*, different types of plants are designed for a different annual utilization and typical number of start-ups a year. The expected technical lifetime considers these assumptions.

In real life, specific plants of similar technology may operate for shorter or longer times. The strategy for operation and maintenance, e.g. the number of operation hours, start-ups, and the reinvestments made over the years, will largely influence the actual lifetime.

Construction time

Time from final investment decision (FID) until commissioning completed (start of commercial operation), expressed in years.

Financial data

Financial data are all in Euro (€), real prices, at the 2020-level and exclude value added taxes (VAT) and other taxes.

Several data originate in Danish references. For those data a fixed exchange ratio of 7.45 DKK per € has been used.

The first catalogue was in 2011 prices. Some data had been updated by applying the general inflation rate in Denmark (2011 prices have been multiplied by 1.0585 to reach the 2015 price level).

Similarly, prices in 2015-Euro were multiplied by 1.0634 to update them to 2020 prices.

European data, with a particular focus on Danish sources, have been emphasized in developing this catalogue. This is done as generalizations of costs of energy technologies have been found to be impossible above the regional or local levels, as per IEA reporting from 2020 (ref. 3). For renewable energy technologies this effect is even stronger as the costs are widely determined by local conditions.

Investment costs

The investment cost is also called the engineering, procurement and construction (EPC) price or the overnight cost. Infrastructure and connection costs, i.e. electricity, fuel and water connections inside the premises of a plant, are also included.

The investment cost is reported on a normalized basis, i.e. cost per MW. The specific investment cost is the total investment cost divided by the *Typical total plant size* described in the quantitative section.

Where possible, the investment cost is divided on equipment cost and installation cost. Equipment cost covers the components and machinery including environmental facilities, whereas installation cost covers engineering, civil works, buildings, grid connection, installation and commissioning of equipment. Cost may be disaggregated in a more detailed cost breakdown if it improves readability or understanding of the given technology.

The rent of land is not included but may be assessed based on the space requirements, if specified in the qualitative description, and if the cost is a noteworthy component in the developer's scope. In that case land rent can be given as either upfront investment cost or yearly rent.

The owners' predevelopment costs (administration, consultancy, project management, site preparation, approvals by authorities) and interest during construction are not included, unless specifically mentioned by a separate parameter in within the cost breakdown. The costs to dismantle decommissioned plants are also not included. Decommissioning costs may be offset by the residual value of the assets.

Contingency

Project owners often add a contingency to a project's capital cost estimate to deal with project overruns due to uncertainties and risks caused by uncertainties in the project definition.
The Association for the Advancement of Cost Engineering International (AACE International) has defined contingency as "*An amount added to an estimate to allow for items, conditions, or events for which the state, occurrence, or effect is uncertain and that experience shows will likely result, in aggregate, in additional costs. Typically estimated using statistical analysis or judgment based on past asset or project experience*.". AACE International further describes contingency as "*…planning and estimating errors and omissions…..design developments and changes within the scope, and variations in market and environmental conditions**. The Technology Catalogues represent general techno-economic data for different technologies; and are not intended as basis for investment decisions. Therefore, the data in the Technology Catalogues aim at not including contingency.

*Source: AACE (2022) Cost engineering terminology (https://library.aacei.org/terminology/welcome.shtml).

Cost of grid expansion

The costs for the connection of the plant to the system are included in the investment cost, while no cost of grid expansion or reinforcement is considered in the present data.

Business cycles

Historic costs of energy equipment can show fluctuations that are related to business cycles. This was the case of the period 2007-2008 for example or more recently around 2021-2022, where prices costs of many energy generation technologies increased dramatically driven by rapid increases in global raw material costs and supply chain costs. The primary objective of the technology catalogues is to establish general representative techno-economic data for different technologies, which can form a basis for energy planning activities and technical and economic analyses. The catalogues do not attempt to reflect fluctuations in technology costs due to fluctuations in costs of labour and materials driven by e.g. global/regional crises or major events affecting short term supply or demand. The technology cost developments in the catalogues thus intend to reflect an average business cycle situation and macroeconomic environment in a general long-term equilibrium.

Economy of scale

The main idea of the catalogue is to provide technical and economic figures for particular sizes of plants. Where plant sizes vary in a large range, different sizes are defined and separate technology chapters are developed.

For assessment of data for plant sizes not included in the catalogue, some general rules should be applied with caution to the scaling of plants.

The cost of one unit for larger power plants is usually less than that for smaller plants. This is called the 'economy of scale'. The basic equation (ref. 2) is:

$$\frac{C_1}{C_2} = \left(\frac{P_1}{P_2}\right)^a$$

Where:

C_1 = Investment cost of plant 1 (e.g. in million EUR)
C_2 = Investment cost of plant 2
P_1 = Power generation capacity of plant 1 (e.g. in MW)
P_2 = Power generation capacity of plant 2
a = Proportionality factor

Usually, the proportionality factor is about 0.6 – 0.7, but extended project schedules may cause the factor to increase. It is important, however, that the plants are essentially identical in construction technique, design, and construction time frame and that the only significant difference is in size.

For technologies that have a more modular structure, such as electrolysers, the proportionality factor is equal to 1.

The relevant ranges where the economy of scale correction applies are stated in the notes for the capacity field of each technology table. The stated range represents typical capacity ranges.

Operation and maintenance (O&M) costs

The fixed share of O&M is calculated as cost per plant size (€ per MW per year), where the typical total plant size is the one defined at the beginning of this chapter and stated in the tables. It includes all costs, which are independent of how the plant is operated, e.g. administration, operational staff, payments for O&M service agreements, network use of system charges, property tax, and insurance. Any necessary reinvestments to keep the plant operating within the scheduled lifetime are also included, whereas reinvestments to extend the life beyond the lifetime are excluded. Reinvestments are discounted at 4 % annual discount rate in real terms. The cost of reinvestments to extend the lifetime of the plants may be mentioned in a note if the data has been readily available.

The variable O&M costs (€/MWh) include consumption of auxiliary materials (water, lubricants, fuel additives), treatment and disposal of residuals, spare parts and output related repair and maintenance (however not costs covered by guarantees and insurances).

Planned and unplanned maintenance costs may fall under fixed costs (e.g. scheduled yearly maintenance works) or variable costs (e.g. works depending on actual operating time), and are split accordingly.

All costs related to the process inputs (electricity, heat, fuel) are not included.

It should be noticed that O&M costs often develop over time. The stated O&M costs are therefore average costs during the entire lifetime.

Start-up costs

The O&M costs stated in this catalogue includes start-up costs and considers a typical number of start-ups and shut-downs. Therefore, the start-up costs should not be specifically included in more general analyses. They should only be used in detailed dynamic analyses of the hour-by-hour load of the technology.

Start-up costs, are stated in costs per MW of typical plant size (€/MW/startup), if relevant. They reflect the direct and indirect costs during a start-up and the subsequent shut down.

The direct start-up costs include fuel consumption, e.g. fuel which is required for heating up boilers and which does not yield usable energy, electricity consumption, and variable O&M costs corresponding to full load during the start-up period.

The indirect costs include the theoretical value loss corresponding to the lifetime reduction for one start up. For instance, during the heating-up, thermal and pressure variations will cause fatigue damage to components, and corrosion may increase in some areas due to e.g. condensation.

An assumption regarding the typical number of start-ups is made for each technology in order to calculate the O&M costs. As a general assumption, biofuel production and production/upgrade of biogas features 5 start-ups per year, while for electrolyzes 50 start-ups a year are assumed. Any change with respect to this general assumption, e.g. for a specific technology which is characterized by a different utilization, is specified in the notes.

The stated O&M costs may be corrected to represent a different number of start-ups than the one assumed by using the stated start-up costs with the following formula:

$$O\&M_{new} = O\&M_{old} - \left(Startup\ cost * n_{startup}^{old}\right) + \left(Startup\ cost * n_{startup}^{new}\right)$$

where $n_{startup}^{old}$ is the number of start-ups specified in the notes for the specific technology and $n_{startup}^{new}$ is the desired number of start-ups.

Technology specific data

Additional data is specified in this section, depending on the technology.

For example, operating temperatures are indicated for electrolysis and other processes in which it is a relevant parameter.

Whenever process heat is available as output, its temperature is specified as well.

For electrolysis technologies, parameters regarding the regulation ability are specified as follow:

Ramp up time, linear to full load (minutes)
Ramp down time, linear from full load (minutes)
Start-up time (minutes)
Minimum load (%)

Relevant emissions to the environment, including emissions to water and air, are reported in g per MWh of total input of fuel at the lower heating value.

All plants are assumed to be designed to comply with the environmental regulation that is currently in place in Denmark and planned to be implemented within the 2020-time horizon.

Definitions

The latent heat of vaporization is the heat absorbed when a substance changes phase from liquid to gas.

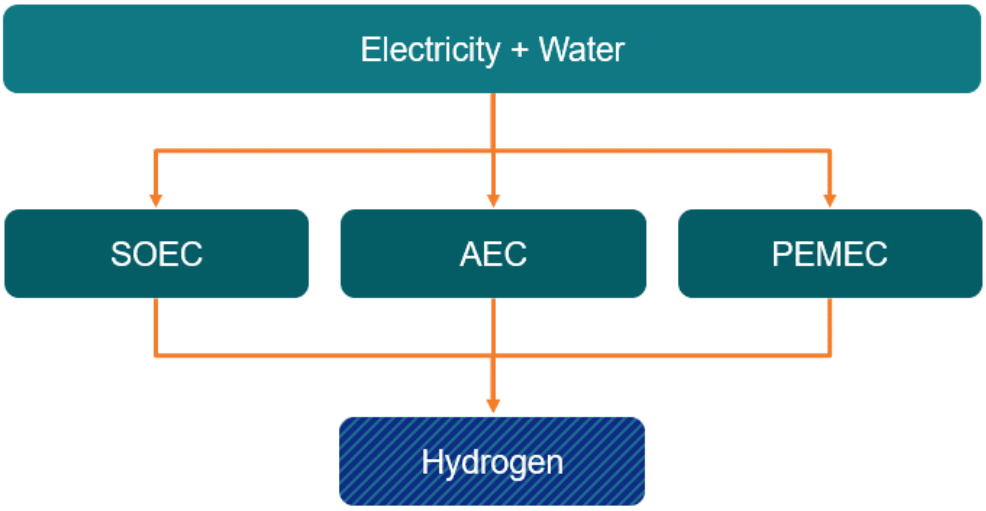
The lower heating value (also known as net calorific value) of a fuel is defined as the amount of heat released by combusting a specified quantity (initially at 25°C) and returning the temperature of the combustion products to 150°C, which assumes the latent heat of vaporization of water in the reaction products is not recovered. The LHV are the useful calorific values in boiler combustion plants and are frequently used in Europe.

Using the LHV for efficiency definition, a condensing boiler can achieve a thermal efficiency of more than 100%, because the process recovers part of the heat of vaporization.

The higher heating value (also known as gross calorific value or gross energy) of a fuel is defined as the amount of heat released by a specified quantity (initially at 25°C) once it is combusted and the products have returned to a temperature of 25°C, which considers the latent heat of vaporization of water in the combustion products.

When using HHV for thermal efficiency definition, the thermodynamic limit of 100% cannot be exceeded.

Hydrogen



This section describes the technology data for the production of hydrogen using electrolysis starting from water and electricity, as shown in the figure above. It includes the three main technologies, namely alkaline, proton membrane and solid oxide electrlysers. The chapter does not include other hydrogen production pathways.

The hydrogen produced in this pathway can be used directly or as feedstock in one of the other pathways highlighted in this catalogue.

The section includes only one chapter:

80 Electrolysis

80 Electrolysis

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Amendments after publication date

Date	Ref	Description
08/2026	DNV	Restructured text to fit the guideline, updates to text regarding the technical and economic data to reflect the new data sheets and industry advancements. Revised architectural configurations, system boundaries, flexibility potentials and new section on offshore electrolysis.
01/2024	Rambøll	Major update with new qualitative text and revised technical and economic data

Quantitative description

The upcoming decades are anticipated to witness a progressive growth in the demand for renewable hydrogen, driven by a commitment to transitioning our energy system towards a low carbon framework. Among the various forms of low carbon hydrogen available, water electrolysis has become the most prevalent technique due to its lower emissions throughout the entire process. When combined with 100% renewable energy sources, this method gives rise to green hydrogen, a sustainable, CO_2 emission-free hydrogen variant.

In recent years, there has been considerable focus on green hydrogen production, with significant emphasis placed on the core equipment involved in water electrolysis, namely

the electrolysis systems. However, the hydrogen production plants are an infrastructure that extends beyond electrolysers.

The offtake uses of the hydrogen impact its specific requirements, such as flow rates, purity or pressure, which in turn influence the production and demand of this fuel. Thus, this chapter presents a comprehensive overview of a green hydrogen production plants.

Brief technology description

Hydrogen Plants

A hydrogen production plant produces hydrogen from electricity and water as primary inputs. The required output is specified by the offtake application and typically includes a production volume target (in tonnes per hour, day or year), a hydrogen purity specification (water and oxygen content), and a delivery pressure and temperature. The offtake application also determines the required dynamic behaviour, for example whether the plant operates as baseload, follows variable renewable generation, or provides grid services.

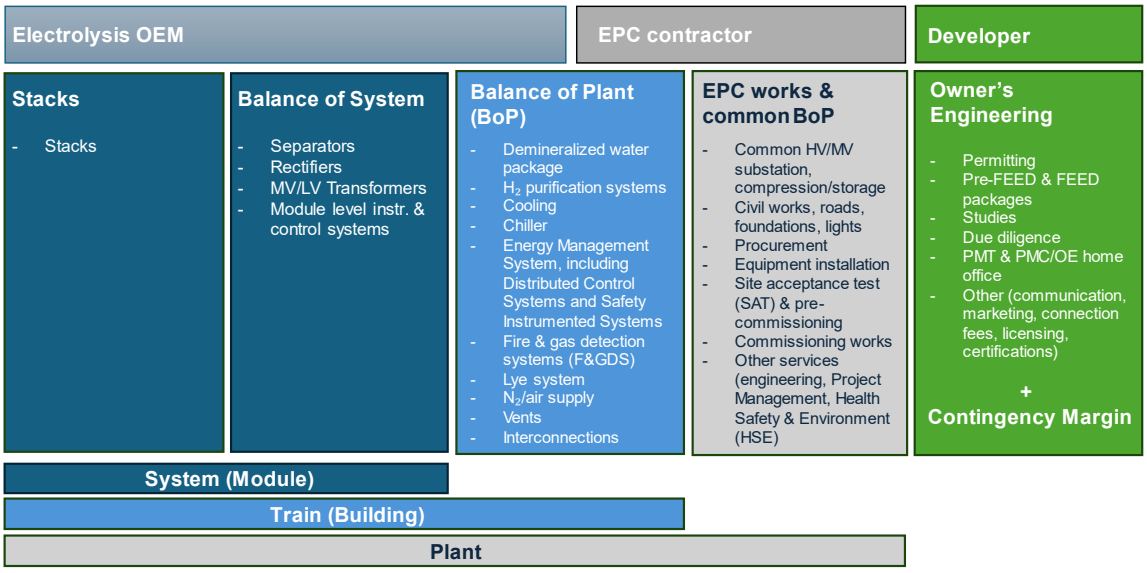
Two distinct architectural configurations are used in the industry to deliver hydrogen plants at the multi-MW to multi-100-MW scale. In the chemical-plant configuration the plant is engineered as a bespoke chemical process facility, with the scope of supply distributed between the electrolyser OEM, an EPC contractor, and the developer with its independent engineer. This configuration is the most common in large scale projects typically above 100 MW due to flexibility in the design and possibilities to achieve economies of scale. In the prefabricated configuration, the plant is composed of a number of standardised, type-approved electrolyser units with the unit OEM delivering an integrated product that includes most of what would be balance of system and some elements from the balance of plant in the chemical-plant configuration. This configuration is common in small to medium scale projects, typically below 100 MW.

The two configurations imply different definitions of the OEM, EPC, and developer scopes, different cost structures, and different scaling logic. The remainder of this chapter describes both, noting where the descriptions differ between architectures.

Hydrogen Plants, chemical-plant configuration

Figure 1 shows the typical scope of supply for the chemical-plant configuration. This normally exceeds the electrolyser OEM's product offering (typically limited to stacks, balance of system and part of the balance of plant, the rest being covered by an EPC contractor and the developer with their independent engineer and advisors).

Figure 3. Chemical-plant configuration equipment and works disaggregation to form the total CAPEX.



(*) EMS (global): main programmable logic controllers (PLC) with supervisory control and data acquisition (SCADA) system and databases interconnected with Power Purchase Agreement (PPA) input data, electrolysis PLC, H2 storage, demand side, etc.

As a result, the following definitions are usual in the hydrogen industry:

- Stack. Assembly of several cells in series, consolidating a unit which is capable of producing hydrogen when direct current (DC) is applied to it.
- System (also referred as 'Module'). Assembly of several stacks with common balance of system.
- Balance of system (BoS). Common peripherals included in an electrolysis module. Normally it includes:
  - Gas-liquid separators
  - H2 compression at module level if applicable (e.g. atmospheric technology),
  - Rectifiers
  - MV/LV transformers
  - Module- and stack level instrumentation and control
  - Pipes and fittings
  - Cables and bus bars

  The industry sometimes differentiates between balance of system and power supply unit (the latter comprising the abovementioned power electronics, i.e. rectifiers and MV/LV transformers).
- Trains (also identified as electrolysis buildings). Assembly of several modules with common Balance of Plant.
- Balance of plant (BoP). Common peripherals included in an electrolysis train. Normally it includes:
  - Demineralised water unit
  - Chiller
  - Water cooling unit
  - H2 purification (depending on the OEM it may be included as Balance of System also)
  - Water supply pumps and tank
  - Electrolyte circulation pump and tank (in case of AEC)

  - Instrument Air compressors
  - Nitrogen supply
  - Vents
  - Energy Management System (EMS) incl. Distributed Control System (DCS) and Safety Instrumented System (SIS)
  - Fire and Gas Detection Systems (F&GDS).
  - Pipelines, fittings, cables, other instrumentation and equipment
  - HV/MV substation
  - Interconnections (electrical, mechanical, piping and instrumentation/control, I&C)
- Hydrogen plant. Infrastructure composed of several electrolysis trains plus additional peripherals. Those peripherals include (often considered also in BoP):
  - Centralized gaseous hydrogen buffer storage (if required)
  - Centralized hydrogen compression (if required)
  - Civil works, roads, foundations, lights and additional interconnections

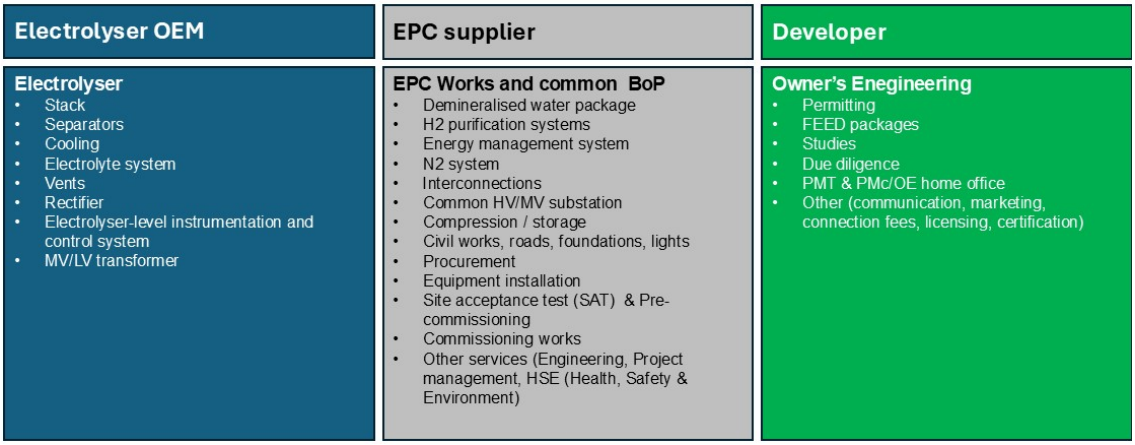
For instance, a common typical configuration offered to consolidate a pressurized alkaline plant of 500 MW in the industry would be:

- 100 x 5 MW stacks
- 25 x 20 MW modules (each module composed of 4 x 5 MW stacks)
- 5 x 100 MW trains (each train composed of 5 x 20 MW modules)

Hydrogen Plants, prefabricated configuration

In the prefabricated configuration the plant is not engineered as a bespoke chemical process facility but is composed of a number of standardised, type-approved electrolyser units. The unit, rather than the stack, is the basic building block, and the plant is built up by repetition of identical units rather than by scaling a single train. This changes the scope of supply, because some elements that the chemical-plant configuration distribute across the balance of system and part of balance of plant is brought inside the unit and delivered by the unit OEM as part of an integrated product. Note, that this configuration is not quantified in the attached data sheets.

Figure 4. Prefabricated plant configuration equipment and works disaggregation to form the total CAPEX.

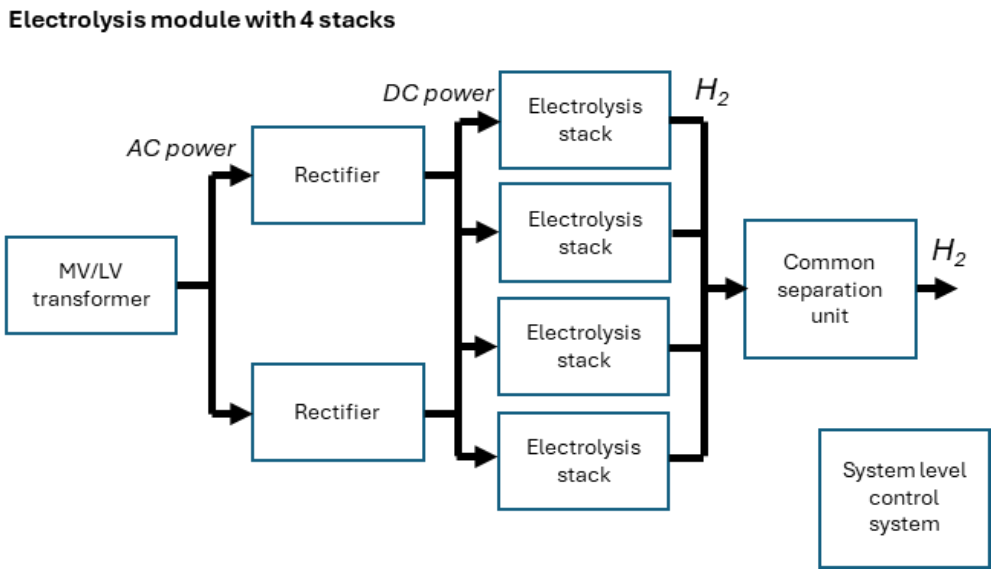


Electrolysis systems

Figure 5 provides a comprehensive overview of the electrolyser system, delineating its key components: the stacks and the balance of system. As mentioned, a system or module includes the stacks (ensemble of cells where the electrochemical reaction takes place) and

all necessary components consolidated into a single unit, such as connectors and piping, collectively referred to as the balance of the stacks. Despite stacks varying in size from a kilowatt to multiple megawatts (depending on the technology and vendor), electrolysis modules remain in a comparable order of magnitude, measured in megawatts.

Figure 5: Schematic of the different components included in an electrolyser system: example for a 4 x stacks module with 1 x MV/LV transformer, 2 x rectifiers and 1 x separation system.

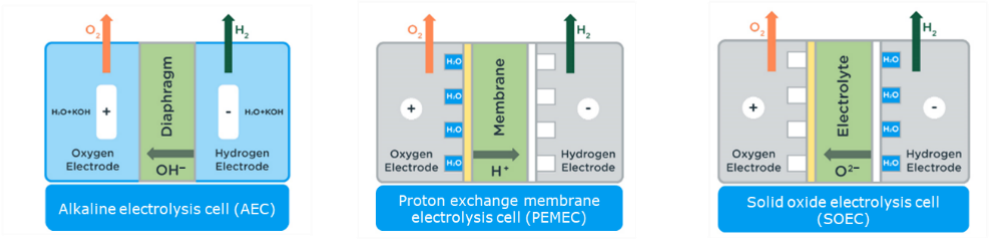


Types of electrolyser stacks

This chapter will focus on the three primary types of electrolyzer stacks and their associated cell technologies: alkaline electrolysis cells (AEC), proton exchange membrane electrolysis cells (PEMEC), and solid oxide electrolysis cells (SOEC). The three different technologies are illustrated in Figure 6.

A distinction is maintained between the electrolysis cell, the electrolysis process, and the equipment. The acronyms AEC, PEMEC and SOEC denote the three cell types, the alkaline electrolysis cell, the proton exchange membrane electrolysis cell, and the solid oxide electrolysis cell, and are used only where the cell and its electrochemistry are meant. The corresponding technologies, and the equipment built from these cells, are referred to as alkaline, PEM and solid oxide electrolysers, and the underlying process as alkaline, PEM and solid oxide electrolysis. Where the scale of the equipment matters, the level is stated explicitly, as electrolyser stack, electrolyser system, electrolyser train, or electrolyser plant.

Figure 6: The three primary electrolyser technologies: AEC, PEMEC, and SOEC.



Type	Low temperature		High temperature
	Alkaline electrolysis cell (AEC)	Proton exchange membrane electrolysis cell (PEMEC)	Solid oxide electrolysis cell (SOEC)

Regardless of the specific technology, the fundamental chemical reaction remains consistent across all types. The operational principle for these three technologies lies in the process of breaking the water molecule using electricity, also known as electrolysis. As a result, hydrogen and oxygen are produced, as exemplified in the following reaction:

$$2\,H_2O \rightarrow 2\,H_2 + O_2$$

The electrochemical reactions at the hydrogen and oxygen electrodes for each technology varies slightly, specifically the charge carrying ion that is transported across a selective barrier differs from OH^- (hydroxide ion) in case of alkaline, to H^+ (proton) in case of PEM, to O^{2-} (oxide ion) in case of SOEC, as shown in Table 3.

Table 3: The reactions at the hydrogen electrode and oxygen electrode for each of the three electrolysis technologies.

	Hydrogen electrode/Cathode	Oxygen electrode/Anode
AEC	$2H_2O + 2e^- \rightarrow H_2 + 2OH^-$	$2OH^- \rightarrow \frac{1}{2}O_2 + H_2O + 2e^-$
PEMEC	$2H^+ + 2e^- \rightarrow H_2$	$H_2O \rightarrow \frac{1}{2}O_2 + 2e^- + 2H^+$
SOEC	$2H_2O + 4e^- \rightarrow 2H_2 + 2O^{2-}$	$2O^{2-} \rightarrow O_2 + 4e^-$

Ann: In water electrolysis, common terminology includes referring to the electrodes as the anode and cathode. The cathode is associated with the reduction side, where the reduction reactions take place, while the anode is associated with the oxidation side, where oxidation reactions occur. Alternatively, different terminology may use the terms 'hydrogen electrode' and 'oxygen electrode'. This terminology allows to understand better each side when fuel cells or reversible systems are also considered. Table 3 shows the partial reaction taking place for every technology in each electrode.

The three electrolyser technologies have some key differences in characteristics in terms of stack efficiency, lifetime and electrolyser unit footprint. These have been summarised in Table 4.

Table 4: State-of-the-art characteristics of different electrolysis technologies (representing status as of 2025)

Status 2025					
Parameter	Units	AEC	PEMEC	SOEC	Ref.
Critical raw materials	Chemical elements	Ni	Pt, Ti, Ir	Co, Ni	
Max commercial stack size| max H_2 output	MWe | kgH_2/h	5 | 91	1.5 | 27	0.01 | 0.25	DNV**
Average stack efficiency	$kWhe/kgH_2$ (% LHV)	52 (64%)	51 (65%)	37* (90%)	[1] [2] [3] [4] [5]
Average stack + BOP efficiency	$kWhe/kgH_2$ (% LHV)	57 (58%)	58 (57%)	40* (83%)	[1] [2] [3] [4] [5]
Average stack lifetime[1]	h	70,000	60,000	25,000	[5] [6], [7]
Electrolyser Stack Footprint	m^2/MW	10	7.5	20	DNV**

Ann.: Electrolyser stack + BoP efficiency is defined as total energy consumption of components including stacks, balance of system and balance of plant.
*Steam input at 150 °C. Efficiency value excludes the energy consumption related to steam input. DNV** = DNV estimate from industry insights.

Alkaline electrolysis

Alkaline electrolysis cells (AEC) are the most mature electrolysis technology, and they use a liquid electrolyte (potassium hydroxide, KOH). The main characteristics of AEC are presented in Table 4. Both large stacks and electrolyser systems can be achieved with the use of pressurised AEC technology, with stacks as large as 5 MW with a maximum output hydrogen flow rate of 91 kg/h and systems larger than 500 MW, with a stack lifetime of 70,000 h (2025) [6], [7]. These stacks allow the design of electrolysis units between 10 MW and 25 MW with the possibility to have their own balance of the system and power electronics.

Regarding the materials used in AEC, pure nickel (Ni), Ni-plated carbon steel and stainless steels are the most common materials, with some expensive and platinum-family metals such as ruthenium (Ru) or iridium (Ir) being significant in some of the solutions offered in the market, although not in the majority. Recent calculations made by the International Energy Agency estimate that AEC uses around 800 kg/MW of Ni [8] if fitted with solid nickel electrodes. If plated or stainless-steel electrodes are used, the specific mass of Ni per MW is significantly lower. Ni-plated carbon steel or stainless steels are proposed to

[1] Stack lifetime is defined as the number of hours of operation required so that the stack electrical efficiency reaches 90% of the beginning of life value.

replace pure Ni in all components, owing to the relative cost and projected scarcity of Ni in the coming decades. Balance of system components, such as electrolyte tanks or gas separators, are mainly made of stainless steel or Ni-plated carbon steel. In addition, stainless steel is used for electrolyser system tubing. Finally, inexpensive catalysts such as Raney Ni, as well as Ni-, iron (Fe)- and copper (Cu)-containing alloys, are the most common materials used as catalysts [6].

Proton exchange membrane electrolysis

Proton exchange membrane electrolysis cells (PEMEC) are characterised by having a solid electrolyte and by operating at much higher current densities, resulting in a significantly smaller electrolyser system footprint. With a relatively high output pressure of ca. 30 bar, it produces high-purity hydrogen (99.999%). Table 4 shows the main characteristics of this technology. Stack sizes are smaller than for alkaline electrolysers, with current maximum commercial sizes at 1.5 MW and 27 kg/h of produced hydrogen (H_2). Plant sizes above 100 MW are nevertheless reached by combining many stacks, as for the other technologies. As shown in Table 2, the stack-level electricity consumption is marginally lower than for AEC, while the consumption at system level, including the balance of system and balance of plant, is slightly higher, and the average stack lifetime is shorter (60,000 h as per 2025) [6], [7].

Regarding the materials, PEMEC is the most demanding of the three technologies in terms of raw materials, as it uses the platinum-group metals platinum (Pt) and iridium (Ir). These metals are scarce, and iridium in particular, which is produced only as a by-product of platinum mining at a few tonnes per year, is a potential constraint on large-scale deployment (>100 MW projects and beyond). Titanium (Ti) is used in some of the stack constituents, such as bipolar plates and porous transport layers (PTLs), due to its good performance and stability in the service conditions (high potentials in acidic media). Pt and Ir are used as catalysts for the demanding electrocatalytic reactions in acidic media, with loads of Ir and Pt of about 0.7 kg/MW and 0.3 kg/MW respectively [8]. In addition, Pt is also used as a coating for some of the Ti constituents described above (mainly PTLs). One of the main advantages of PEMEC technology is the use of fewer balance of system components, as no electrolyte tanks, electrolyte circulation, or gas scrubbing are needed; gas-water separation is, however, still required on the water circulation loop. The use of stainless steel for electrolyser system tubing is also still necessary, as is the case with AEC technologies.

Solid oxide electrolysis

Solid oxide electrolysers (SOECs) are characterised by their ability to operate at high temperatures (i.e., 550–850°C), making them the most efficient technology of the three electrolysis technologies. Additionally, they are made of abundant materials (i.e., ceramic oxides). Table 4 shows the main characteristics of this technology. Compared to PEMEC and AEC, SOECs use much smaller stacks due to the current difficulties with scaling up high-quality and reliable ceramic technology. However, these electrolyser systems can already achieve the MW scale, allowing for their commercial deployment and paving the way for continuous development. The main advantage of SOECs over the other electrolysis technologies is their much higher electrical efficiency. They operate at or near the thermoneutral voltage for steam electrolysis (1.29 V), at which essentially all the electrical input is converted to hydrogen; the corresponding stack efficiency is 90% on an LHV basis when the energy required for steam raising is excluded, as given in Table 2. The average electricity consumption for SOEC stacks, when fed with steam at 150°C, is 37 kWh/kg H2. The stack lifetime is the shortest of all the technologies: 25,000 h.

SOEC materials include zirconium (Zr), iron (Fe), manganese (Mn) and stainless steels. There are also other materials, such as cerium (Ce), lanthanum (La) or yttrium (Y), that are

less abundant but still cost-effective and readily available [6]. Special mention must be made of nickel (Ni) and cobalt (Co), as both materials are used quite extensively in SOEC constituents. However, the current use of Ni and Co is only 200 kg/MW and 25 kg/MW, respectively; for Ni this is a quarter of the figure cited above for AEC fitted with solid nickel electrodes [8]. The use of high temperatures is another material concern, as more advanced stainless steels need to be used in both stack constituents and hot boxes when the operating temperature is at the high end of the range, above about 700°C, as stacks will be connected in series and operated at the desired temperature. However, recent developments show a trend towards operating temperatures below this level, where cheaper stainless steels can be used.

Balance of System

The specific components required to connect different electrolyser units vary depending on the technology used. Such components can encompass gas separators, gas scrubbers, gas purifiers, connectors, piping, and more. For instance, in alkaline electrolyser system, due to the liquid electrolyte, gas separators, scrubbers and purifiers are needed (the latter can be aggregated for several systems to profit from economies of scale and be included in the Balance of Plant instead). These components have considerably large footprints and are normally shared for few stacks. The electrolyser system is composed of both the piping and its corresponding electrical connections. On the other hand, SOEC and PEMEC do not have the need of some of these components (i.e., purifiers) as the output of their stacks is of higher purity (both hydrogen and oxygen).

MV/LV transformers

Normally OEMs also offer the MV/LV transformers required to obtain low voltage required to feed the rectifiers from medium voltage. As observed in industry for large-scale projects, the high voltage to medium voltage is frequently within the scope of the EPC contractor. Similar to rectifiers, the MV/LV transformers can vary in size but it is frequent that in large H2 plants using the chemical plant configuration several rectifiers can be paired with a single unit common to the entire electrolysis system. In H2 plants using the prefabricated configuration, each stack is typically combined with single MV/LV transformer and a single rectifier.

Rectifiers

Typically, one or several electrolysis stacks are paired with a set of rectifiers, varying in size and type (for instance, thyristor based versus IGBT based). These components will convert the AC current to DC current required for the electrolysis process. Different options exist, and specific designs could be done in function of the different characteristics of the technology and the size of the hydrogen plant.

Separators

In alkaline electrolysis, separators extract pure hydrogen isolating it for downstream treatment. Inside an electrolysis system, it is common to use a single separation unit to collect the hydrogen from the stacks in it.

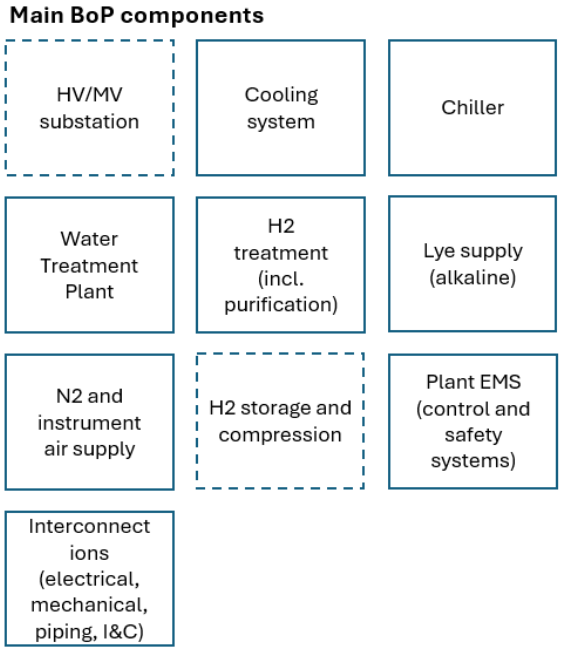
Control System

The electrolyser system is a highly automated assembly governed by a control system that manages the entire operation, ensuring seamless functioning and efficiency. This could include gas sensors, safety shutdowns, Programmable Logic Controllers (PLC) panels, and Supervisory Control and Data Acquisition (SCADA).

Balance of Plant

Beyond the electrolyser system, a hydrogen plant needs several components to deliver the desired hydrogen to the final offtaker. The most relevant are shown in Figure 7 and described in the subsections below.

Figure 7: General components of the balance of the plant in a H_2 plant. HV/MV substation, compressors and storage system are shown with dotted lines as they are not required in every hydrogen plant, and their cost is highly sensitive to project specifics.

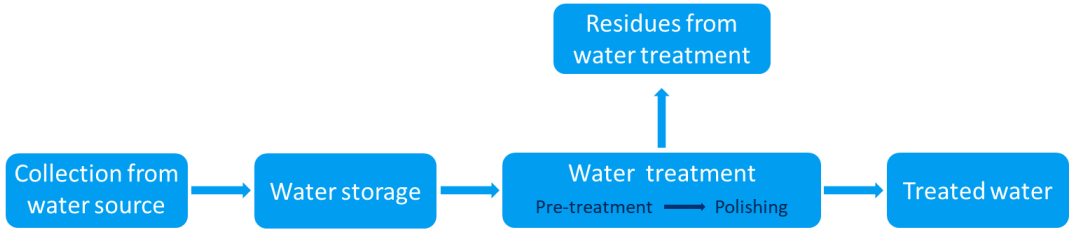


Water Treatment Plant

High quality water is required during the water electrolysis process, as impurities of water will lead to undesired products in the electrolysis cells and can also irreversibly contaminate the cell. Generally, large volumes of water will be needed for hydrogen production and therefore, water treatment plants will be coupled to hydrogen production.

In general, water sources can be split into several types such as surface water, groundwater, city or drinking water, seawater or effluent treated wastewater. Each type will need a slightly different approach as they have varying contents of minerals, sediments, and other contaminants. Furthermore, depending on the amount of treatment necessary, different amounts of water extracted will be necessary. For 1 m^3 of ultrapure water, it will be required to source 1.4 m^3 of groundwater, 1.5 m^3 of wastewater or surface water, and 3.3 m^3 of seawater. 1 kg of H_2 requires approximately 9 kg of ultrapure water for the electrolysis process [9].

Figure 8: General flow diagram of water.



Depending on the type of electrolysis and the source of water, there will be different requirements for water treatments (Figure 8). However, in each case there will be a pre-treatment step followed by polishing within the water treatment. The source of water determines the pre-treatment, and the polishing step is determined by the electrolyser technology. The steps may include filtration, aeration, UV, desalination, followed by softening, demineralisation, degassing and polishing EDI.

Compression

Hydrogen leaves the electrolyser at ambient or moderate pressure and must often be compressed for storage, transport, or delivery at the pressure required by the offtaker. The required compression duty is set by the inlet pressure, the delivery pressure, and the flow rate, and it varies widely between applications.

Pressurised alkaline and PEM systems typically deliver hydrogen at 15 to 35 bar, which is sufficient for many Power-to-X applications without further compression, whereas storage, pipeline injection, and some downstream processes require higher pressures.

Several compressor types are used, principally reciprocating compressors, including metal piston, diaphragm, and ionic liquid piston designs, which are well suited to high-purity hydrogen because they avoid oil contamination. Less conventional options include hydride compressors, which adsorb hydrogen at ambient conditions and release it at higher pressure on heating, and electrochemical compressors, which drive hydrogen across a proton exchange membrane under an applied voltage. Hydrogen compressors are commonly arranged in several stages with interstage cooling, and the associated energy use comprises both shaft power and the cooling load [10].

For a more extensive overview of hydrogen compression, see the technology catalogue “Technology data for energy transport” section “Introduction to transport of gases and liquids” [10].

Nitrogen supply

Nitrogen is used in green hydrogen production for safety and for component preservation. It purges and inerts the system to reduce the risks associated with hydrogen's flammability, removing residual gases before start-up and establishing an inert atmosphere during maintenance. It is also used to blanket stored hydrogen and to protect stack components during shutdown.

Instrument air supply

In some hydrogen plants, an instrument air system provides clean, dry, compressed air for pneumatic valves, actuators, and safety devices across the plant. It typically comprises air compressors, filtration units, dryers, and distribution manifolds meeting defined purity and dew-point specifications. The need for instrument air is a feature of the chemical-plant configuration, where pneumatic actuation is standard process-industry practice. In the prefabricated configuration, valves and actuators are typically electrically operated, and no plant-wide instrument air system is required.

Vent system

A hydrogen plant requires a means of safely discharging hydrogen during start-up, shutdown, depressurisation, and purging, in process upsets, and on activation of pressure relief devices. In electrolysis plants this duty is normally met by a vent system rather than a flare. Hydrogen is far lighter than air and disperses rapidly, so it can be released through

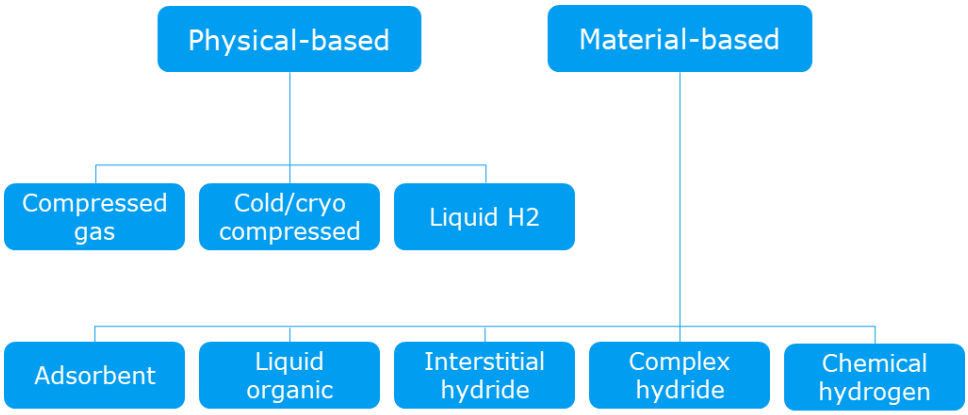
elevated vent stacks designed for safe dispersion, and the small and intermittent release volumes of an electrolysis plant do not justify a flare with its continuous pilot and support systems. Flares are generally found only where hydrogen is processed together with hydrocarbon streams, as in refineries, where safe dispersion of a vented gas cannot be assumed.

The location of the vent system follows the plant architecture. In the chemical-plant configuration the relief and purge streams are collected in a vent header and discharged through one or more common vent stacks at plant level. In the prefabricated configuration each unit carries its own vents as part of the integrated product, and no plant-level vent header or stack is required. In either case the vent points must be located and sized so that the discharged hydrogen disperses without forming a flammable atmosphere at equipment or personnel locations, which is addressed in the hazardous area classification of the plant.

Storage and transport

Before the usage of hydrogen, it may need to be stored or transported. This needs to be done in a manner that is both safe and efficient to ensure its viability as a fuel source. Some methods include utilising gas compression, cryogenic liquefaction, or chemical carrier methods. Chemical carriers may include liquid organic hydrogen carriers (LOHC), ammonia (NH_3), or methanol (CH_3OH).

Figure 9: Different options to store hydrogen.



Ann.: Developed from [11].

Physical methods are the most common in industrial use. Hydrogen can be stored as a compressed gas, as a cryogenic liquid, or, at large scale and for longer periods, in salt caverns. Compression reduces the gas volume but consumes energy, except where the hydrogen is already delivered at pressure from the electrolyser. Cryogenic storage holds hydrogen as a liquid at very low temperature and consumes more energy still, for liquefaction and to maintain the low temperature. Transport follows the same divide, with high-pressure pipelines and tube trailers for gaseous hydrogen and cryogenic ships for long-distance movement of liquid hydrogen.

Material-based methods, in which hydrogen is held in metal hydrides, chemical hydrides, or sorbent materials, are still largely under development. They can store hydrogen at near-ambient conditions and offer higher energy density, but face challenges of weight, cost, and the rate at which hydrogen can be taken up and released. Chemical carriers such as liquid organic hydrogen carriers (LOHC), ammonia, and methanol are a related route, storing hydrogen in a more readily handled liquid form.

Storage requirements depend strongly on the project, the location, and the offtaker, and storage is therefore omitted from the cost estimates in this chapter.

For a more extensive overview of hydrogen storage, see the DEA's technology catalogues "Technology data for energy transport" section "Introduction to transport of gases and liquids" [10] and "Technology Data – Energy storage" section "Hydrogen Storage" [11].

Cooling system

Cooling systems are essential in green hydrogen production, particularly during AEC and PEMEC electrolysis and compression stages. They prevent damage and decreased efficiency by managing excessive heat generated by electrolysis. Efficient cooling is crucial for optimal operation and longevity of the electrolyser system, and will involve coolants, heat exchangers, and advanced techniques like liquid cooling. These cooling systems maintain components at optimal temperatures, contributing to stability, durability, and overall process efficiency.

Chiller system

A chiller provides coolant below ambient temperature, by means of a refrigeration circuit, where the cooling duty cannot be met by ambient-temperature cooling alone. Whether one is needed depends on the system design and the site conditions rather than the electrolyser technology in isolation. In practice, electrolyser stacks – particularly AEC and PEMEC systems – operate within relatively narrow temperature ranges to maintain efficiency, limit degradation, and protect internal materials. Where the stack and balance-of-plant heat can be rejected to ambient air or water at the available temperatures, a chiller is not required, and several pressurised alkaline systems operate without one. However, a chiller becomes necessary where tighter thermal control is required than ambient heat sinks can provide, for example to ensure stable stack performance under varying load conditions, operate in hot climates, or condition specific process streams. Where fitted, the chiller forms part of the plant's thermal management strategy within the balance of plant. It consists of a closed-loop refrigeration unit with heat exchangers and circulation pumps, sized to the heat load and the operating regime. While it adds both capital cost and an additional electrical load, it can enable improved temperature control, operational flexibility, and overall plant reliability.

Electrolyte supply (alkaline)

Alkaline electrolysers circulate a liquid electrolyte, a potassium hydroxide (KOH) or 'lye' solution, between the stack and the electrolyte management system. This system prepares, stores, circulates, and conditions the electrolyte, and comprises storage tanks, circulation pumps, filtration, and concentration control. Its functions are to maintain the electrolyte at the correct concentration and purity, to remove particulates and impurities that would otherwise shorten stack life, and to carry heat away from the stack as part of the cooling duty. Because KOH is corrosive, the system is built from compatible materials and includes containment and monitoring. The electrolyte system is specific to alkaline technology; PEM and solid oxide electrolysers use solid electrolytes and have no equivalent.

The location of the electrolyte system depends on the plant architecture. In the chemical-plant configuration it is a plant-level or train-level installation shared across several stacks. In the prefabricated configuration the electrolyte loop can be integrated within each electrolyser unit, so that there are no separate plant-level electrolyte system and the electrolyte handling is contained entirely within the delivered product.

H2 purification

Electrolysers produce hydrogen at high purity, but additional purification is often required to meet the offtaker's specifications for moisture, residual oxygen, or particulates. The purification step may use catalytic deoxygenation to remove oxygen, drying to remove water and where higher purity is needed, pressure swing adsorption (PSA), or membrane separation. The duty depends on both the electrolyser technology and the intended use. Hydrogen from alkaline electrolysers generally needs deoxygenation and drying, whereas PEM and solid oxide electrolysers deliver hydrogen at higher purity and require less treatment. An offtaker such as an ammonia synthesis loop tolerates lower purity than one requiring fuel-cell-grade hydrogen. Where purification is shared across several stacks it may be located at plant level within the balance of plant, while in integrated unit designs the conditioning to delivery purity is carried out within the unit.

Plant HV/MV substation

In hydrogen plants, the required electrical power connection, often referred to as the voltage level, depends on the hydrogen plant's size and specific operational requirements. In addition, the available power will be very country dependent as every system operator will have different power options. For example, in Denmark, a 10 MW hydrogen plant would typically have a 10 kV power connection, while a 1 GW plant would have a 400 kV power connection.

While small projects (<10 MW) would have most of the power electronic components within the scope of the electrolysis system design, as the projects become larger (i.e., >10 MW) specific electrical designs will be performed to reduce costs and due to the need of more components as further transformers and components such as switchgear will be required.

Switchgear in a hydrogen electrolysis plant is essential for safe and efficient operations and is an integral part of the HV/MV connection. It controls and protects the power distribution system, ensuring reliable electricity flow. It includes power supply switchgear, main switchgear for distribution, and control switchgear for electrolysis cell operation. Protection switchgear safeguards against faults and abnormalities. Communication and monitoring capabilities enable real-time data acquisition and remote control. Safety measures are crucial due to hydrogen's flammability. Overall, switchgear ensures reliable power distribution, protection, and efficient plant operation.

Power requirements such as large step-down transformers and further switchgears may be required depending on the project specific or location specific infrastructure.

Note: In the cost figures included in the financial data for this chapter, the HV/MV transformers are not included.

Plant Energy Management System (EMS)

The plant control system monitors and manages the production process, from water treatment through electrolysis to hydrogen delivery, and regulates parameters such as current, temperature, and pressure. It handles data acquisition from sensors across the plant, supporting monitoring, fault diagnosis, and maintenance planning, and it interfaces with the safety systems, including gas detection and emergency shutdown. In larger plants the EMS also coordinates the operation of multiple electrolysis trains and modules and manages the plant's response to the electricity market and to grid signals.

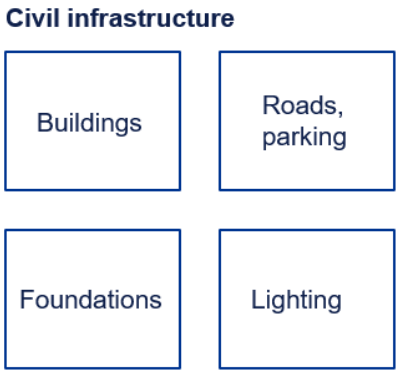
Interconnection systems

The plant's subsystems are linked by four kinds of interconnection. First: electrical interconnections comprising the cabling, switchgear, grounding, and protection that carry power from the grid or renewable source to the electrolysers and auxiliaries. Second, piping interconnections, carrying hydrogen, oxygen, nitrogen, cooling water, treated water, and electrolyte between subsystems designed to the relevant pressure, purity, and material-requirements. Third, mechanical interconnections, the supports, structures, and skids, that physically integrate equipment and secure proper alignment, vibration control, and accessibility for maintenance. Fourth, instrumentation and control interconnections linking sensors, transmitters, and actuators, to the supervisory systems that ensure accurate monitoring of process variables and enable automated, safe operation of the entire hydrogen plant.

Plant civil infrastructure

Civil infrastructure components relate to the physical foundation of the plant, such as construction of buildings or access roads, and potentially added infrastructure like train tracks, among others. These foundations can vary by size depending on soil characteristics of the location, including soil improvements such piling. Average foundation sizes, rough grading, roads, site paving and a water treatment building are included in the estimates in this chapter, but no other buildings (Figure 8).

Figure 10: General components of the civil infrastructure category of a hydrogen project.



Buildings

A hydrogen plant may include several buildings, but most of the equipment does not require one.

In the chemical-plant configuration a multi-100 MW plant is divided into electrolysis buildings, each housing a train of several modules, typically 100 MW per building, together with other buildings such as a control room, offices, and a spare-parts warehouse. Buildings that are provided must meet operational, safety, and environmental requirements as well as the applicable codes for hydrogen handling, including the International Building Code, NFPA 2 Hydrogen Technologies, and the International Fuel Gas Code, with ventilation and gas detection to prevent hydrogen accumulation and fire detection and suppression where required.

In the prefabricated configuration, part of the plant can be installed outdoors, depending on the climatic protection, so the electrolysis system and part of the balance of plant do not always require a building. However, some parts of the EPC supply, e.g. compressors are usually housed indoors.

Roads, access, parking

The plant requires road access for staff, deliveries, and hydrogen transport, designed for the heaviest vehicles expected and including emergency-services access, evacuation routes, signage, and parking. Access for maintenance and equipment replacement should be provided for.

Foundations

Foundations support the electrolyser systems, compressors, storage, and cooling equipment, and must be designed for the equipment loads, the local ground conditions, and any machinery vibration. Drainage should be provided to prevent water accumulation beneath the structures.

Lights

Lighting must provide adequate visibility for operation and maintenance and include emergency lighting for evacuation. Energy-efficient fixtures are preferred, and explosion-protected fittings are required in areas where hydrogen may be present.

Inputs

For all three electrolysis technologies, the inputs are mainly water and electricity. For SOEC, PEMEC and AEC, high- purity water is needed. The water quality is normally determined by the conductivity of the water but also by the content of other impurities. Water categories are normally described as Type I (ultrapure water), Type II and Type III (drinking water). While SOEC and AEC can accept Type II water with conductivities around 1 µS/cm, PEMEC technology needs Type I and conductivities below 0.1 µS/cm. In addition, SOEC use steam rather than water and as it operates at the thermoneutral point is thus a consumer of heat. If the hydrogen produced was to be used for synthetic fuel, then the waste heat of these synthesis processes (e.g., Fischer-Tropsch synthesis, Haber-Bosch process) could be used to produce steam for further SOEC electrolysis and increase SOEC electrolysis efficiency further [12].

As mentioned in previous sections, the required electricity input for stacks is DC power (high current levels in low voltage). This is the reason for why MV/LV transformers and subsequently rectifiers are needed. For balance of system and balance of plant components the supply can be taken from the LV bus in the transformers to deliver three phase 400 V electricity covering various subsystems.

Output

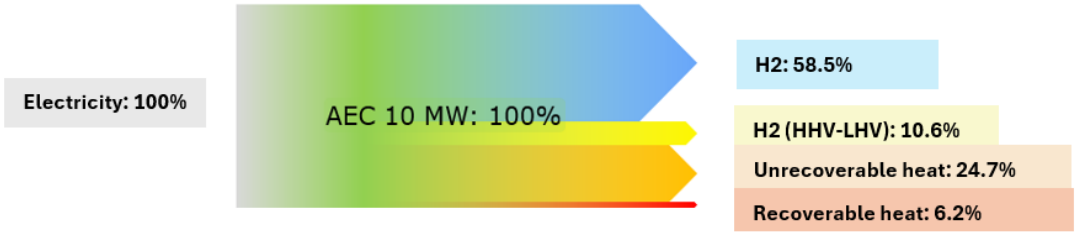
The outputs of all three technologies are hydrogen and oxygen. Alkaline and PEM electrolysers also produce excess heat, whereas solid oxide electrolysers, operating at or slightly below the thermoneutral point, absorbs heat rather than rejecting it [13]. The oxygen is a by-product with potential uses in industries such as steel, glass, pulp and paper, medical supply, and aquaculture, but few projects have found a business case for capturing it, since the cost of purifying, storing, and transporting the oxygen cannot generally compete with oxygen from large-scale air separation [14]. The excess heat from AEC and PEM electrolysers can be used for district heating; the systems operate at 60 to 80 °C, which suits low-temperature district heating networks, typically supplying at 40 to 50 °C with a 25 °C return.

Energy balance

The energy balances for the three electrolyser technologies are shown in Figures 9, 10, and 11. They are given at plant level, including the balance of plant, and reported on a lower heating value (LHV) basis. For alkaline and PEM electrolysers the input is entirely electricity. For solid oxide electrolysers, electricity supplies about 80% of the energy input and heat the remainder, since the water must be raised to steam and the stack operates

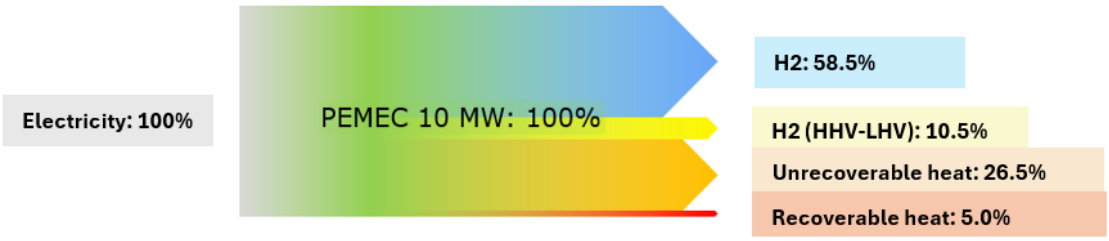
above 600 °C. Once water or steam reaches the cells, hydrogen and oxygen form and heat is released; only the fraction of that heat available at a useful temperature can be recovered, for example for district heating.

Figure 11. The energy inputs and outputs of an AEC (10 MW hydrogen plants).



Source: Data is derived from DNV references.

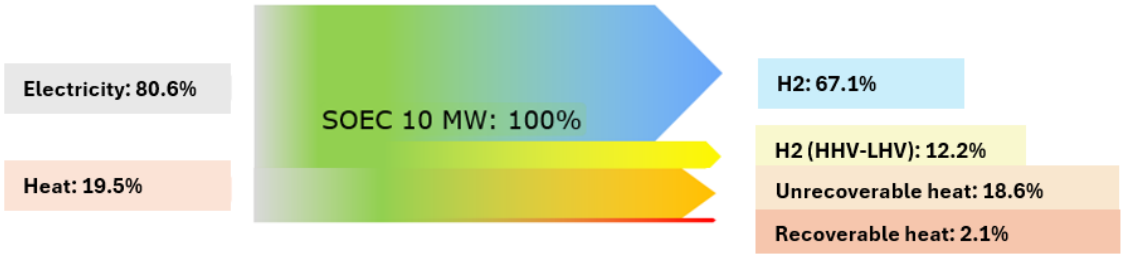
Figure 12: The energy inputs and outputs of a PEMEC (10 MW hydrogen plants).



Source: Data is derived from DNV references.

Once water or steam is supplied to the electrolysis cells, formation of H_2 and O_2 takes place along with heat dissipation. Since the heat needs to be at a certain temperature to be fed into the district heating grid, only a percentage of the heat generated can be directly utilised. An important aspect related to the analysis performed herewith does not consider the latent heat of vaporization of steam in the product and normally referred as the Low Heating Value (LHV) and High Heating Value (HHV) of hydrogen. This is performed in order to get an accurate analysis of the usable energy produced by electrolysis.

Figure 13: The energy inputs and outputs of an SOEC (10 MW hydrogen plants).



Source: Data is derived from DNV references.

Typical capacities

A plant is built up from multiple stacks. The maximum commercial stack size today is given in Table 2 and the data sheets set out how it is expected to grow. AEC and PEMEC

electrolyser plants are planned at the multi-100 MW and multi-GW scale while SOEC projects are in the multi-10 MW range.

Flexibility

Most electrolysers can be operated dynamically. The practical limits are the minimum partial load set by hydrogen crossover, and the response times of the external components. The cold-start time, hot-start time, ramp rates, and load range for the three technologies are given in Table 5.

A cold start is start-up from ambient temperature after a prolonged shutdown, sometimes with the system depressurised for non-atmospheric stacks. A hot start, or warm start, is start-up from a heated stand-by condition, in which the system is held near its operating temperature and, for pressurised systems, pressurised. The power response is the time taken to move between minimum and nominal load while producing hydrogen.

An electrolyser stack is more efficient at lower current density, and therefore at lower load, down to the point where hydrogen crossover sets the safe minimum. Working against this, shunt losses and the balance-of-system and balance-of-plant loads do not fall in proportion to output, so they represent a larger share of consumption at low load. The two effects typically combine to place the highest system efficiency at around 50% load.

Table 5. Cold and warm start-up time, ramp rates and load range for 10 MW AEC, PEMEC and SOEC technologies (Status 2025).

	AEC	PEMEC	SOEC
Cold start-up time (from 0 to 100%) [minutes]	< 60	15	480
Hot start-up time (from 0 to 100%) [minutes]	20 (atmospheric) 5 – 15 (pressurized)	< 8	60
Ramp rate (up) [%/sec]	5 (atmospheric) 15 – 25 (pressurized)	10	0.8
Ramp rate (down) [%/sec]	10 (atmospheric) 30 – 50 (pressurized)	20	0.2
Load range per electrolyser stack (%)	30 – 102	15 – 105	80 – 100
Degradation rate (%/1000 h)	0.13	0.17	0.75

Source: Data in the table is derived from DNV references and sources listed in the text.

Degradation

Flexible operation shortens stack lifetime. Frequent load changes and start-stop cycles accelerate degradation, as do operating temperature and water quality. A stack is generally

considered to have reached end of life once its performance has fallen by about 10%, at which point refurbishment or replacement is required. This matters most for systems coupled to variable renewable generation, where the fluctuating supply forces dynamic operation and repeated cycling. The main mechanisms are accelerated catalyst degradation, reverse currents at start-up and shutdown, and increased thermal and mechanical stress in the stack, though their relative importance differs between technologies and there is little firm comparative evidence across them. Some of electrolysers' degradation mechanisms can be reversible, while other degradation phenomena are irreversible.

Partial load operation

Flexible operation can improve electrolyser efficiency at partial load. Stacks are generally more efficient at lower current densities because activation overpotential, ohmic losses and mass transport limitations increase at higher loads. However, operation cannot be reduced indefinitely, as hydrogen crossover eventually limits safe operation.

At plant level, BoS and BoP electrical loads do not scale linearly with hydrogen production, meaning their relative consumption is higher at low loads. Considering both effects, the highest system efficiency is typically reached around 50% load. Oversizing plant capacity relative to upstream substation power could therefore provide efficiency gains, although these are partly offset by higher CAPEX and are typically modest, around 5% compared with nominal operation.

Flexibility (AEC)

Alkaline electrolysers are robust and long-lived under steady operation but less tolerant of frequent load changes. Dynamic operation can cause gas bubble entrapment, deterioration of the nickel-based electrodes, and increased hydrogen crossover through the diaphragm, while rapid shutdowns and frequent cold starts add the risk of electrolyte crystallisation, seal degradation, and electrode fatigue. The operating range and dynamics depend on whether the system is atmospheric or pressurised. Both can turn down to 30% of rated load, but they differ markedly in ramp rate.

Pressurised alkaline systems are more responsive and deliver hydrogen at higher pressure, in the range of 15 to 35 bar, and have already demonstrated response times fast enough for grid services. For these systems the ramp rate is now limited mainly by the balance of plant rather than the stack. The hot-start time is currently 5 to 15 minutes, and ramp rates in production are typically 5% per second up and 10% per second down, with some OEMs claiming 15 to 25% per second for ramp up and 30 to 50% per second for ramp down. Atmospheric alkaline systems operate at 0.1 to 4 bar and take around 20 minutes to ramp from hot stand-by to nominal load, where pressurised systems take 5 to 15 minutes. It is uncertain whether atmospheric alkaline will see comparable improvement in ramp rate, or whether the emphasis will shift to pressurised systems where fast response is needed.

Flexibility (PEMEC)

PEM electrolysers offer fast, flexible operation and are often preferred for coupling to renewable power where no stabilising grid connection is available. They can turn down to 5 to 15% of rated load and reach nominal load from hot stand-by within seconds, at ramp rates above 10% per second up, and 20% per second down, which is already sufficient for grid services. Some OEMs permit short-term overload to 120% of nominal capacity by raising the current density, at the cost of increased degradation and the risk of overheating the balance-of-plant components, particularly the power electronics, unless these are oversized. Repeated start-up and shutdown can still induce reverse currents that degrade the membrane-electrode assembly, and dynamic operation increases hydrogen crossover through the membrane, an effect that many OEMs limit with recombination catalysts at the

anode. Documented improvement targets relate mainly to cold-start time, which some manufacturers already place below 5 minutes.

Flexibility (SOEC)

Solid oxide electrolysers are suited primarily to steady baseload operation. The ceramic electrolyte is fragile, and the high operating temperature gives the stack large thermal inertia, so output cannot be ramped quickly. Frequent cycling causes thermomechanical stress and accelerated degradation, and some OEMs recommend not reducing below 80% of nominal load because of the difficulty of thermal management, which limits the use of partial load. Given the lower maturity of the technology relative to alkaline and PEM electrolysers, there is scope for improvement in the longer term.

Overall, the choice of technology should be matched to the intended operating regime, including the expected supply profile and the degree of load variability. Realised stack life in practice is governed more by the degradation associated with the operating regime than by the nominal design figures.

Ancillary services participation

Electrolysers are grid-connected loads and, being controllable and fast-responding, can provide ancillary services by adjusting their power draw in real time, increasing consumption when system frequency rises and reducing it when frequency falls. Ancillary services are not fully standardised across Europe, but ENTSO-E is harmonising the main categories. The EU Electricity Balancing Guideline defines Frequency Containment Reserve (FCR, activated within seconds), Automatic Frequency Restoration Reserve (aFRR, within minutes), Manual Frequency Restoration Reserve (mFRR), and Replacement Reserve (RR) for slower balancing. Some operators also pay industrial loads for interruptibility; Spain is one example, where large consumers are paid to be available to reduce consumption on request.

Denmark illustrates how flexible loads fit into these markets. The country spans two price zones in different synchronous areas, DK1 in continental Europe and DK2 in the Nordic area, so the available products differ between them. Denmark procures a full set of frequency reserves, including FCR, aFRR, and mFRR in DK1, and additional services such as FCR-N, FCR-D, and Fast Frequency Reserve in DK2.

Space requirement

The stack footprints for the three technologies are given in Table 3. PEM electrolysers are the most compact at 7.5 m²/MW, against 10 m²/MW for alkaline and 20 m²/MW for solid oxide electrolysers. These are stack figures only. The overall plant footprint is larger and the differences between technologies narrow once the balance of plant is included, and they narrow further where a compressor is added for high-pressure delivery or where the power electronics, rectifiers, transformers, and switchgear, take significant space.

Environment

The only process outputs of all three technologies are hydrogen, oxygen, and heat. The environmental considerations relate mainly to the materials in the stacks. AEC depend on nickel, whose use in batteries and other renewable technologies is making it a more contested resource.

PEMEC use a fluoropolymer (PFAS) membrane, which must be disposed of or recycled at end of life, and platinum-group catalysts, platinum at the hydrogen electrode and iridium and ruthenium at the oxygen electrode, which are scarce and a potential constraint on large-scale deployment [18]. The PFAS membrane in PEMEC warrants particular attention, as the European Commission is considering a broad restriction on PFAS substances under REACH. Were such a restriction adopted, it could affect both the cost trajectory and the

long-term availability of PEMC, and this is not currently reflected in the cost projections in this chapter.

SOEC use cerium, lanthanum, and yttrium, which are less abundant but available as oxides and relatively low in cost.

Advantages/disadvantages

In this section, a summary of advantages and disadvantages of AEC, PEMEC and SOEC are displayed in

Table 6. The data is collected from literature [15], [16].

Table 6: Summary of advantages and disadvantages of electrolysis technologies.

	Advantages	Disadvantages
AEC	1. The technology is mature and scalable. 2. Low operating temperature. 3. Quick start up (pressurised) for response in grid services 4. Suitable for use as a flexible technology. 5. Long stack lifetime 6. MW scale electrolyser systems are already being deployed. 7. Low presence of critical raw materials 8. Lower CAPEX per installed power	1. Less flexibility (for atmospheric systems) 2. The use of highly caustic electrolyte. 3. Corrosivity of KOH. 4. High membrane resistance. 5. Low maximum operational current density, nominally operated around 0.6-1 A/cm^2 as average [17]. 6. Higher footprint 7. Less hydrogen purity than PEMEC 8. For pressurized systems, lower levels of pressure are reached in comparison with PEMEC
PEMEC	1. Low operating temperature, low noise, high power density. 2. Quick response time. 3. Pressurised hydrogen can be produced. 4. Current densities >2.0 A/cm^2 can be used for operational systems leading to compact system sizes. 5. MW scale electrolyser systems are already being deployed. 6. Smaller footprint than AEC.	1. Sensitive to impurities, with a prerequisite of very pure water (Type I) as input. 2. Lifetime of the commercially available systems is lower than AEC 3. Presence of scarce and expensive critical raw materials, like Platinum or Iridium used as catalysts in electrode layers 4. Higher CAPEX per installed power
SOEC	1. High efficiency if coupled to a heat source. 2. Low presence of critical raw materials 3. Possibility for synthesis gas from co-electrolysis of steam and CO_2. 4. Potential to be used in reverse mode as fuel cell.	1. Still in demonstration phase for large scale applications for hydrogen production 2. Low scalability potential for large projects: SOEC's stacks are about 10 times smaller in H_2 output than PEMEC and AEC. 3. Instability of electrodes and ceramic electrolyte 4. Heat management issues 5. Stack components are susceptible to corrosion.

Advantages	Disadvantages
	6. Commercially available lifetime system is shorter than PEMEC and even shorter to AEC. 7. SOECs can be operated only at current densities up to 0.5 A/cm^2. 8. Higher CAPEX than AEC or PEM systems 9. Higher minimum partial load and less dynamic response than AEC or PEMEC

Research and development perspectives

For AEC, the main objectives are higher efficiency and durability while retaining low-cost materials, pursued through both catalyst and stack design. Examples include Hysata, which has reported a capillary-fed cell design with efficiency as high as 95 per cent [19]. and Hydrogen Pro, working on catalysts that avoid expensive metals [20]. In parallel, OEMs are increasing stack size and modularising the balance of system and balance of plant to produce more compact and cost-competitive designs for large projects.

For PEMEC, stack cost is the main barrier to large-scale deployment, and research addresses the cost of catalysts and bipolar plates, degradation, and the substitution of scarce materials. TNO has reported a reduction in iridium loading by a factor of about 200, pointing to substantially cheaper stacks [21].

For SOEC, the focus is scalability and durability. ISPT has reported that larger cells, with greater area and higher current density, could yield hot modules of around 1 MW and so enable GW-scale systems, and lowering the operating temperature would allow cheaper materials [22]. In addition, lowing the temperature of operation of SOECs will allow the use of even cheaper materials to make SOECs stacks more cost-effective.

Examples of market standard technologies

This section discusses recent full-scale commercial projects. As it can be seen, several are focusing on what is considered hard-to-abate industries. Fertiliser, steel and chemical industries are the most typical examples of hard-to-abate industries. These industries require large quantities of hydrogen continuously (> 10 000 tonnes per year) which requires the construction of a hydrogen plant with an electrolyser system as large as 100 MW. This leads to profiting from economies of scale (lower cost per unit of installed power of electrolysis). In addition, pressure and purity will depend on the final use. Most of these industrial demands would require relatively low pressures (< 50 bar) and not very demanding hydrogen purities (99.5%), which opens to the use of the three technologies. In addition, in case that the operational pattern allows it, SOEC could profit of the excess heat released in these industrial processes, which would make the hydrogen production very efficient.

Table 7 summarises some examples of different electrolysis plants deployed either commercially or in demonstration phase in recent years. The electrolyser systems at different locations are deployed for various applications and their specifications are also described in the table.

Table 7: Examples of operational hydrogen plant projects in recent years across locations, sizes and technologies.

Project	Location	Type	Size (MW)	H_2 prod. (kt/y)	COD year	Off-taker Sector
Iberdrola – Puertollano I	Spain	PEM	20	3	2022	Ammonia
HySynergy, phase 1	Denmark	ALK	20	3.4	2025	H2
Refhyne	Germany	PEM	10	1.5	2021	H2
Green fertilizer project Porsgrunn-Heroya, phase 1 - SKREI	Norway	PEM	24	3.6	2024	Ammonia
H2 Pilotanlage Lingen, phase 1	Germany	ALK	10	1.7	2024	H2
UpHy, phase 1	Austria	PEM	10	1.5	2025	H2
REDDAP - Renewable Distributed & Dynamic Ammonia Plant	Denmark	ALK	10	1.7	2024	Ammonia
P2X Harjavalta project	Finland	ALK	20	3.4	2025	CH4
Hy4Chem - BASF Ludwigshafen	Germany	PEM	54	8.1	2025	H2
Hofors rolling project	Sweden	ALK	17	2.9	2023	H2
Trailblazer - Siemens-Air Liquide Oberhausen, Phase 1	Germany	PEM	20	3	2024	H2
Port of Aabenraa, Kasso Power-to-X facility	Denmark	PEM	52.5	7.9	2025	MeOH
Szazhalombatta refinery	Hungary	PEM	10	1.5	2024	H2
Orlen Hydrogen Hub in Wloclawek, phase 1	Poland	Other electrolysis	15.1	3	2024	H2

Source: IEA and DNV

Quantitative description

Prediction of performance and costs

The data sheets present techno-economic data and projections towards 2050 for AEC, PEM, and SOEC at several sizes in the chemical-plant concept. Future CAPEX estimates, for 2030 and beyond, are subject to substantial uncertainty. Cost estimates for the pre-fabricated units are not available for iteration of the chapter, due to the novelty of the concept and the limited number of suppliers.

The total plant CAPEX for each technology is split into seven cost items in the data sheets (for definitions on what is included in each item, see Table 8):

1. Electrolyser Stacks
2. Balance of System (BoS)
3. Balance of the Plant (BoP)
4. Compression
5. Control System
6. Civil Infrastructure
7. Indirect costs such as EPC works and commercial margins applied to equipment and materials by suppliers and contractors.

Table 8: Electrolysis cost taxonomy used in this chapter.

Scope of Supply - Categories		Typical Scope of Supply for two configurations		Included in cost figures of this report	Cost category
Category	Sub-category	Chemical plant	Prefabricated		
Stacks	Stacks	OEM	OEM	TRUE	1. Electrolyser Stacks
Balance of System (BoS)	Separators	OEM	OEM	TRUE	2. Balance of System (BoS)
	Rectifiers	OEM	OEM	TRUE	
	MV/LV Transformers	OEM	OEM	TRUE	
	Module level instr. & control systems	OEM	OEM	TRUE	5. Control System
Balance of Plant (BoP)	Demineralized water package	EPC	EPC	TRUE	3. Balance of Plant (BoP)
	H_2 purification systems	EPC	EPC	TRUE	
	Cooling	EPC	OEM	TRUE	
	Chiller	EPC	EPC	TRUE	
	Energy Management System, including Distributed Control Systems and Safety Instrumented Systems	EPC	EPC	TRUE	5. Control System
Balance of Plant (BoP, continued)	Fire & gas detection systems (F&GDS)	EPC	EPC	TRUE	3. Balance of Plant (BoP)
	Lye system	EPC	OEM	TRUE	
	N_2/air supply	EPC	EPC	TRUE	
	Vents	EPC	OEM	TRUE	
	Interconnections	EPC	EPC	TRUE	

Scope of Supply - Categories		Typical Scope of Supply for two configurations		Included in cost figures of this report	Cost category
Category	Sub-category	Chemical plant	Prefabricated		
EPC works & common BoP	Common HV/MV substation	EPC	EPC	FALSE	N/A
	Hydrogen compression	EPC	EPC	TRUE	4. Compression
	Hydrogen storage	EPC	EPC	FALSE	N/A
	Civil works, roads, foundations, lights	EPC	EPC	TRUE	6. Civil Infrastructure
	Procurement	EPC	EPC	TRUE	7. Indirect costs
	Equipment installation	EPC	EPC	TRUE	
	Site acceptance test (SAT) & precommissioning	EPC	EPC	TRUE	
	Commissioning works	EPC	EPC	TRUE	
	Other services (engineering, Project Management, Health Safety & Environment (HSE)	EPC	EPC	TRUE	
Owner's Engineering	Permitting	DEV	DEV	FALSE	N/A
	PreFEED & FEED packages	DEV	DEV		
	Studies	DEV	DEV		
	Due diligence	DEV	DEV		
	PMT & PMC/OE home office	DEV	DEV		
	Other (communication, marketing, connection fees, licensing, certifications)	DEV	DEV		
	Contingency Margin	DEV	DEV		

Source: DNV

CAPEX scope and system boundaries

The CAPEX in the data sheets includes compressors but excludes hydrogen storage and the HV/MV transformers. It also excludes grid connection infrastructure, beyond the initial medium-voltage to low-voltage transformers and rectifiers required by the electrolysers which are typically supplied by the OEMs (balance of system). This is because HV/MV transformers, switchgear, and the wider grid connection are highly project-specific and location-specific depending on the connection capacity, the voltage level, the distance to the nearest existing substation, and the local grid code, and are treated separately in this chapter rather than included in the generalised figures. Including full grid connection within the scope can increase total plant CAPEX substantially relative to the figures shown, which should be borne in mind when comparing these estimates with other studies. All CAPEX values are expressed at a 2025 cost level, excluding inflation, and exclude owner's cost and contingency, in line with the catalogue methodology.

Hydrogen compression

Compressor cost is highly project-specific since it is driven by inlet pressure, the delivery pressure and the flow rate increase from 30 to 80 bar and should be treated as indicative, because both the pressure rise and the flow rate strongly affect the result. Table 7 gives indicative compression costs, normalised to electrolyser capacity, for three inlet pressures and three plant sizes. The ranges in the table reflect the different nominal flow rates of the three technologies at a given MW rating, not an uncertainty band. The figures show clearly that the lower the compression duty, the lower the cost, and that a system delivering hydrogen already at elevated pressure avoids the largest part of it: at 10 MW the compression costs fall from 418 to 500 €/kW for compression from 1 bar to 87 to 104 €/kW for compression from 30 bar.

Table 9. General compression costs depending on pressure increase requirements.

Scope	Electrolyser output presusre	Compressor pressure range	10 MW_{ELX}	100 MW_{ELX}	1000 MW_{ELX}	Units
Compression unit cost	1 bara	1 > 40 bara	418 - 500	167 - 200	69 - 87	€/kW_{ELX}
Compression unit cost	15 bara	15 > 40 bara	198 - 236	75 - 91	30 - 37	€/kW_{ELX}
Compression unit cost	30 bara	30 > 40 bara	87 - 104	35 - 41	14 - 16	€/kW_{ELX}

Power connection

Power-connection infrastructure, the switchgear and high-voltage to medium-voltage step-down transformers can be a major cost, and depend heavily on the project site. Where no suitable substation exists nearby, building or expanding one raises upfront CAPEX, or alternatively raises long-term OPEX through the tariffs charged by the utility that owns it. Projects supplied from dedicated renewable generation carry different cost implications again. Because these costs are so project-specific, the datasheet CAPEX excludes power-connection infrastructure beyond the initial rectifiers and transformers at the electrolysers, which are included.

Key assumptions for the base year data

The assumptions underpinning the cost and performance estimates presented for the base year (2025) are outlined below.

Costs

The 2025 cost data are drawn from DNV's own reports and advisory projects supplemented by scientific and industrial literature. Unlike DNV's Energy Transition Outlook, which does not differentiate cost by electrolyser technology, this study derives a separate average cost per technology.

For AEC the reference case is a 100 MW onshore plant, the most common size in projects at advanced development with costs for 10 MW and 1 GW then derived by applying scaling laws consistently across all direct cost components. The same approach is used for PEMEC, again from a 100 MW reference.

For SOEC, where commercial plants at 100 MW do not yet exist and published data covers roughly 1 to 10 MW, that range is used as the reference and scaled up to 100 MW.

Indirect costs are calculated as a percentage markup on direct costs; in the data sheets, in line with the catalogue methodology, only the EPC-related part of that markup is included, and it is assumed to be higher in early years and to fall as the industry matures.

Performance

Current density and stack capacity are held constant across plant sizes for a given year and, current density is assumed to rise over time across all technologies as a result of R&D advancements traded off against efficiency.

For AEC and PEM electrolysers, stack dimensions (cell area, number of cells) are assumed to stay close to today's values due to practical limitations to stack dimensions and weight relating to transportability. Therefore, the rising current density translates into higher stack capacity over time.

Stack footprint and weight per MW are likewise held constant with scale for a given year

and fall over time. The change in plant footprint per MW with increasing capacity, is driven by more efficient layouts; reduced duplication of equipment at larger scale, and centralisation of balance of plant items such as chillers and purifiers across several modules or trains also lowers cost at the 1 GW scale relative to the 100 MW scale.

Offshore systems are assumed to require about 60 per cent of the footprint of the equivalent onshore plant, due to offshore weight and space constraints being more stringent, driven by cost.

Data for the base year

CAPEX

The 2025 data presented is based on DNV's expertise and ETO data [23], supplemented by inputs from a range of additional sources from industrial and academic literature to establish baseline values for key parameters such as CAPEX breakdowns [24] [25] [26] [27] [28] [29] [30] [31] [32] [33] and stack cost modelling [34] [35]. Collectively, these sources were used to inform:

- Cost composition shares and overall electrolyser plant cost structures;
- Total electrolyser plant CAPEX, including balance-of-plant components;

- Bottom-up cost assessments based on manufacturer input;
- Material cost breakdowns for electrolyser stacks;
- Cost contributions of individual stack sub-components.

The 2025 CAPEX estimates for all three electrolyser technologies and the different plant sizes considered are presented in Table 10. These estimates represent the aggregate of the cost components listed in section "Prediction on performance and costs".Costs related to temporary hydrogen storage, HV/MW transformers, grid connection, owner's cost, and contingency allowances are not included in these CAPEX estimates.

Table 10. CAPEX estimates for 2025 for all three technologies and different plant size.

Technology	Capacity	2025 costs (€/kW)
AEC	10 MW	1,849
	100 MW	1,254
	1 GW	936
PEMEC	10 MW	2,477
	100 MW	1,765
	1 GW	1,324
SOEC	1 MW	2,959
	10 MW	2,024
PEMEC Offshore (decentralised)	10 MW	4,105
PEMEC Offshore (centralised)	100 MW	2,853

Source: Data in the table is derived from DNV references

OPEX

Annual fixed OPEX, expressed as a share of total investment cost, is assumed to be broadly comparable for AEC and PEMEC, with the primary differentiating factor being plant size. For a 100 MW installation, the fixed annual OPEX for both AEC and PEMEC is assumed to fall within the range of approximately 3–6 per cent of total investment cost.

For SOEC systems, higher fixed OPEX is assumed. In the case of a 100 MW SOEC plant, annual fixed OPEX is estimated to range between 3 and 9% of total investment cost, resulting in a higher average fixed OPEX relative to AEC and PEMEC.

The average fixed OPEX applied for each electrolyser technology and plant capacity are summarised in Table 11. It should be noted that for electrolysis systems, OPEX costs include preventative and corrective maintenance, operation and monitoring costs, and spare parts (excluding stack replacements). Electricity and water costs are generally not included in electrolysis fixed OPEX, and have therefore been excluded in the presented estimates.

Table 11. Average annual OPEX for 2025 for all three technologies and different plant size.

Technology	Capacity	2025 annual OPEX (% of specific investment/year)
AEC	10 MW	5.5
	100 MW	5
	1 GW	4.5
PEMEC	10 MW	5.5
	100 MW	5
	1 GW	4.5
SOEC	1 MW	7
	10 MW	6.5
PEMEC Offshore (decentralised)	10 MW	10
PEMEC Offshore (centralised)	100 MW	9.5

Source: Data in the table is derived from DNV references

Performance

For the electrolyser technologies technological performance data, several sources were used looking at current and projected future stack capacities, current densities and efficiencies [36] [37] [38] [39] (Carlos Bernuy-Lopez, Ramboll / HydrogenTechWorld, 2023) [41] [42].

Assumptions for projecting costs into future years

The assumptions applied for future years build on those defined for the base year, with several of the same sources underpinning the forward-looking data.

Costs

Cost projections beyond 2025 are based on learning rates derived from DNV's Energy Transition Outlook (ETO) 2025 data. In the ETO, these learning rates assume:

- gigawatt-scale global deployment of electrolysis and
- cross-technology learning effects, whereby cost reductions are driven by cumulative deployment across all electrolyser technologies rather than being treated independently.

In this chapter, however, it is assumed that learning rates differ by technology, resulting in technology-specific cost-reduction trajectories (see section below).

- AEC: The 100 MW onshore configuration is used as the reference case for the 2025 data, and subsequently applied as the basis for future cost projections in the 2040 and 2050 scenarios.

- PEMEC: The 100 MW onshore configuration is used as the reference case for the 2025 data, and subsequently applied as the basis for future cost projections in the 2040 and 2050 scenarios.
- SOEC: The 10 MW onshore configuration used as the reference case for the 2025 data, and subsequently applied as the basis for future cost projections in the 2040 and 2050 scenarios.
- EPC costs are assumed to decline over time, reflecting a reduction in perceived project risk as the industry matures and electrolyser deployment becomes more established.
- Standard deviations for the cost estimates for 2050 are based on a more limited set of data points than earlier years, resulting in a higher degree of uncertainty. Furthermore, there is an inherent uncertainty in the cumulative global GW deployment towards 2050, which is the main driving factor in the learning curve calculations. Deviations in the cumulative deployment will have a large influence on the achieved cost reductions.

Performance

AEC and PEMEC

- stacks are assumed to be close to practical design limits, and the data sheets therefore assume that manufacturers are unlikely to increase the physical size of stacks over time.
- future performance improvements are assumed to arise from increases in current density, leading to higher MW ratings per stack while maintaining constant physical stack dimensions. Linear scaling of MW rating is applied for current density increases.

SOEC:

- it is assumed that, even in long-term projections, SOEC stack sizes remain significantly smaller than those of AEC and PEMEC technologies.
- stack size is assumed to scale with current density, cell area, and the number of cells per stack, reflecting the lower maturity of the technology and greater scope for design evolution.

Learning curves and technological maturity

Technological Maturity Categories

As noted previously, projections of future costs and performance for the three electrolyser technologies take learning effects into account. In its catalogue, the technologies are classified according to four technology maturity categories: research and development phase (category 1), pioneer phase (category 2), commercial technologies with moderate deployment (category 3) and commercial technologies with large deployment (category 4). These categories can be directly compared with Technology Readiness Levels (TRLs), which classify technological maturity on a nine-level scale. The TRL framework, as defined in Horizon Europe guidance for TRL self-assessment [43]. Based on these definitions, AEC and PEMEC are considered to be at TRL 9, while SOEC are assessed at TRL 8 [44].

Table 12. Technology maturity levels

TRL	Europe Horizon 2020 terminology	DEA Technology Maturity Category
TRL1	Basic principles observed	Category 1
TRL2	Technology concept formulated	
TRL3	Experimental proof of concept	
TRL4	Technology validated in lab	Category 2
TRL5	Technology validated in relevant environment (industrially relevant environment in the case of key enabling technologies)	
TRL6	Technology demonstrated in relevant environment	
TRL7	System prototype demonstration in operational environment	Category 3
TRL8	System complete and qualified	
TRL9	Actual system proven in operational environment	Category 4

The mapping of the TRL framework results in the correspondence shown in Table 12:

- AEC and PEMEC are generally classified as Category 4 technologies, reflecting their commercial status and level of deployment. However, large-scale PEMEC installations may in some cases still be considered Category 3 due to more limited deployment at scale.
- SOEC technology is considered to fall within Category 3 across all capacities, reflecting its lower level of commercial deployment.

Learning Curves Applied

The learning curves and cost projections applied to build the future costs and assumptions described in the previous section and shown in the datasheets build upon a selection of literature sources looking at cast projections and learning curves [45] (Fraunhofer ISE (for Clean Air Task Force), 2022) [47] [48] [49] [50] [51] [52] [53] [54] [55] [44] [56] [57] [58] [24] [59] [60] along with deployment scenarios [61] [62] [63] [64] [65] and efficiency projections [66] [67].

The forward cost projections presented in the data sheets are derived using a standard experience-curve formulation commonly applied in energy technology modelling:

$$C = C_0 \left(\frac{Q}{Q_0}\right)^{-b}$$

Where:

- (C) = cost after cumulative deployment [€/kW]
- (Q) = cumulative deployment capacity [GW]

- (C_0) = initial cost [€/kW]
- (b) = learning exponent

The learning exponent is related to the learning rate through the following expression:

$$LR = 1 - 2^{-b}$$

Where *LR* represents the learning rate in %.

In the literature, there is a differentiation between system-level (or "blended") learning rates, and component-level (e.g. stack) learning rates. Typically, authors distinguish between stack learning rates and 'other component' learning rates such as BoS, BoP, power electronics or other components. On the other hand, many meta studies consider the use of 'blended' learning rates which consider a single learning rate for the full system and use regression to match this with historical costs. It should be noted that the component-level learning rates exhibit a wider spread than that of system-level learning rates. These two approaches can be harmonised by applying the individual components' learning rates, assuming the components' percentage share of unit capex and calculating from the resulting sum of unit capex one single learning rate that applies to the full system. This approach has also been applied in this study, with careful quality assurance of the effective component-level learning rates that are assumed. Below, the "blended" learning rates at system level are presented.

Although AEC, PEMEC and SOEC technologies exhibit relatively high technology readiness and maturity, learning effects are still applied when projecting future costs. Table 13 presents typical learning-rate ranges reported in the literature for each technology, together with the learning rates assumed in this study.

Table 13. Learning rates applied to each electrolyser technology.

Technology	System-level learning rate ranges from literature	Assumed system-level learning rate used in the datasheets
AEC	12-15%	14%
PEMEC	16-18%	17%
SOEC	18-22%	20%

Source: Data in the table is derived from DNV references combined with sources listed in the text.

The data sheets use 2025 as the base year, for which the baseline CAPEX values defined in Table 10 are:

- **AEC 100 MW:** 1,250 €/kW
- **PEMEC 100 MW:** 1,765 €/kW
- **SOEC 10 MW:** 2,950 €/kW

Future cost projections apply the learning rates presented in Table 13 together with the assumption that cumulative electrolyser deployment approximately doubles every three to four years. This assumption is aligned with global electrolyser project pipeline projections. In addition, cumulative gigawatt-scale deployment trajectories applied in the learning-curve calculations are assumed to be consistent with those used in DNV's ETO 2025 modelling.

In addition, beyond the use of scaling laws and projection formulae, DNV has also applied market knowledge and estimates provided in our projects for longer term initiatives, existing literature and calibration (for instance, to consider technology differences for the same project size) for coherence of the output values presented.

Uncertainty

Given that the electrolyser industry is still evolving, there are several key uncertainties associated with the data presented, particularly for the long-term projections to 2040 and 2050. One of the principal uncertainties in these forward projections relates to assumptions on cumulative electrolyser deployment, expressed in gigawatts. Should global deployment follow a trajectory different from that assumed, this could result in either significantly higher or lower installed electrolyser capacity than projected, with corresponding implications for the cost estimates across all three technologies.

A further source of uncertainty relates to the treatment of deployment across technologies. In the presented analysis, learning rates have been applied to cumulative electrolyser deployment as a whole, rather than modelling the deployment of AEC, PEMEC and SOEC technologies separately. This introduces an additional uncertainty, as individual technologies—particularly SOEC—may not scale at the same rate or to the same extent as AEC and PEMEC.

Uncertainty is particularly pronounced for the 2050 projections, reflecting the limited availability of reliable data when looking this far into the future. Ultimately, cost development will scale with cumulative global deployment, which may vary significantly towards 2050 under different scenarios.

For the 2025 cost data, uncertainty has been addressed by reviewing a wide range of literature sources and applying statistical analysis, including the assessment of average values and standard deviations. This approach has also been used to define upper and lower cost bounds within the presented 2025 cost data.

Offshore production

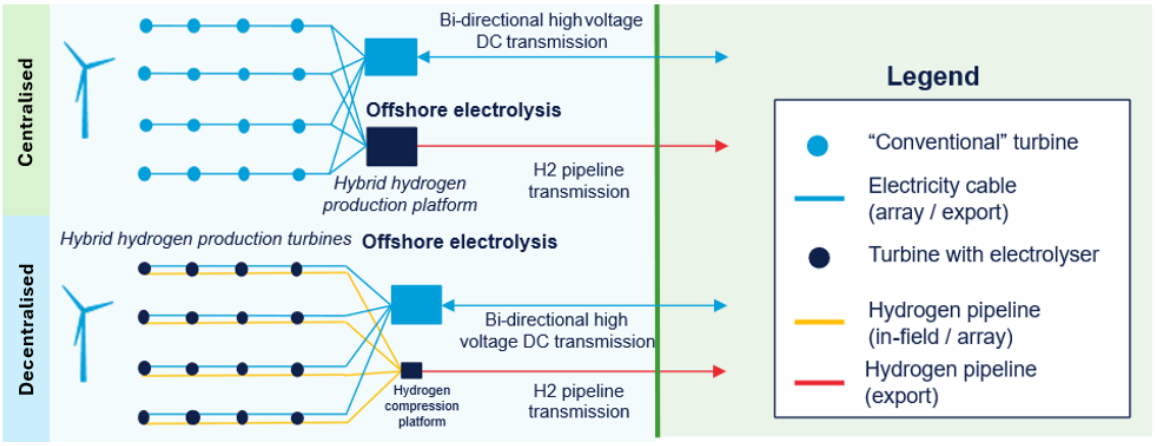
Offshore hydrogen production is increasingly considered as a complement to onshore electrolysis, particularly where offshore wind resources are strong and onshore grid capacity or land is constrained. Several studies have found that a hybrid offshore infrastructure, carrying energy to shore as both electricity and hydrogen, can lower the overall societal cost of bringing offshore wind energy ashore [75], [76]. Offshore production raises distinct technical, economic, and operational issues that affect technology choice, system design, and cost.

Offshore production concepts

Two offshore production concepts are commonly considered.

- In the centralised concept, electrolysis is performed at large scale on a central offshore platform or energy island that aggregates the electricity from several wind farms, and the hydrogen is transported to shore by pipeline or repurposed gas infrastructure.
- In the decentralised concept, electrolysers are placed at or integrated with individual wind turbines, so that hydrogen production is distributed across many installations; the hydrogen is gathered through array pipelines to a central offshore compression platform and exported to shore by pipeline.

Figure 14. Two offshore hydrogen production concepts. Centralised production is illustrated at the top and decentralised production is illustrated at the bottom.



There is also a third option, onshore centralised production, in which the wind farm is connected to shore conventionally by electrical transmission and the electricity feeds an onshore electrolyser plant. This is typically chosen where an onshore grid-connected substation already exists, allowing the electrolyser to draw a stabilised supply, with voltage fixed by the grid, while consuming the imported wind power. This option is not illustrated in Figure 12, as it is identical to a standard onshore plant and is covered by the onshore data sheets. The two offshore concepts, by contrast, differ in their implications for electrolyser sizing, redundancy, infrastructure sharing, and cost allocation.

Both concepts presented in Figure 14 have different implications for electrolyser sizing, redundancy requirements, infrastructure sharing, and cost allocation across the system. The third alternative mentioned has no implications for the electrolysis since the plant will be onshore and similar to the standard concept, so the 'non offshore' data sheets provided are valid for those.

Main system components

An offshore hydrogen production system typically comprises

- Offshore wind generation,
- Electrical infrastructure (offshore substations, transformers, and distribution),
- Electrolyser systems with their balance of plant, cooling, and control,
- Hydrogen compression, conditioning, and drying,
- Hydrogen export infrastructure such as pipelines, and
- Offshore platforms or floating foundations with their structural and marine systems.

Compared with onshore plants, offshore systems place greater constraints on compact design, marinization against corrosion, and reliability, because access for maintenance is limited.

Key advantages of offshore hydrogen production

Offshore hydrogen production offers several potential advantages that are relevant when assessing electrolyser technologies:

- Reduced reliance on onshore grid infrastructure, particularly in regions where grid connection capacity is limited or costly to expand

- Improved utilisation of offshore wind resources, including the potential to mitigate curtailment during periods of high generation
- Direct conversion of electricity to hydrogen offshore, which may reduce the need for high-capacity electrical export cables in some configurations
- Scalability at large project sizes, particularly in centralised concepts where very large electrolysis capacity can be deployed

These advantages may favour electrolyser technologies that can operate at large scale and integrate effectively with variable renewable power.

Key disadvantages and cost drivers

At the same time, offshore hydrogen production introduces additional cost and complexity relative to onshore installations:

- Higher CAPEX for offshore structures, including platforms, foundations, and marine installation
- Increased costs, for example for offshore electrical equipment, driven by stricter design standards, redundancy requirements, and environmental exposure
- More challenging operations and maintenance, due to limited access, weather constraints, and higher intervention costs
- Accelerated stress and degradation of electrolysis stacks due to direct connection to wind turbines and the fluctuations in their electricity supply, particularly for the decentralized alternative. Batteries may be required to be coupled in parallel to electrolysers to stabilise voltage.
- Stricter requirements on reliability and availability, which can influence technology selection and system redundancy

Electrolyser technologies with higher maintenance intensity or sensitivity to operational disturbances may face additional challenges in offshore environments.

Technology-specific considerations

Offshore deployment is generally considered for AEC and PEMEC. Alkaline brings maturity and proven large-scale deployment, but needs careful attention to footprint, dynamic operation, and maintenance access offshore. PEMEC offers higher power density and stronger load-following, which suit direct coupling to offshore wind and allow more compact designs, though typically at higher equipment cost. In this study, PEMEC is selected as the basis for the evaluation.

Key considerations for future projects

When integrating offshore hydrogen production into electrolyser technology assessments, several factors should be considered:

- The trade-off between centralised and decentralised offshore concepts, including scale, redundancy, and infrastructure sharing
- The interaction between offshore wind variability and electrolyser operational characteristics
- The impact of offshore installation and maintenance constraints on technology choice and system design
- The relative importance of footprint, modularity, and reliability in offshore environments

- The implications of offshore CAPEX on overall hydrogen production costs and competitiveness

Overall, offshore hydrogen production introduces additional cost and complexity compared with onshore solutions, but it also provides a pathway to unlock large-scale renewable hydrogen production in regions with abundant offshore wind resources. These offshore-specific factors should therefore be explicitly considered alongside conventional onshore metrics when comparing electrolyser technologies in future hydrogen system studies.

The offshore data in the datasheets

The data sheets include performance and cost data for two offshore hydrogen production concepts:

- A 10 MW PEMEC with offshore decentralised configuration
- A 100 MW PEMEC with offshore centralised configuration

The cost assumptions applied are based on a separate DNV study submitted to the Danish Energy Agency, which assessed offshore hydrogen production costs under a defined reference case. The study assumed 10 GW of offshore wind capacity located within the Danish Exclusive Economic Zone (EEZ), supplying 7 GW of electrolysis capacity, with cost estimates developed for 2040 and 2050. Offshore wind capacity was assumed to be allocated such that 70% was dedicated to hydrogen production and 30% to bidirectional high-voltage direct current (HVDC) electricity export.

The reference site was assumed to be located at a water depth of approximately 45 m and at a distance of 300 km from shore, with electrical export via an HVDC connection to Germany. Across all scenarios, hydrogen was assumed to be compressed from 30 to 80 bar and transported via a 28-inch pipeline sized to carry hydrogen equivalent to 6 GW_{LHV}, complemented by 14–16 inch collector pipelines.

The results of the study indicate that, for centralised offshore concepts, the offshore environment increases costs by approximately 132% in 2040 and 125% in 2050 relative to onshore production. For decentralised concepts, the corresponding cost increases are estimated at around 164% in 2040 and 159% in 2050. While these results are case-specific, they illustrate the magnitude of the cost uplift associated with relocating hydrogen production offshore.

Offshore hydrogen production is expected to see centralised offshore electrolysis reach commercial maturity by around 2040 and scale to multi-gigawatt levels by 2050. Offshore hydrogen could enable large-scale utilisation of North Sea renewable resources, support economic value creation in Denmark, and contribute to a secure domestic supply of green hydrogen for Europe.

The main assumptions and findings of the offshore cost study have been applied in the data sheets for both the 10 MW decentralised PEMEC configuration and the 100 MW centralised PEMEC configuration. Based on these assumptions, total CAPEX estimates have been derived for 2025, 2030, 2035, 2040 and 2050, as presented in the attached datasheets.

Energy/technical data

Typical total plant size

The typical total plant size represents the total input capacity of a single electrolysis plant or facility, expressed as MW thermal (MW_{th}) and reflecting the sum of all energy inputs to the process. In the dataset, this corresponds to the nominal electrical input capacity of the electrolyser plant (e.g. 10 MW, 100 MW or 1 GW), which is used consistently across technologies (AEC, PEMEC and SOEC) and configurations (onshore and offshore) to define plant scale. This parameter provides a standardised basis for comparing performance, costs and physical characteristics across technologies and over time, independent of downstream hydrogen output or utilisation assumptions.

Input

Electricity is the primary energy input across all electrolysis technologies and plant scales considered. For AEC and PEMEC, electricity accounts for 100% of the total energy input, expressed as 1 MWh/MWh of total input. In the case of SOEC, electricity represents a lower share of the total input, with the remaining fraction supplied as thermal energy for steam generation; the electrical share typically ranges from around 80–83%, depending on year and plant scale (generally assumed to increase in the future). All energy inputs are reported on a lower heating value (LHV) basis to ensure consistency across technologies.

Electricity consumption is reported both at stack level and at full system level, including balance of system (BoS) and plant (BoP). Stack-level electricity consumption decreases over time for all technologies, reflecting assumed efficiency improvements and higher current densities. For AEC and PEMEC, stack-level consumption typically declines from the low 50 kWh/kgH_2 range in 2025 towards the high-40s kWh/kgH_2 by 2050. When BoP losses are included, system-level electricity consumption is higher, typically by 3–6 kWh/kgH_2, but follows a similar downward trend over time. Lower values of BoP losses are generally achieved at larger plant scales in case optimized shared infrastructure is used between modules. SOEC exhibits substantially lower electricity consumption per kilogram of hydrogen at both stack and system level, reflecting the contribution of thermal energy input and higher electrical efficiency due to operating at high temperature.

Demineralized water required for electrolysis is treated as an auxiliary input and is expressed in kg per MWh of electrical input. Water consumption varies by technology but remains broadly stable over time for a given configuration, as it is largely governed by electrochemical stoichiometry, electro-osmotic drag and cooling flow requirements rather than efficiency improvements. AEC and PEMEC show comparable water requirements (with PEMEC being a little higher), while SOEC generally exhibits higher water demand per unit of electrical input due to differences in process configuration and the inclusion of steam generation.

Output

Across the dataset, outputs are reported relative to the total electrical energy input (expressed as % of total input_e or MWh/MWh of total input_e) and include LHV- and HHV-based efficiencies, process heat losses (split into recoverable and unrecoverable fractions), and the resulting hydrogen yield (kg/MWh input_e). For AEC and PEMEC, the LHV-based efficiency (reported on a system basis including BoP) improves over time and for AEC it is slightly higher at larger scales. Over time, AEC LHV efficiency is estimated to increase from ~58–61% in 2025 to ~64% by 2050 across 10–1000 MW; while PEMEC increases from ~58% in 2025 to ~64% by 2050 at 10–1000 MW). The corresponding HHV-based efficiencies are higher (reflecting the HHV/LHV difference) and follow the same improvement trajectory.

Heat loss (as a share of electrical input) decreases over time for AEC and PEMEC, and the dataset separates heat into an unrecoverable fraction and a fraction recoverable for district heating (typically a few %-points of electrical input for AEC/PEMEC, with the recoverable share varying by year). Where specified in the notes, the recoverable heat is associated with low-temperature heat levels (e.g. around 50°C, with an expectation of higher temperatures in future assumptions). For SOEC, the sheets explicitly include both electrical and thermal inputs in the energy balance, and the output section reflects this by showing materially different efficiency and heat-accounting patterns compared with AEC/PEMEC.

Hydrogen yield, expressed as kg H_2 per MWh of electrical input, increases over time for all technologies in line with decreasing specific electricity consumption (e.g. AEC rises from ~17.5 kg/MWh in 2025 towards ~19–20 kg/MWh by 2050; PEMEC from ~17.4 kg/MWh towards ~19.2 kg/MWh; SOEC at 10 MW is higher, rising from ~25 kg/MWh towards ~28.6 kg/MWh).

Forced and planned outage

Plant availability and lifecycle parameters are also reported in the output section: forced outage assumptions are typically a few percent and, for AEC, decline over time (e.g. ~4% in 2025 reducing to ~2% by 2050 at 10 MW), while PEMEC and SOEC are commonly held at around 4.5% and 5.5% respectively in the current data. Planned outage allowances range from roughly one to two weeks per year depending on technology/scale and generally decrease in later years for some configurations.

Technical lifetime

Technical lifetime assumptions increase with time in several cases (e.g. AEC increasing from ~20 years towards 25 years), while SOEC is often kept constant at around 25 years in the current sheets. Stack replacement intervals (hours) lengthen over time across technologies (e.g. AEC from ~70,000 h towards ~88,000–100,000 h; PEMEC from ~60,000 h towards ~76,900–100,000 h; SOEC from ~25,000 h towards ~70,000–90,000 h), reflecting assumed durability improvements.

Construction time

Construction time is presented as an indicative schedule duration that decreases over time but increases with plant scale (e.g. around ~1.75–3 years for 10 MW in the near term, rising to ~2-3.5years at 100 MW and ~3-3.5 years at 1 GW in 2025, with shorter durations assumed in later years). Normally the prefabricated plant concept is able to reduce construction time to lower band values while the critical path is common to the entire industry which is marked by reception of long lead items and integration & commissioning works.

Financial data

All financial figures in the catalogue are expressed in real Euro (€), at a 2025 price level, and exclude VAT and other taxes.

Investment costs are reported as specific investment per unit of total electrical input capacity (€/kW of total input_e) and are provided both as total values and as a detailed breakdown by major cost components. Across all technologies, specific investment costs decline over time and decrease with increasing plant scale, reflecting assumed learning effects, standardisation, and economies of scale. The degree in which such investment costs decline is also marked by the plant design concept, with higher margin for cost savings in the case of the chemical plant concept due to integration benefits and aggregation of balance of plant equipment into larger assets versus the prefabricated approach. AEC exhibits the lowest specific investment costs across all scales, followed by

PEMEC, while SOEC remains the most capital-intensive in early years but shows comparatively steep cost reductions towards 2050. Offshore PEMEC configurations show substantially higher costs than onshore plants, driven by offshore installation, marinization, and support structure requirements, with cost reductions over time subject to higher uncertainty.

The cost breakdown is aligned with the electrolysis cost taxonomy as presented in Table 8. Stack costs decrease over time for all technologies, reflecting assumed learning rates that differ by technology (lower for AEC, higher for PEMEC and SOEC), and they represent a smaller share of total investment at larger scales. BoP costs also decline with scale but remain a significant cost component especially at lower scales, particularly for PEMEC and SOEC. Additional cost items include compression, which shows limited cost-down potential over time, as well as control systems and civil infrastructure, which scale weaker with capacity and therefore contribute proportionally more at smaller plant sizes.

Finally, indirect costs—covering engineering, procurement and construction (EPC) service - are reported explicitly and applied as technology- and scale-dependent mark-ups in line with DNV energy technology outlook assumptions. For onshore plants, indirect costs typically decline as a share of total investment with increasing scale, while for offshore configurations they remain comparatively stable due to installation complexity and project risk. For the chemical-plant configuration, EPC costs will be higher than for the prefabricated configuration. Therefore, the prefabricated configuration has potential to reduce overall cost, in case direct cost allocation for certain scope items is cheaper than the equivalent EPC cost.

Investment costs

Investment costs are reported as specific investment per unit of hydrogen production capacity, expressed in €/kg H_2/day of maximum output. This metric is derived by normalising the total investment cost by the typical total plant size and corresponding maximum hydrogen output, allowing consistent comparison across technologies, scales, and configurations.

Across all technologies, specific investment costs decrease with increasing plant scale and decline over time, reflecting assumed learning effects, standardisation, and economies of scale. AEC shows the lowest specific investment costs, followed SOEC. Offshore PEMEC configurations display significantly higher specific investment costs than onshore plants, driven by offshore installation requirements and support structures, as well as equipment marinization requirements, with cost reductions over time subject to higher uncertainty. It is important to note also that economies of scale are limited by the concept and design approach of the plant, being more relevant when the balance of plant and balance of system can be integrated and aggregated into larger units (chemical plant concept) than in the case of prefabricated plants.

Where reported, the investment cost is split between equipment and installation. Equipment costs cover the electrolyser stacks, system components, balance of plant, and associated machinery, while installation costs include engineering, civil works, buildings, grid connection, installation, and commissioning activities. For AEC and PEMEC onshore plants, equipment typically accounts for the biggest share at around 61-62%, with installation around 38-39%, reflecting the modular and equipment-dominated nature of these projects. The share for installations cost is expected to decrease in the future, reflecting on a learning industry.

Economy of scale

Electrolysis plant costs can be significantly reduced through economies of scale, particularly with the chemical plant approach. Several system components do not scale

linearly with increasing capacity, in particular, vessels, tanks, and piping, which constitute a substantial share of both the electrolyser system and the balance of plant (BoP). In addition, with a larger size it is possible to centralize certain balance of plant equipment into single units as it is the case of using large purifier trains (100 MW electrolysis train level) or chiller/cooling infrastructure (at +100 MW plant level). Besides (also applicable for prefabricated plants), unitary EPC costs reduce with the scale as it is the case with civil works because there are certain fixed initial costs prior to the execution of the construction works which maintain or do not grow in the same rate as the plant size.

A commonly used rule of thumb to represent such effects is the six-tenths rule (also known as the 0.6 rule). This method, widely applied in chemical engineering, uses a power-law relationship to estimate the capital cost of a larger plant or component based on the cost of a smaller reference unit. The approach reflects the fact that while costs increase with capacity, they do so at a slower, non-linear rate, typically by raising the capacity ratio to the power of 0.6.

However, DNV's numerical analysis of electrolyser components indicates that economies of scale are less pronounced for electrolyser systems than suggested by the six-tenths rule alone. The effect is moderate for balance of stack (BoS) and Balance of Plant (BoP) equipment and even more limited for electrolyser stacks, largely due to their modular design. A significant share of total electrolyser plant costs is associated with the stacks, and once the maximum stack capacity is reached—currently in the range of approximately 0.1–10 MW depending on the technology and manufacturer—further scaling is achieved by deploying additional stacks in parallel. As a result, stacks exhibit limited economies of scale beyond the first few megawatts, whereas BoP components continue to benefit from scaling effects.

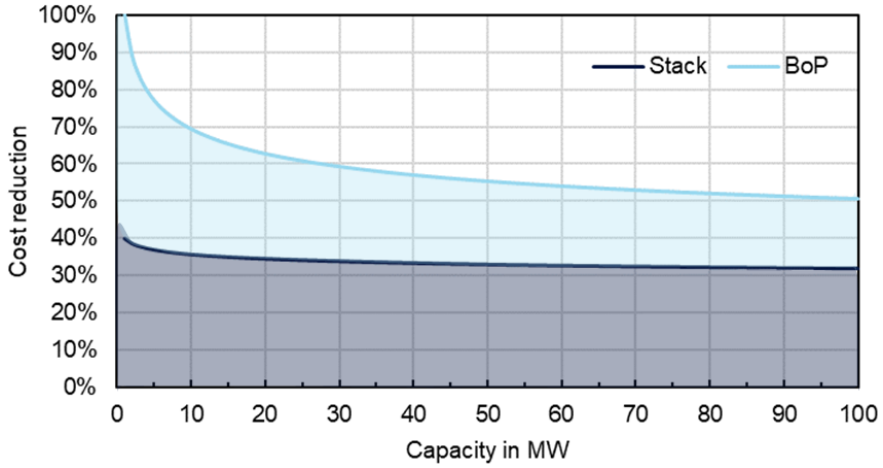
To reflect these differences, a simplified scaling approach is applied in which separate scaling exponents are used for stacks and Balance of System (BoS) or Balance of Plant (BoP) components, both of which are hereafter referred to as the BoP scaling law. Based on manufacturer data and publicly available sources, DNV identified appropriate scaling relationships for each component. Stack costs were found to scale close to linearly, with a scaling exponent of 0.95 providing a good fit, while BoS and BoP costs showed stronger economies of scale, represented by a scaling exponent of 0.75. These scaling factors can be applied to estimate the cost advantage associated with increasing plant capacity, using the system cost values presented earlier. The resulting economies of scale apply only to electrolyser system costs and explicitly exclude installation costs.

$$\text{Scaling advantage stacks (\%)} = (\text{Electrolyser capacity (MW)}^{0.95})/\text{Electrolyser capacity (MW)} * 100\%$$

$$\text{Scaling advantage BoS + BoP (\%)} = (\text{Electrolyser capacity (MW)}^{0.75})/\text{Electrolyser capacity (MW)} * 100\%$$

Beyond electrolyser capacities of approximately 500 MWel, the cost advantage from further scaling diminishes and scaling behaviour approaches linearity. This reflects practical limits to economies of scale, as component sizes cannot increase indefinitely. In addition, this scaling approach is inherently simplified. As the electrolyser industry continues to develop and mature, other cost drivers—such as supplier-specific pricing, design choices, and market dynamics—may influence observed scaling behaviour. In particular, plant designs based on a high degree of repetition of smaller-scale units tend to exhibit more limited economies of scale. The impact of economies of scale is illustrated in Figure 15, based on an indicative cost split of 40 per cent for stacks and 60 per cent for BoP.

Figure 15. Economies of scale for an electrolyser system



Operation and maintenance (O&M) costs

Operation and maintenance costs in the dataset are reported as fixed O&M expressed as a percentage of the specific investment per year (% of specific investment/year), representing average lifetime costs. Fixed O&M covers all costs that are largely independent of operating hours, including administration, operational staff, O&M service contracts, insurance, network and system charges, and routine maintenance activities.

Across all technologies, fixed O&M costs decline over time and generally decrease with increasing plant scale, reflecting assumptions of improved reliability, standardisation and operational learning. AEC shows fixed O&M levels typically decreasing from around 5–5.5% of specific investment per year in early years towards 2–3% by 2050, depending on scale. PEMEC follows a similar trajectory but at slightly higher levels for smaller plants, while SOEC exhibits the highest fixed O&M shares, particularly in early years (6.5-7%), reflecting higher complexity and lower technology maturity. Offshore PEMEC configurations show substantially higher fixed O&M shares than onshore plants, driven by offshore access, marine operations and maintenance logistics. Variable O&M costs are not explicitly quantified in the current dataset. Stack replacement costs are not included in the fixed O&M %.

Technology specific data

Technology-specific parameters describe the electrochemical performance, operational flexibility, durability, and physical characteristics of the electrolysis systems.

Current density increases over time for all technologies, reflecting assumed improvements in cell design and materials. For AEC, current density increases from around 0.5 A/cm² in 2025 to about 2.0 A/cm² by 2050, largely independent of plant scale. PEMEC operates at significantly higher current densities, increasing from around 2.5 A/cm² in 2025 to 10 A/cm² by 2050, enabling much higher power density. SOEC current densities are lower than PEMEC but increase from roughly 0.8 A/cm² towards about 2.0 A/cm² by 2050. Higher current density raises the capacity of a stack of given physical size. The physical dimensions of AEC and PEMEC stacks are assumed to stay close to today's values, constrained by transport and handling, so the rising current density is expressed as a higher MW rating rather than a larger stack. On this basis AEC stack capacity rises from current commercial values towards about 20 to 25 MW per stack by 2050, and PEM from around 1.5 to 2 MW towards about 6 MW over the same period.

Operational flexibility is characterised here by cold-start time, the time required to ramp from zero to full load. PEMEC is the most flexible, with cold-start times of around 5 to 15

minutes in 2025, decreasing to well below 5 minutes in later years. AEC flexibility depends strongly on whether the system is atmospheric or pressurised. Atmospheric AEC has slow dynamics, with cold-start times of around 60 minutes in 2025, improving towards about 30 minutes by 2040 to 2050. Pressurised AEC is more responsive, with hot-start times of the order of 5 - 15 minutes and ramp rates triple that of atmospheric AEC. SOEC has markedly longer start-up times because of its thermal inertia, of the order of several hours in 2025 (about 480 to 600 minutes), decreasing over time but remaining well above the low-temperature technologies.

Degradation rates, expressed as % per 1,000 operating hours, decrease for all technologies over time. AEC declines from around 0.13% per 1,000 h in 2025 towards about 0.11% by 2050, while PEMEC follows a similar path from a slightly higher starting point. SOEC shows higher degradation in early years, around 0.75% per 1,000 h at 10 MW in 2025, falling towards about 0.12% by 2050.

The physical characteristics are described by stack footprint, total plant footprint, and stack weight, each normalised per MW of electrical input. For AEC, stack footprint decreases from around 10 m²/MW in 2025 to about 2.5 m²/MW by 2050, and total plant footprint from roughly 280 m²/MW to about 70 m²/MW over the same period at 10 MW scale. PEMEC achieves more compact, with stack footprint falling from around 7.5 m²/MW to about 1.9 m²/MW and total plant footprint from around 200 m²/MW to about 50 m²/MW. SOEC requires more space, with stack footprint around 20 m²/MW in 2025 falling to about 8 m²/MW by 2050, and correspondingly larger plant footprints. The footprint reductions over time follow directly from the assumed increases in current density, so they should be read as conditional on those assumptions.

Stack weight per MW follows similar trends and falls over time for all technologies. PEMEC is the lightest, decreasing from around 1 tonne/MW towards about 0.3 tonne/MW, followed by SOEC, while AEC is the heaviest of the three.

Together these parameters illustrate the fundamental differences between the technologies in power density, flexibility, durability, and space requirement. PEMEC and pressurised AEC are both well suited to dynamic operation, while atmospheric AEC and SOEC are more constrained – SOEC by thermal inertia and atmospheric AEC by slower ramping – but both offer other system-level advantages depending on the application. Ultimately, the appropriate technology depends on the application, the operating regime, and the scale.

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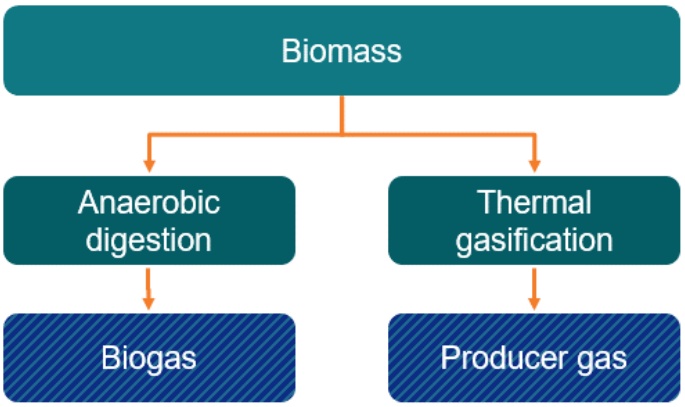
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Biogas and Producer Gas



This section focuses on the production of gaseous fuels as biogas and gasification producer gas. These gases can be used directly for heat and electricity production or can be cleaned and upgraded to other fuels. The upgrade to other products can be found in the suqseqent chapters of this catalogue. Methane production is handled separately in under its own section.

The type of biomass feedstock for these processes is to a large extent different, whereas anaerobic digestion uses a wet type of biomass as organic waste or manure, thermal gasification requires dry biomass as woodchips or straw. Straw can on the other hand also be used for biogas production.

The section includes the following chapters:

81 Biogas Plants

83 Gasifier Plants (for Producer Gas)

81 Biogas Plants

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Date	Ref.	Description
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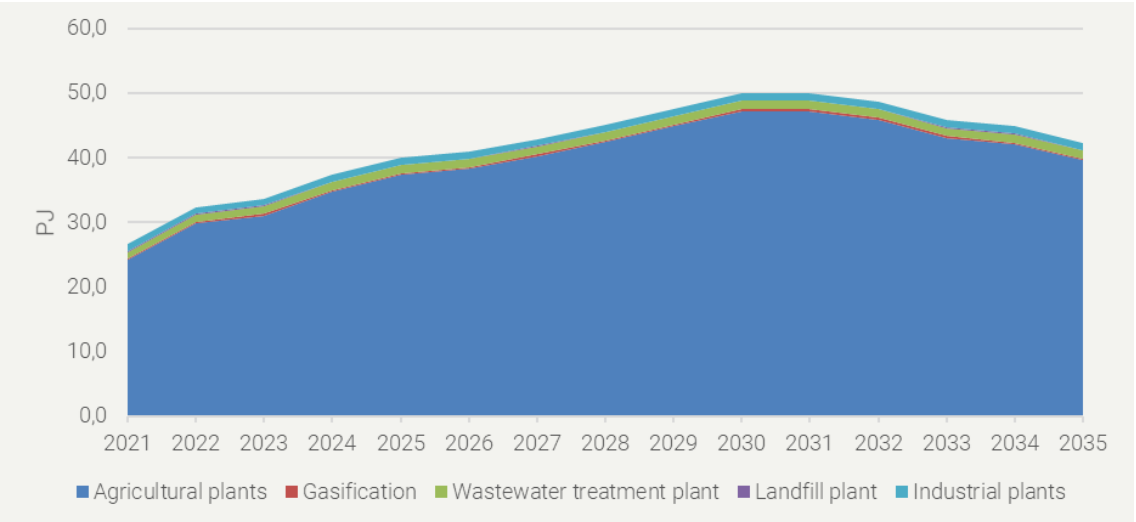
Qualitative description

Brief technology description

In biogas plants, organic matter is biologically converted under anaerobic conditions into a methane (CH_4) and carbon dioxide (CO_2) rich gas and digestate. The biogas can be used in industrial processes, for producing heat and electricity or for upgrading to biomethane, which can be injected into the natural gas grid. Upgrading of biogas to biomethane is treated in a separate chapter of this technology catalogue (chapter 93).

Today, Danish biogas production is distributed across approx. 100 agricultural biogas plants that use liquid livestock manure (slurry) as their main input co-digested with residues from households, industry and agriculture, 50 wastewater plants and 27 landfill biogas plants. Biogas production has been increasing significantly the past decade, and the increase is expected to continue. The biogas production from agricultural plants amounted to 25 PJ in 2021-22 and it accounts for the major share of total Danish biogas production [1]. As illustrated in figure 1, the forecasted future increase in biogas production is expected to come from new agricultural biogas plants. It is noted that Figure 1 is based on Denmark's Climate Status and Outlook 2023, which refers to a frozen-policy scenario.

Figure 1: The expected deployment of biogas production - by type of biogas plant (2021-2035) [2]



This chapter solely addresses biogas production at agricultural plants. The technology datasheets cover larger plants, compared with the previous standard-sized plants, with an approximate yearly biogas production of respectively 25 and 50 mill. Nm^3 (equivalent to 3,000 and 6,000 Nm^3 CH_4 per hour). These plant sizes are expected to represent the market standard, as most new plants are built with a biogas production capacity within this range [3].

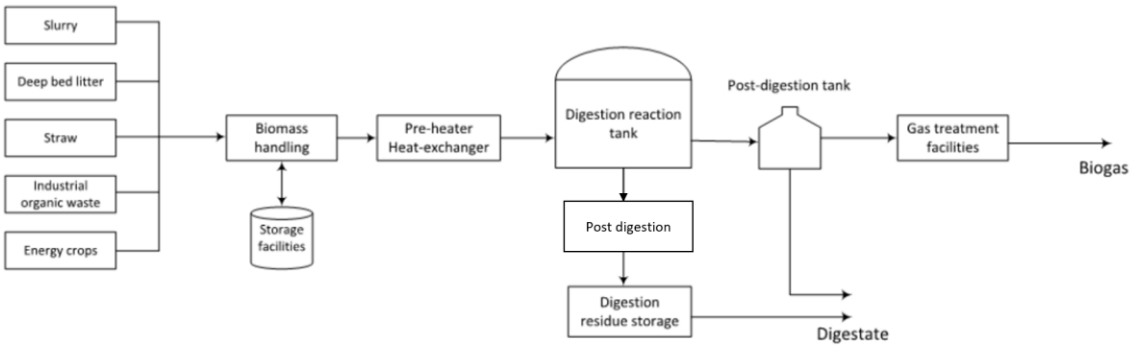
The input of biomass is usually transported to the plant by road, but there are also plants where the low dry matter (DM) feedstock is pumped in pipes, thereby reducing local nuisance from truck transport [4, 5]. The biomass is received and stored in pre-storage tanks and later processed in digesters (reactors). The digesters are normally heated to either 35 - 40 °C (mesophilic digestion), or 50 - 55 °C (thermophilic digestion). For new biogas plants with gas upgrading, the heat in the digesters will typically be supplied with excess heat from the upgrading facility. For plants that are not connected to an upgrading plant, the heat supply depends on local conditions, and can be supplied by either boiler (gas or biomass-fired), heat pumps, or district heating.

Typical processing time in the digesters (Hydraulic Retention Time, HRT) is assessed to be between 60 and 100 days in newer Danish plants, depending on the biomass input and the plants' technical specifications [3]. It is further stated that 65-70% of Danish biogas plants have an HRT of 60-80 days, with a tendency towards longer retention time for new biogas plants [3]. An overview of existing plants indicates that the HRT is a bit shorter on average, with a weighted average of 47 days [6]. As new plants are expected to have a longer HRT than the existing, the calculations in the datasheets are based on an HRT of 65 days.

Danish plants use continuous stirred-tank reactors (CSTR). This implies continuous removal of a small quantity of digested biomass from the digesters and replacement with a corresponding quantity of fresh biomass, typically several times a day.

After being processed in the main digester, the digestate is pumped to post-processing tanks where post-digestion takes place and additional gas is produced and collected. After the biogas process, the volume of the digestate is roughly the same, or slightly reduced, as that of the initial feedstock. The digestate can be recycled as a fertilizer in agriculture either directly or after being separated into solids and fluids. Finally, the gas is treated to reduce water and sulphur contents to the desired concentrations. Figure 2 shows the typical components and flows in a biogas plant.

Figure 2: Typical components in a biogas plant



Ann.: OBS alternative terms for technical description might be used by some actors: e.g., biomass handling might be referred to as pre-treatment, digestion reaction tank as a digester and digestion residue storage as digestate storage or post-storage of digestate.

The composition of the biomass input (feedstock) is important for the economy, and for the dimensioning and operation of the biogas plants. As the existing plants use CSTR, they are built to handle pumpable biomass, i.e., slurry and wet industrial waste [7].

There is an upper limit to how much high DM feedstock, e.g., straw, that can be handled in a CSTR. This is due to the risk of floating layers and the longer decomposition time of straw and similar biomasses. In the last couple of years, there has been a technical development towards biogas plants being able to handle a larger share of biomass with a high DM content, such as deep litter and straw. For instance, floating layers are prevented with an increased stirring frequency. Based on the current market standard, biogas plants should be operated with a DM content of a max of 13-14% in the reactor [8, 3]. Recirculation of the liquid fraction after separation of the digestate makes the biogas plants able to use a higher share of feedstocks with a high DM content, as it will be mixed with the liquid digestate and thereby decrease the average DM content in the reactor [3, 9]. It is also noted, that the DM content of the input biomass is higher than the DM content in the reactor due to degradation of part of the DM input.

Input

Biodegradable organic material such as livestock manure/slurry, organic waste from food processing and households, agricultural residues (e.g., straw), energy crops, etc.
Electricity for mechanical processing equipment
Process heat for preheating and heating the reactor tanks

Output

Biogas and digestate, for use as fertilizer.

Biogas typically contains 50-75% methane (CH_4), 25-50% carbon dioxide (CO_2) plus a minor content of hydrogen (H_2), nitrogen (N_2), oxygen (O_2), hydrogen sulphide (H_2S), and ammonia (NH_3). The composition of the biogas varies with the specific mix of the feedstock.

Quantitative description

Energy balance

In the biogas industry, it is not common practice to measure the energy content of the input material as a calorific value, as is often done for other energy conversion technologies in this catalogue. Instead, the input is measured as tons of biomass along with information on the amount of dry matter in the input, expressed by the DM factor, and the share of organic materials, expressed by the share of volatile solid (VS). Using the energy balance as a yardstick for comparing different technologies is mainly interesting for biomasses (or other energy sources) with alternative uses such as straw, energy crops or certain types of industrial waste, which e.g., could be used in combustion plants or in thermal gasification processes. The lack of focus on the energy balance for biogas plants is partly due to difficulties in measuring the energy content of the input biomasses. Further, the high-water content and fertilizer value of some of the biogas feedstocks, particularly slurry and manure, make them unsuitable for combustion in traditional energy plants both seen from an energy production perspective and a nutrient recycling perspective.

To calculate the energy balance of biogas production, the energy content of the biomass going into the plant and the output of biogas need to be calculated. Table 1 provides an overview of the energy content of some of the most used biomasses. The energy content depends on the DM content, the VS share, and the calorific value of the biomass. The energy content is directly proportional to the DM content and the VS share. Further, the VS share of the DM represents the fraction of the DM that may be transformed into energy.

Dry matter and volatile solids [22]

The Dry Matter (DM) content is the mass of solid remaining after a sample has been dried in an oven at 103°C for 24 hours, divided by the original mass of the sample.

The Volatile Solid (VS) measures the organic matter of a liquid or slurry. From a chemical perspective, the organic matter is the part that burns, and this is also the portion that may potentially be converted to biogas. Important to mention, most plants and other material that a nonprofessional would term as organic contain a portion of inorganic matter.

To determine the share of VS, the DM sample is heated at 550°C for 1 hour. The lost mass is the VS. The remaining part, the ash, is also called the fixed solids (FS).

Table 1: Data and energy balances for selected biomasses.

	DM content	VS Share	The energy content of input in GJ per ton of VS	The energy content of output gas in GJ per ton of VS	Conversion efficiency, biomass to biogas
Straw	85%	95%	17.4	9.5	55%
Slurry	4.5-7%	80%	n.d.	9.2	-
Maize	31%	95%	17.5	11.6	66%
Grass	32%	90%	18	11.5	64%
Beet	18%	95%	17.1	13.2	77%
Beet greens	12%	85%	18.2	12.4	68%

Output data is given under the assumption of a retention time of 65 days. The conversion efficiencies will vary from plant to plant depending on the specific operations characteristics and specific properties of the biomass – and thus the values are only guiding. Longer retention time would increase the output from the plant and hence the conversion efficiency, and vice versa with respect to shorter retention time.

The conversion efficiency (biomass to methane) depends on several factors, including the composition of the feedstock, the processing time, the organic loading rate, and the effectiveness of process control. Fatty biomasses, proteins, and certain carbohydrates (sugars and starches) are relatively easily converted to biogas, whereas only part of the cellulose is converted, and almost none of the lignin [10].

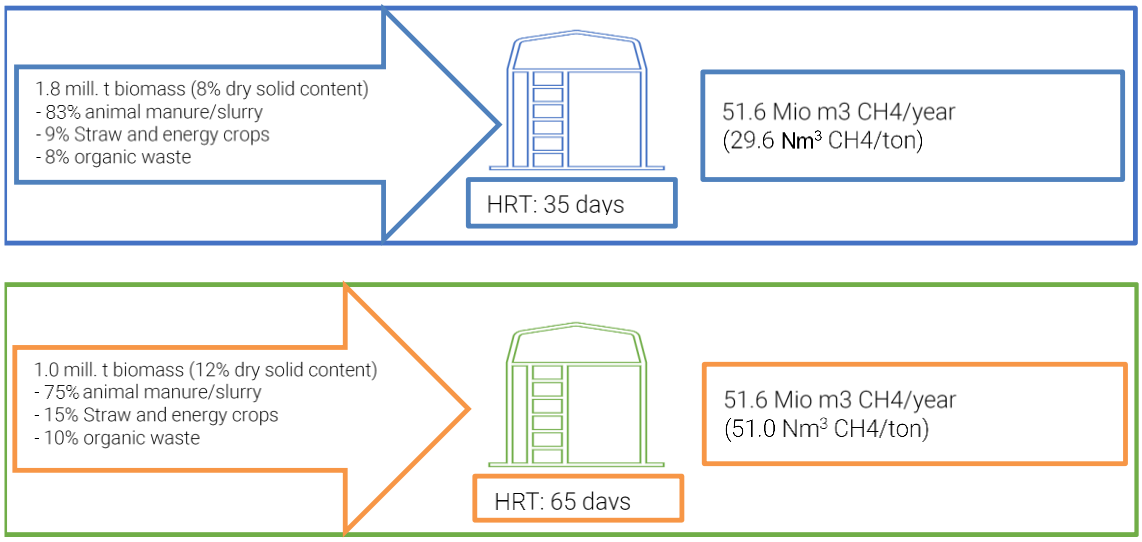
As an example, the energy content of straw is approx. 17.4 GJ per ton of VS. When straw is used as feedstock in a biogas plant with an HRT of 65 days, 260 Nm^3 methane/ton biomass will be produced, with an energy content of 9.5 GJ. When comparing this to the energy content of straw it implies that approx. 45% of the energy content is not converted into gas. As mentioned, some biomasses are more easily converted to biogas than others, thereby giving a higher biogas yield per ton of biomass being digested. The “energy loss” therefore depends on the type of biomass

input as well as the HRT in the plant. When using a large share of straw, the energy loss will decrease if the HRT is increased, e.g., to 80 days instead of 65. Thus, conversion efficiency and methane production per ton varies depending on HRT, and the difference in methane output for straw and industrial waste would accordingly be different than the one displayed in Table 1 if the HRT had been different. It is noted, that the energy loss should not be perceived as a loss *per se*; thus, the carbon not converted to energy is not lost but returned to the fields, where it is stored and contributes to plant growth.

The heating value of biogas depends on the share of methane, which depends on the type of feedstock and the production pathway [11]. Therefore, measuring the output in Nm^3 methane rather than Nm^3 biogas is practical to allow comparisons across plants. Methane has a lower heating value (LHV) of 35.9 MJ/Nm^3, whereas biogas with a 65% methane content has a LHV of 23.3 MJ/Nm^3.

Figure 3 shows an example of two different plants that produce the same volume of biogas. The first plant uses 1,770,000 t of biomass with an HRT of 35 days, while the second uses 1,030,000 t of biomass and has an HRT of 65 days. The reason why the output is the same is due to differences in feedstocks and HRT across the two plants [12].

Figure 3: Biogas output in two different biogas plants – based on EUDPs very large plants (4a-A and 4b-A).



Biomass input in biogas plants

The Danish legislation requires that all biogas plant operators annually report the volume in ton of the feedstock used in their plant to the Danish Energy Agency.

Table 2 shows the average feedstock composition for biogas production at agricultural plants in the accounts for 2021-22; the table also specifies the average estimated DM content and methane yield for the different biomass input categories.

Table 2: Biomass for biogas production in Denmark 2021-22 [1].

	Input share (%)	Methane production (% of total)	Dry matter content (%)	Methane output per ton biomass input (GJ CH4/tons)
Cattle slurry	40.4	12.7	7.6	0.5
Pig slurry	25.2	6.8	5.0	0.4
Deep litter and similar	8.4	11.3	30.0	2.2
Energy crops	5.3	11.7	32.1	3.6
Straw and similar	2.0	9.3	84.3	7.5
Industrial waste	12.7	28.0	29.9	3.5
Glycerine and similar	1.5	12.0	75.0	12.7
Household waste	3.4	5.7	18.0	2.7
Other crops and residues	1.2	2.4	30.0	3.3
Total/average	100	100	16.1	1.6

Ann.: The biomass category "Mixed slurry" from the accounts is included in the "Cattle slurry" and "Pig slurry" categories in the table with 50% allocated to each of the categories.

The methane production per unit of input material for straw, glycerine, and industrial waste substantially exceeds that of slurry. Specifically, the methane yield is 25-fold greater for glycerine, 15-fold greater for straw, and 7-fold greater for industrial waste, when compared to slurry. These differences are to a large extent the result of differences in DM content. However, differences in chemical composition also play a role. Thus, the degradability and conversion into biogas is faster for industrial waste than for straw, due to straw having a high content of cellulose, while industrial waste is often characterised as a fatty feedstock. Because of the variation in methane output per ton of feedstock, the methane output of plants with a certain input capacity (measured in tons) varies depending on the feedstock, HRT, and several other factors as mentioned above. In 2016-17 the average methane yield in Danish biogas plants was 1.21 GJ per ton of biomass, this has increased to 1.60 GJ/ton in 2021-22 [1].

While feedstocks with high DM content may yield a high methane output per ton of biomass input, some feedstocks, e.g. straw and deep litter material, also set additional requirements for pre-processing systems and increase the auxiliary energy consumption. Straw for instance needs to be mechanically chopped, grounded, or otherwise treated before being fed into the digester to avoid a floating layer and obtain a shorter processing time. Thus, higher yields when using feedstocks with a high DM content are usually followed by increased investment and O&M costs [9]. In addition, they may also require post-treatment of the digestate in order to optimize the value as fertilizer and fulfil the legislation regarding emission of ammonia.

The composition of biomass input primarily depends on three factors:

- Cost and availability of the biomass
- Technological barriers/cost
- Regulation

Typical capacities

Most of the biogas production in Denmark is based on large CSTR plants with co-digestion and upgrading facilities with direct injection of renewable gas into the gas grid.

The capacity of existing plants varies from small-scale plants, with a biomass input of 10,000 tons/year, to large-scale plants with a biomass input of 1 million tons/year. The average annual biomass input for existing agricultural plants is approx. 160,000 tons. The production of methane depends on the capacity of the plant combined with the composition and DM content of the feedstock and the HRT in the digester. Looking forward, the majority of the biogas is expected to be produced at plants with a production capacity of 25 to 50 mill. Nm^3 CH_4/year (corresponding to a feedstock input on approx. 0.8-2 mil. tons/year, depending on the composition of the feedstock).

Regulation ability

Biogas production at existing facilities can be increased by adding organic materials with high methane potential or by prolonging the HRT. However, there is a biological limit to how fast production can be regulated. For example, a biogas plant digesting only animal slurry during summer may increase the gas yield from 14 Nm^3 methane per ton to about 45-50 Nm^3 methane per ton during a period of three to four weeks if feedstock with a higher methane output is added [13]. Regulation of the production may require additional feedstock storage capacity. Seasonal regulation of biogas production is mainly relevant for biogas plants that are directly connected to a combined heat and power plant supplying a heat load, as heat demand is higher during winter.

Today, the flexibility of biogas is mainly related to the possibility to store the upgraded gas in the large storage facilities that form an integral part of the Danish and European gas grid.

Biogas plants typically have short-term storage in connection with the facility. For new biogas plants with an upgrading facility, the storage will most likely be in connection to the upgrading facility and with a capacity equivalent to half an hour's production on a large biogas plant. The cost of the storage is therefore not included in the datasheet for biogas production but as a part of the upgrading facility (Chapter 93 in the technology catalogue).

Area requirements

The area requirements vary depending on the design and layout of the plant. The area requirements, including the area for storage, of newer plants range between 2 and 9 ha depending on the capacity and the type of biomass input. Biogas plants are typically placed in open farmland, in order to minimize transport costs of the biomasses. Figure 4 shows two Danish biogas plants, SBS Kliplev from 2022 (to the left), which occupies an area of 8.5 ha and has an annual production of 48 million Nm^3 of biomethane, and Solrød biogas from 2015 (to the right), which occupies an area 2 ha and has an annual production of 11 million Nm^3 biomethane.

Figure 4: SBS Kliplev (left) is approx. 8.5 ha and has an annual production of 48 milion Nm^3 of biomethane, whereas Solrød Biogas (right) is approx. 2 ha and has an annual production of 11 million Nm^3 biomethane



Source: [23, 21]

Advantages:

When manure is used for biogas production, the emission of greenhouse gasses from handling and storage of manure is reduced
Wet and difficult biomasses with few or no alternative uses can be transformed into a high-value energy carrier (biomethane) that may be stored at a low cost in the existing gas system
The output gas contains a high level of CO_2, which makes it attractive for subsequent carbon capture and storage (CCS) or carbon capture and utilization (CCU)
Saved expenses for handling and storage of slurry
Important nutrients, primarily nitrogen, phosphorus, and potassium, are recycled and can be redistributed between farms. The risk of nitrate leaching is also reduced
The fertilizer value of the digested biomass is improved compared to that of the untreated slurry. The fertilizer value is also better known and documented, and it is, therefore, easier to apply the right dose to the crops. With a high content of straw in the biomass input, the share of nitrogen (in ammonia form) in the digestate will, however, often be low and the balance between nitrogen and phosphor and potassium may be less optimal for the farmer. Also, a high content of straw may have a negative effect on the viscosity of the digestate. Thus, the improvement in fertilizer value is reduced/not applicable, for digestate from biogas plants with a high share of straw. The problems can however be addressed by separating the digested biomass in a liquid and dry fraction.
For waste fractions with high water content, co-digestion of manure and waste can often provide a low-cost option compared to other forms of waste handling, such as incineration
Application of digestate reduces smell compared to the application of raw slurry
When straw is used as feedstock and the digestate from the biogas production is used as fertilizer, the content of carbon in the topsoil is not depleted, as it would be if the straw was incinerated in boilers or power plants

Disadvantages:

Methane emissions from biogas plants are unavoidable but can be kept to a minimum if monitored and handled properly (e.g. cooling and coverage) [14]
Use of straw and other solid biomasses in biogas production yields a lower energy output than if used for thermal gasification and/or combustion
The successful operation of biogas plants is relatively complex and requires large experience

The utilisation of large quantities of biomass with low DM content (manure) makes transport and sourcing radius a critical parameter
Biogas plants without upgrading facilities often deliver the biogas to a CHP-facility or a gas boiler in the local district heating system. As the demand for heat is seasonally dependent, it requires a flexible operation or storage capacity, and may limit biogas production during summer.
Substantial road transport of biomass

Environment

Biogas can substitute fossil fuels in the energy system and thereby avoid emissions of CO_2. Furthermore, the emission of greenhouse gasses from agriculture can be reduced. Methane is emitted from manure and slurry when it is stored in stables or slurry tanks, and the higher the temperature in the stables or slurry tanks, the faster the emission of methane will happen. In biogas plants, this methane is captured and utilised instead of being released into the atmosphere during manure storage. When the manure is treated at a biogas plant, the emission of methane during storage may be reduced by up to 70% [7].

Methane leakage is an environmental issue related to biogas production. Methane is the second most important GHG contributor to climate change following CO_2. On a 100-year timescale, methane has 28 times greater global warming potential than CO_2 per kg [15]. An investigation from 2021 covering 69 Danish biogas plants showed a weighted average leak of methane of 2.5%, ranging from 1.9 % in agricultural plants to 7.7 % in wastewater treatment plants. [14]. The findings resulted in a new regulation aiming at reducing methane leakage to a minimum via a third-party inspection and enforcement notices from the Danish Energy Agency. The new regulation entered into force January 1, 2023 and it aims to reduce methane leakage from biogas plants as far as possible.

Odour from biogas plants is often mentioned as a concern, but can be avoided with proper filtering of the off-gases, treatment of the air from all parts of the biogas plant and good management during operation. The odour nuisances from field application are reduced when slurry is anaerobically digested compared to the direct application of untreated livestock manure.

Hydrogen sulphide (H_2S) makes up a small part of the produced biogas. H_2S is highly toxic and represents an environmental issue. It is, however, easy to detect as the chemical has a strong odour, and a reduction in odour will therefore also solve the toxicity issue. The content of H_2S in the biogas varies depending on the feedstock. When livestock slurry is the main biomass input, the raw gas typically contains 2,000-8,000 ppm, whereas biogas produced from household waste typically exhibits hydrogen sulphide levels of 600-800 ppm [8].

Multiple methods can be used to remove the H_2S. Common techniques involve using either iron chloride, biological filters, or activated carbon. Iron chloride is dosed into the digester or into the substrate pre-storage tanks when needed. Depending on the substrate the iron chloride needed for the reduction of the hydrogen sulphide levels varies. In biofilters, the off-gases are led through a chamber filled with products with a large surface on which microorganisms that degrade the unwanted substances live. When activated carbon is used, the gas is led through a filter where activated carbon absorbs the H_2S. Over time, the activated carbon will be saturated and has to be re-activated or renewed. The CAPEX of the activated carbon technology is very low; however, it has a high OPEX meaning that it is mostly applied in smaller plants or used as a final polishing of the off-gases from biological filters or in the ramping up of new biogas plants where the biological filters are not fully matured. The cost of H_2S removal using activated carbon is approx. 0.012 Euro per Nm^3 methane [12].

Biogas engines tolerate only small amounts of H_2S in the biogas. Therefore, the H_2S content must be reduced below the acceptable level to meet the specification from the engine suppliers and the environmental legislation. When the biogas is upgraded to biomethane and injected into the gas grid, complete H_2S removal may be necessary, and this is normally an integrated part of the

upgrading process. As most biogas is upgraded, the cost of H_2S removal is not included in the costs for the biogas plant in the datasheet, but instead in the cost of the upgrading plants.

Biomass is a limited resource that often has many uses, i.e., the use of some types of biomasses in biogas plants competes with alternative uses of biomass, e.g. maize and straw. It should be ensured that biomass is used where it adds the most value.

Research and development

The Danish biogas R&D activities focus on several areas to increase energy production, improve the economy of the plants, reduce the climate impact and optimize the value of the digestate as fertilizer.

To increase energy production the focus is especially on developing technologies enabling increased use of "difficult" biomasses with higher methane potential per ton such as straw, which are readily available. A development towards increased use of straw has occurred during the recent years, and this development is expected to continue although it is recognized that there might become increased competition for straw in the future, as straw has many alternative uses.

Biotechnological advances within microbial enzymatic hydrolysis may improve biogas production, in particular from lignocellulolytic material. However, today, the high cost of commercial enzyme production limits its application [16].

To reduce the climate impact and ensure sustainable production of biogas, a significant focus is on developing the operation and technologies of the plants to reduce methane leakages. Gas collection from several tanks in the process is under development, including collecting gas from the pre- and post-storage tanks. This is seen as an important development to reduce methane emissions from leakages.

Further development activities are related to the optimisation of control systems and logistics, for instance, transport systems integrated with larger stable systems, and possibilities for higher DM content in the livestock slurry.

Examples of market standard technology

As mentioned, most of the biogas production in Denmark comes from large plants, where the biogas is upgraded for injection into the gas grid. One of the largest biogas plants today is SBS's Kliplev, from 2022. The plant has a capacity of 48 million Nm^3 of biomethane with an input of approx. 1 mill. tons biomass per year and an HRT of approx. 40 days [6].

Ausumgaard Biogas from 2017 has an HRT of 70-80 days and an annual production of 15 mil. Nm^3 of biomethane [9].

Danish biogas plants generally allow a maximum DM content in the digester of 13-14% with current state-of-the-art technology. Further the HRT typically varies between 60 and 100 days, with most plants lying in the interval of 60-80 days.

Financial data

Prediction of performance and costs

The potential for improving technologies is linked to their level of technological maturity. Biogas plants are assessed to be a category 4 technology, meaning that it is a commercial technology with large deployment.

It is expected that the investment costs will continue to decrease gradually due to learning curve effects, but at a slower pace than previously. The reason for this is that many elements of a

biogas plant consist of mature technologies from other industries, e.g., civil construction works and general process equipment, where learning curve effects are expected to be limited.

The greatest cost reductions are expected to arise from the use of biomasses with a higher methane output per ton input combined with increased professionalization and technical optimization of operations, which are likely to increase efficiency. Further, due to the increasing size of biogas plants, economies of scale are also expected to lead to a reduction in cost.

Learning rates for energy technologies typically vary between 5% and 25%. In 2015, Rubin et al. published "A review of learning rates for electricity supply technologies", which provides a comprehensive and up-to-date overview of learning rates for a range of relevant technologies. 10-15% seems to be the typical level for many technologies, with solar PV being an exception demonstrating learning rates well above 20% [17]. Studies on learning rates for biogas plants are scarce, however, a 2006 study [18] finds a learning rate of 12% for the investment cost of Danish biogas plants based on data from 1988 to 1998. This improvement is however related to higher yield from the plants (i.e., lower investment cost per methane output) due to feedstock changes.

It should be noted that using a learning curve as a method for forecasting price developments is less applicable for biogas plants, than for solar panels and other module technologies. The expectations for cost development applied in the datasheet are therefore further substantiated by the EUDP-report "Production of upgraded biogas - optimising costs and climate impact" [12]. The report has analysed a variety of specific cost-reduction measures for modern biogas plants of different capacities. The report finds that the greatest reduction potential in biogas plants lies within biomass pre-treatment, biogas production, slurry handling and energy integration. Within biomass pre-treatment, re-digestion/selective digestion contributes with approx. 2/3 of the reduction potential and technologies for mechanical shredding account for the remainder. Within biogas production, reduction of downtime contributes by approx. 1/3 of the potential, whereas the remaining improvement potentials concern optimization of electricity and heat consumption, and reduction of methane loss. For slurry handling, reduction of washing water, rapid discharge of pig slurry and mixing of deep litter in the slurry contribute about half of the potential and filter box for separation with the other half. For energy integration, important measures concern the use of heat pumps, heat exchange and regular cleaning of pipes and heat exchangers. Overall, the report identifies cost optimization potentials between 10% and 16%, depending on plant size and configuration.

In the subsequent analyses, a progress rate of 0.9 is assumed, corresponding to a 10 % cost reduction for every doubling of the installed biogas capacity. The cost reduction is applied to investment costs as well as to O&M costs.

Data for 2020

The production of biogas in Denmark has increased from approx. 3 PJ in 2000 to 13 PJ in 2018 and 21.3 PJ in 2020, and the increase has almost exclusively been in agricultural biogas plants. The average size of new plants has been increasing, and so has the biogas production per ton biomass input. This is due to longer HRT, changes in biomass input and technological improvements.

Currently, most new biogas plants are built with a capacity to produce 25-50 million Nm^3 methane and they are built together with a corresponding upgrading facility allowing the plant to produce biomethane, which can be fed into the gas grid. In the datasheets, the model biogas plants represent the expected interval for future large biogas plants assumed to have an annual production of 25-50 million Nm^3 methane.

The cost of producing biogas is correlated with the cost of the feedstock, rather than being uniquely dependent on the efficiency improvements in technology. In larger biogas plants, the cost of feedstock accounts for more than half of the expenses. In the datasheet, the cost of

feedstocks and transport are not included in the OPEX as they are highly dependent on local conditions and may vary widely. It is, however, important to include costs for feedstock when calculating the operational cost of biogas plants – estimated costs for different feedstocks (including transport) based on [12] can be found in Table 3. The cost is however dependent on various factors, e.g., the price for biomass, transportation cost, bulk purchase benefits and the local supply of different kinds of feedstocks. The cost is also depended on the supply-demand balance. In 2022, there has been an increase in feedstock prices, due to changes in the supply and demand. The costs of feedstocks in the table below are estimated for 2020, meaning that the prices do not represent the current situation, which is considered temporary and unrepresentative.

Table 3: Overview of estimated prices for feedstocks including the cost of transport

	Estimated cost of feedstock incl. transport (Euro/ton)
Cattle slurry	3.5
Pig slurry	3.5
Deep litter	8
Manure	8
Energy crops (Grass – sugar beet)	40-80
Straw (timothy grass straw – wheat straw)	35-82
Industrial waste (dairy waste – molasses)	4-113
Household waste	18
Glycerine	182
Other	Depends on input

Source: [12]

The table gives an overview of estimated prices for feedstocks including the cost of transport. The estimates are based on prices in [12], a more in-depth overview can be found in the report. For the categories, where prices are given as an interval, more types of biomasses, with varying prices, are included in the category. The biomass with the lowest and highest costs within those categories is indicated in parentheses.

Assumptions for the period 2020 to 2030

For the period 2020 – 2030 the total biogas production in Denmark is expected to more than double from 21.3 PJ to approx. 50 PJ [2]

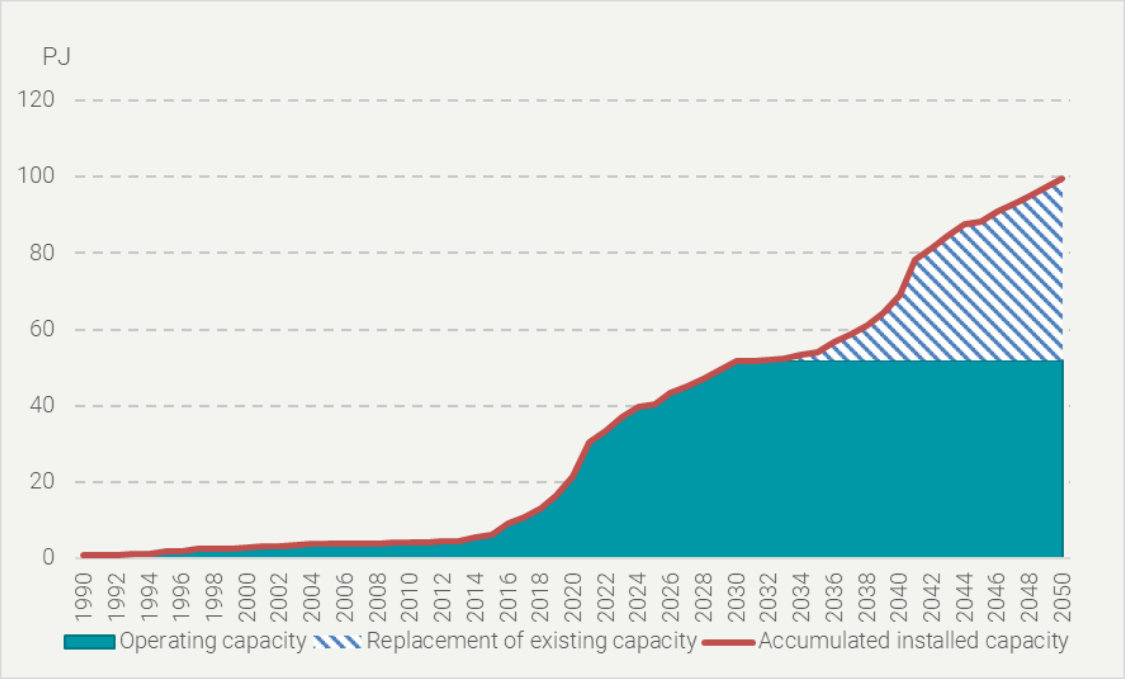
Due to the expected learning curve effect, it is estimated that an increase in installed capacity from 21.3 to 50 PJ and a learning curve progress rate of 0.9 (corresponding to a 10% cost reduction for every doubling of the installed biogas capacity) will lead to a 12% reduction in costs between 2020 and 2030.

Assumptions for the period 2030 to 2050

Forecasting the development of biogas production for the period 2030-2050 is associated with significant uncertainty. Stakeholders have different perceptions of how the deployment will be past 2030.

Denmark's Climate Status and Outlook 2023 assumes that biogas production will flatten out at a level of around 50 PJ and decrease slightly between 2030 and 2035. This projection is based on a frozen policy approach and assumes that existing support schemes will gradually be phased out [2].

Figure 5: Forecast of accumulated installed biogas production capacity in Denmark, based on Denmark's Climate Status and Outlook 2022 and assuming the need for replacement capacity beyond 2030.



In the Danish Energy Agency's Climate Programme 2021 [19] four long-term scenarios are established exploring how Denmark could become climate neutral by 2050 focusing on different mitigation options. Depending on the scenario, biogas production reaches somewhere between 35 and 52 PJ by 2050, i.e., on the same level as 2030 or lower. On this basis, we assume that the installed production capacity remains stable at around 50 PJ after 2030. There will, however, be a need for replacement capacity as existing biogas plants reach their technical lifetime of around 20 years, which leads to additional learning. Several of the existing biogas plants are older than 25-30 years and have within recent years been renewed and enlarged and are hence brought up to a state-of-the-art level. Therefore, by 2050 the accumulated installed (not operating) capacity is estimated to reach around 100 PJ. Under the assumption of a 10% learning rate, a cost reduction of 21% is reached by 2050 compared to 2020.

Uncertainty

The general uncertainty when calculating energy generation costs for biogas plants is high, but the investment costs seem to contribute less than the operating costs. Data from existing biogas plants in Denmark show that the energy production per ton input as well as other cost-determining factors are quite different from plant to plant. As previously mentioned, the methane output per ton of biomass input has also been increasing from 2016 to 2020, which shows that the plants have become more efficient, or are using biomasses with higher biogas potential. Key parameters of the energy output are the composition of the input material and the processing time. The data in this technology catalogue consider a fixed composition of the input and a fixed processing time. The input in the basic plant represents the current distribution of biomass input in Danish biogas plants, with small adjustments due to future limits on the use of energy crops. The HRT is 65 days, which is considered representative of current state of the art plants. Concerning the costs, biomass purchase, transportation, auxiliary energy, and labour costs are

important but may vary widely. In the datasheet, the biomass purchase and transportation costs are not included. The cost for auxiliary energy and labour costs are included.

Datasheets

There is a total of 6 datasheets for biogas plants; three datasheets for a plant with a production capacity of approx. 25 million Nm^3 biomethane, and three for a plant with a production capacity of approx. 50 million Nm^3 biomethane. The three datasheets for each plant size refer to different biomass input scenarios:

Basic plant: Datasheet for a basic plant with an input composition reflecting current biomass use at Danish agricultural biogas plants, but with minor adjustments to reflect future cap on the use of energy crops
Scenario A: Datasheet for a plant without energy crops, but with higher input of straw and slurry compared to the basic plant
Scenario B: Datasheet for a plant with reduced input of: energy crops, industrial waste, household waste and slurry, but increased input of: straw, glycerine and other crops and residues, compared to the basic plant.

The capacities of the biogas plants in the data sheets are stated in MW output to ensure comparability between the different technologies in the catalogue. The capacities of the plants are 29.5 MW and 59 MW (lower heating value), which corresponds to approx. 3,000 and 6,000 $Nm^3\,CH_4$ per hour. Using MW as the unit is however not common in the industry, therefore, an additional row has been created showing the output capacity in Nm^3 methane[2].

The capacity of a biogas plant is commonly stated as tons per year and for most of the input material, a calorific value is not relevant (e.g., manure). Thus, for comparability with the template used throughout the catalogue, the energy of the biogas output is assumed to be 100%. The stated auxiliary energy consumption is stated in percentages of the output energy.

In the data sheets methane emissions from the biogas plant are set to 0.9%, based on the expectation that total emissions from biogas and upgrading at new plants will be max. 1% and that emissions from upgrading using amine scrubbing technology is approx. 0.1%.

The HRT of the plants in the data sheets are set to 65 days, which is considered to be representative of future large plants.

Methane output per ton input of a specific biomass varies depending on e.g., retention time and the specific properties of the biomass such as dry matter content. The methane outputs per ton biomass input used in the data sheet calculations are based on the estimated average methane yields currently realised in Danish biogas plants. Acknowledging that the average HRT of existing plants is likely to be shorter than the anticipated average HRT of new plants, it is recognized that the used yields may in fact somewhat underestimate the methane yield per ton biomass input, particularly for biomasses with longer degradation time, e.g., deep litter, energy crops (grass) and straw. However, lack of consistent data on the correlation between HRT and methane yield for different biomasses, implies that the used data is considered to represent the best available data. It is furthermore noted, that most of the used biomass categories includes a range of different biomasses, and for some categories – particularly industrial waste – the biomasses may have quite different properties in terms of methane yield. Despite such differences biomasses have nevertheless been grouped in order to reduce complexity and give a meaningful picture of the

[2] The conversion is based on a calorific value for methane of 9.97 kWh/Nm^3 (equivalent to 35.9 MJ/Nm^3).

biogas sector as a whole. The biomass categories used in the data sheet calculations, including their assumed DM content and methane output per ton, are listed in Table 4.

Table 4: The biomasses used in the biogas plants, their DM content, and methane output/ton

	DM content (%)	Methane output per ton biomass input (GJ CH_4/tons)
Cattle slurry	7.6	0.5
Pig slurry	5.0	0.4
Deep litter	30.0	2.2
Energy crops	32.1	3.6
Straw (and similar)	84.3	7.5
Industrial waste	29.9	3.5
Glycerine (and similar)	75.0	12.7
Household waste	18.0	2.7
Other crops and residues	30.0	3.3

In the datasheet for the basic biogas plant, the biogas plant presented is based on data and expectations for biogas plants established from 2020 towards 2050. The basic biogas plant is based on the biomass input presented in Table 5, which reflects the current input composition (see table 2) adjusted for the 4% cap on energy crops, which will be operational from 2024. The average DM content of the biomass input for the basic plant is 16% and the methane yield is 1.59 GJ/ton input.

Table 5: Basic biogas plant scenario

	Input share (%)
Cattle slurry	40.5
Pig slurry	25.5
Deep litter	8.5
Energy crops	4.0
Straw (and similar)	2.0
Industrial waste	13.0

Glycerine (and similar)	1.5
Household waste	3.5
Other crops and residues	1.5
Total	100

There are significant uncertainties regarding the future feedstock input for biogas production just as there are significant variations across individual plants. To address these uncertainties and variations, two additional datasheets for each of the two plant sizes are included: Scenario A and Scenario B.

It is noted that the cost calculations for the different biogas plants consider that digester capacity and pretreatment requirements depend on the composition of the feedstock. Costs for the reactor are reduced for biomasses with higher gas output, while costs for pretreatment increase when the proportion of biomass with a demand for pretreatment (e.g., straw) increases. The lifetime is shorter for e.g. pretreatment plants than for the biogas plant itself; this is also considered in the calculations.

In Scenario A, the share of energy crops is reduced to zero, while the share of straw is increased markedly and the shares of cattle and pig slurry are increased slightly. Scenario A builds on the expectation that many new plants are assumed to avoid reliance on energy crops due to political ambitions to reduce use of energy crops further than the 4% cap that applies from 2024. Accordingly, further reductions in the use of energy crops below 4% are considered to represent a relevant scenario. It is furthermore assumed that technological development implies that onwards it will be possible to increase the use of straw, possibly in combination with other biomasses such as catch crops or grass. In practice, the options for increasing the share of straw depend on the design of the plant and the pre-treatment of the feedstocks. Moreover, costs are assumed to increase with increased use of straw. Today there are plants where straw makes up more than 10% of the feedstock input [9, 20].

The cost of production in the datasheet excludes the price for the biomass input, which generally constitute one of the single most important components of production costs. However, pre-treatment of the biomass is included in the costs. It is assumed that the plant uses grinding and milling as mechanical pre-treatment methods for e.g., straw.

The average DM content of the biomass input in Scenario A is 17.3% and the methane yield is 1.67 GJ/ton input.

Table 6: Scenario A

	Input share (%)
Cattle slurry	41.0
Pig slurry	26.0
Deep litter	8.5

Energy crops	0
Straw (and similar)	5.0
Industrial waste	13.0
Glycerine and similar	1.5
Household waste	3.5
Other crops and residues	1.5
Total	100

In Scenario B, the input shares of energy crops, industrial waste and household waste are reduced markedly, while the input shares of cattle and pig slurry are reduced slightly. The reduction in industrial and household waste reflects the uncertainties regarding the extent to which plants continuously will be able to find new and economically attractive waste feedstocks, e.g., by importing waste from other countries. The reduction in of energy crops reflects the uncertainties related to the future regulation on energy crops, e.g., further reductions of the 4% cap on energy crop use in 2024. The reductions are countered by increases in the input of straw, glycerine and other crops and residues, where the latter to some extent is expected to be able to ease the introduction of an increased amount of straw, e.g. by co-ensiling. The average DM content of the biomass input in Scenario B is 21% and the methane yield is 2 GJ/ton input.

Table 7: Scenario B

	Input share (%)
Cattle slurry	40.0
Pig slurry	25.0
Deep litter	8.5
Energy crops	2.0
Straw (and similar)	10.0
Industrial waste	7.5
Glycerine (and similar)	2.0
Household waste	1.0
Other crops and residues	4.0
Total	100

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83 Gasifier Plants (for Producer Gas)

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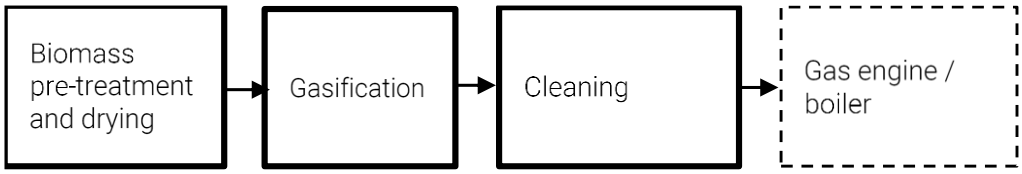
Publication date
March 2017

Qualitative description

Brief technology description
Fixed bed gasifiers are smaller scale plants (<10 MW output) with direct gasification processes that can be either updraft or downdraft, and that can be staged into different process steps.

The primary use of the gas will be in co-generation of heat and power (CHP), or in heat-only boilers. In this catalogue the device for conversion of the producer gas is not included.

For the fixed bed technologies, it is assumed that atmospheric air is used as gasifying agent in direct gasification. Thus, the gas will contain nitrogen. The nitrogen content and the limited possibilities for upscaling make the fixed bed technologies less interesting for larger plants with further upgrading to natural gas quality or production of liquid biofuels based on syngas.



The updraft (or counter current) gasifier has been used for the last 75-100 years with fossil fuel for electricity, heat, steam and industrial processes such as burning of ceramics, glass making, drying and town gas.

It is characterized by the biomass feedstock and the gas having opposite flow directions. The biomass is converted through several stages. Up to 100°C the water is vaporized. By pyrolysis (extra heating and limited addition of oxygen) the dry fuel is converted to a tarry gas and a coke residue. Subsequently, the coke residue is gasified at 800-1,200°C, while water vapour and/or oxygen (air) is added. The gas has low temperature (~75°C) but a large content of tar, typically 30-100g/Nm^3. Depending on the process, the tar shall either be incinerated or cracked before it is cleaned of particles etc.

Producer gas primarily consists of the components N_2, H_2, CO, CO_2, CH_4, and water. The use of atmospheric air and direct gasification limits the calorific values of the gas to about 6 MJ/Nm^3 for the dry-cleaned gas from an updraft gasifier [8]. For internal combustion engine applications, gas from updraft gasifiers needs tar removal and possible effluents from the cleaning step need to be handled.

Figure 1: Updraft gasifier, principle

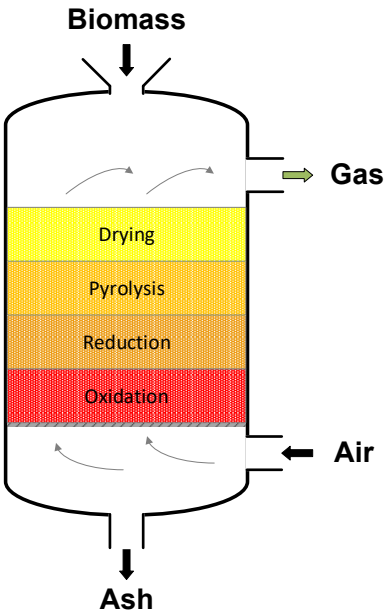
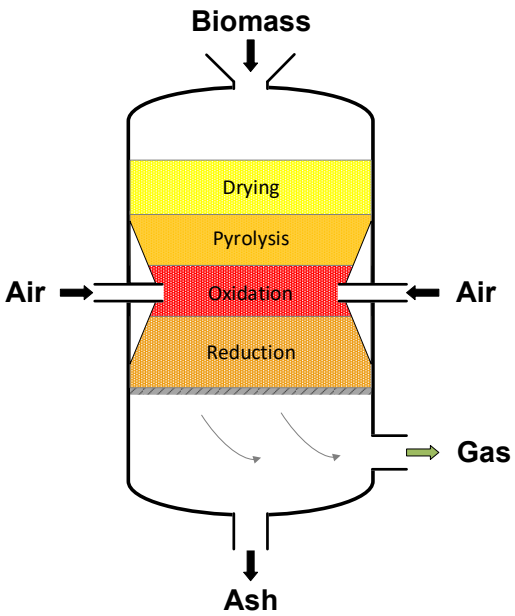


Figure 6: Downdraft gasifier, principle



The downdraft (or co-current) gasifier has the same flow direction of the biomass feedstock and the gas. The biomass is converted through several stages. Up to 100°C the water is vaporized. By pyrolysis the dry fuel is converted to a tarry gas and a char residue. Subsequently, the char residue is gasified at 800-1,200°C, while water vapour and/or oxygen (air) is added. By adding air to the char zone, the tar content in the producer gas is reduced and amongst fixed bed gasifiers the downdraft type produce gas with the lowest level of tar.

In staged downdraft gasification, pyrolysis and gasification are separated in two reactors, enabling a partial oxidisation of tar products between the stages. Thus, staged gasifiers are producing a gas with low tar content, which is essential for engine operation. The tar content is often below 100 mg/Nm^3 and can be below 10 mg/Nm^3.

The pyrolysis process can be driven by either internal or external heating. Internal heating is performed by addition of air/oxygen consuming a part of the energy content in the fuel, while external heating utilises waste heat from the produced gas and from the engine to dry and pyrolyse the fuel.

The data in the table are valid for external heating, as this results in higher efficiencies.

Producer gas primarily consists of the components N_2, H_2, CO, CO_2, CH_4, and water. The use of atmospheric air and direct gasification limits the calorific values of the gas to 4.5-6 MJ/Nm^3 for the dry, cleaned gas from a downdraft gasifier [8].

For internal combustion engine applications, producer gas from downdraft gasifiers may need only cooling and dust removal.

Input

Solid biomass such as wood chips, pellets, chunks and briquettes, industrial wood residues, demolition wood and energy crops can be used. Auxiliary electricity for process machinery.

Requirements to moisture content and size of the fuel depends on the design of the reactor and the process: Updraft gasifiers can take fuels with up to 50% water content, whereas downdraft gasifiers require fuel with a maximum of 15-20% water. In practice, artificial drying is often

integrated with the gasification plant to ensure a feedstock of constant moisture content [8]. Downdraft gasifiers typically need homogeneous sized biomass input to avoid packing of bed and subsequent pressure loss across the fuel bed.

Output

Producer gas suitable for combustion in gas engines, gas turbines or boilers.
Recoverable heat for domestic heating.
Ash, slag and possibly tar and/or effluents from cleaning step.

The range of composition of the producer gas is rather broad according to technology, fuel, operational conditions etc. Levels from two concepts appear from the table below [8].

Table 1: Ranges of composition of producer gas from fixed bed gasifiers.

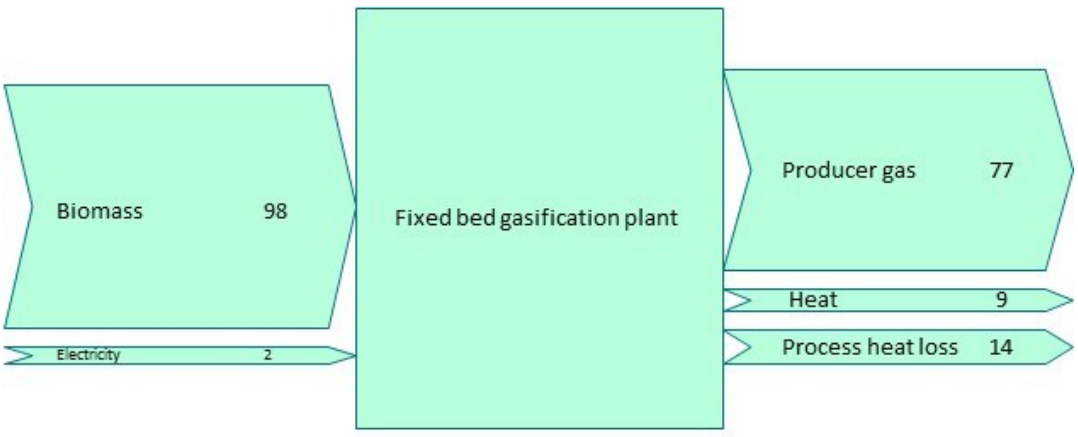
Component	vol%
H_2	19 - 31
CO	18 - 23
CO_2	12 - 15
CH_4	1 - 5

Energy balance

Updraft gasifier: Based on an energy input of wet biomass (100%), a producer gas energy output of 40-65%, and a heat output of 10-20% can be obtained [8].

Staged downdraft gasifier: Based on an energy input of wet biomass (100%), a producer gas energy output of 75-85%, and a heat output of 10-20% can be obtained [8].

Figure 2: Sankey diagram of fixed bed gasifier in 2030.



In many cases, a fixed bed gasifier will be part of a CHP system with an ICE genset that provides electricity also to cover the internal electricity demand. In this case, where the gasifier is standing alone and the system output is product gas and heat, an electricity input is needed.

The heat loss may in many cases be lowered by condensation of the producer gas and circulation of the heat to drive the gasification process.

Typical capacities

Updraft gasifier: 0.08 – 10 MWe (0.2-25 MJ/s fuel)

Downdraft/staged downdraft gasifier: 0.04 – 2 MWe (0.15-5 MJ/s fuel)

Capacities above these levels are typically increased by parallel installation of units. [8], [2].

Regulation ability

Gasifier output can be regulated within few seconds for downdraft gasifiers, and within minutes for updraft gasifiers. Start-up time from cold condition depends on plant sizes and design, in any case several hours to days. Minimum loads of 10-20% can be obtained for updraft-, and 25-30% for downdraft gasifiers [6]. Gasifiers are typically to be kept in continuous operation.

Space requirement

The main space requirements typically relate to the storage and handling of biomass feedstock, which can be assumed to correspond to biomass boilers.

Advantages/disadvantages

Compared with other gasification technologies, fixed bed gasifiers - and especially the downdraft types - provide a simple way of generating a gas clean enough to be used in an internal combustion engine for CHP. However, they generally have limited possibilities for upscaling, especially the downdraft types, as maintenance of a stable bed becomes increasingly challenging in larger cross sections. This is the reason behind parallel installation of units to increase capacity of a site. Furthermore, air as gasification media makes the gas unsuitable for methanation.

The updraft gasifier has limited requirements to fuel quality, i.e. the contents of moisture and ash. Furthermore, the gasifier can ramp up and down thereby offering flexibility both electricity generation and for supplying heat to district heating grids.

The downdraft gasifiers can also be tailored to a large variety of fuel qualities and capacity demands, and generally produces less tar.

Gasification of biomass for use in decentralized combined heat and power production can decrease the emission level compared to power production with direct combustion and a steam cycle.

Compared with alternative small-scale biomass-based electricity generation technologies, the gasifier / engine plants can reach higher net electrical efficiencies, typically up to 30% in CHP mode [2]. Existing natural gas fuelled engines can be converted to run solely on producer gas, or on a combination of producer gas and natural gas. When a spark ignition engine is converted to operation on producer gas its energy input capacity is derated to about 40-50% due to the lower calorific value of the gas [7]. One disadvantage compared to a natural gas-powered engine is the long start-up time of the gasifier (from cold). Also, excessive soot-formation may occur at start/stop.

Environment

Emissions from generation of biomass gases are very limited. Emissions from utilisation of gases from gasifiers may occur at each process step:

- gaseous emissions (exhaust gas, possible leakages)
- liquid emissions (scrubbing water, scrubbing wastes, condensates, bio-oil)
- solid emissions (ash, dust)

Generally, the environmental aspects of biomass gasification are comparable to those of biomass combustion processes; however, as the producer gas from fixed bed gasifiers is filtered

thoroughly before it is fed into the IC-engine, the standard emissions are CO, NO_x and UHC. From a stable operation of a demonstration plant utilising a two-stage gasifier at DTU, the below emissions have been measured [10]:

Table 2: Example of emissions from a plant with a two stage down draft gasifier.

CO (mg/Nm^3 at 5% O_2)	970.0
NO_x (mg/Nm^3 at 5% O_2)	1197.0
UHC (mg/Nm^3 at 5% O_2)	21.4

Source: [10]

This performance does not comply with the current emission regulations in Denmark. A possible commercial plant would apply primary or secondary emission reducing measures to comply with regulations.

Dependent on technology, trace metals, especially cadmium contained in the biomass, may be entrained with the gas or end up in the ash from the biomass gasifier. Further, the ash may contain polycyclic aromatic hydrocarbons (PAHs). Therefore, spreading of ash in forests or on agricultural land must be carried out with considerable caution. It has been demonstrated that in some cases thermal gasification may as a side effect entail the possibility to extract trace metals. In Denmark utilisation of the ash is regulated by a ministerial order for biomass ash.

No emission data is stated in the data sheets below, as the specific utilisation of the producer gas is not covered by this technology data sheet.

Research and development perspectives

Updraft

Up-draft gasification technology with CHP has been demonstrated over a long time in Denmark and abroad.

R&D is carried out, aiming at solving operational problems such as corrosion, process regulation etc. The main issues to be addressed include:

Ability to handle a wider range of fuel properties, in particular waste wood and other biomass residues
Establishing references of up-draft gasification plants for waste wood and other biomass residues to drive the incremental development
Establishing updraft demonstration plants with oxygen and steam as gasification agent to be able to produce methane.

Other issues that should be addressed to support small-scale biomass gasification:

Purification of wastewater containing tar; in particular capital cost reduction
Meeting emissions regulations
Reactor calculations; kinetic models of significance for design and control

Downdraft

There exist a number of suppliers of smaller down draft gasifier plants for CHP, ranging from 10 kW_e to 2 MW_e, and as such the technology seems to have reached a level where it enters technological maturity [15]

Research and development activities seem to focus on incremental operation and design optimisations, including better process regulation and automation for unmanned operation, scaling up, and improving gas engine operation with gasification gas.

Examples of market standard technology

Updraft

At Harboøre Fjernvarme a 3.6 MJ/s updraft counter-current moving bed gasifier was installed in 1994. The gasifier is used for CHP production and has a gross electrical output of 1.0 MW. The gasifier is fuelled by wet forest woodchips. The gasifier is supplied and operated by Babcock & Wilcox Vølund A/S. [2]

Downdraft

Biosynergi Proces had installed a 300 kW_e and 750 kJ/s heat CHP demonstration plant at Hillerød district heating company. The plant came online in 2016. The concept is designed supply a clean gas on basis of wet forest wood chips that are dried on site as an integral part of the process. Output heat is used for district heating. The process is an "Open Core" downdraft and is a successor and upscaling of the Græsted pilot project (450 kJ/s fuel). [2], [5]. The plant has been dismantled by the end of 2017 due to lack of financing to solve minor technical start-up problems.

In Innsbruck, Austria, SynCraft has installed a 260 kWe and 600 kJ/s heat CHP plant at the municipal water treatment company, IKW. The plant is a staged downdraft type with an innovative floating fixed bed char gasifier vessel and came online in 2017. The plant used wet wood chips that is dried on site. Output heat is used for district heating.

A number of suppliers and projects outside Denmark are mentioned in [2] and in [30].

Prediction of performance and costs

Small scale gasification plants for CHP production based on biomass are offered by many suppliers worldwide on a commercial basis [2]. However, commercial deployment is for larger plants still moderate and the technology can be characterized as being in a transition between demonstration and commercial maturity (Category 3).

Further development potentials exist, for example for using new fuels types, technical optimizations, upscaling and better control of un-manned installations. Many suppliers tailor their equipment to certain fuels and needs and offer turnkey solutions. A larger commercial deployment may lead to incremental price reductions [2].

The projection of investment cost assumes that the accumulated production capacity will increase by 40 % between 2015 and 2020, double between 2020 and 2030 and further double between 2030 and 2050. Applying a typical learning curve progress rate of 90 % this yields a 5 % decrease in investment costs between 2015 and 2020, a further 10 % reduction between 2020 and 2030 and additional 10 % reduction between 2030 and 2050. It should be stressed that this projection is associated with considerable level of uncertainty. The statistical data on existing plants is very limited, impairing more detailed analyses. O&M costs are assumed to follow the same trend as investments costs.

Due to the limited possibilities for upscaling it is not expected that applying fixed bed gasifiers to production of methane or other synthetic fuels will be commercially interesting. This would require small to medium scale oxygen production and methanation to reach commercial level. In that case, small to medium scale gasification combined with biogas production for methane production could become an attractive solution.

Uncertainty

Even though several plants have been in successful operation for several years the uncertainty regarding price and performance for future developments remains considerable. The data assumes considerable learning curve effects. However, there is a widespread number of different principles and variants of the technology, of which many are pioneer projects, and it is not clear which improvements can be realized, and how far.

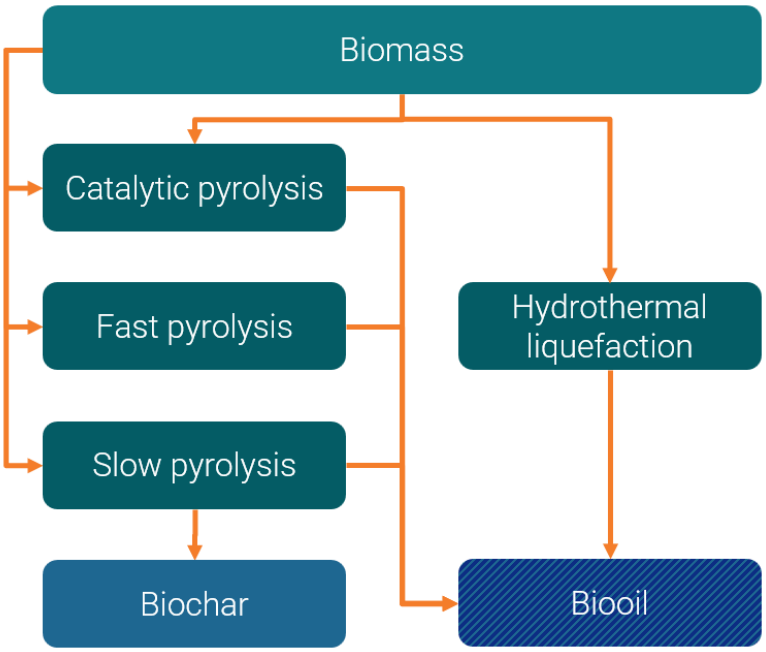
Additional remarks

Today, fixed bed gasifiers are usually integrated with an internal combustion engine gen-set. Besides the described fixed bed gasifiers, a number of suppliers offer CHP technologies based on bubbling fluid bed gasifiers in the 1-2 MW_e range, e.g. the Spanish Eqtec. [2].

References

Please refer to chapter 99 Jet Fuel from Biomass Gasification for references.

Pyrolysis and Hydrothermal Liquefaction



This section focuses on the decoposition of biomass at high temperatures in pyrolysis process or at high temperature and pressure in the process of hydrothermal liquefaction. The resulting outputs are liquid and solid products. Biooil is the primary product of Fast and Catalytic Pyrolysis, Hydrothermal Liquefaction and some types of Slow Pyrolysis. Biochar is the primary output of Slow Pyrolysis.

Biochar can be used directly for carbon sequestration and as fertilizer, while biooil can be burned in power and heat production or upgraded to drop in fuels. All pyrolysis pathways require dry biomass of higher or lower quality, while Hydrothermal Liquefaction can also handle wet biomass types.

The section includes the following chapters:

85 Pyrolysis Overview

86 Fast and Catalytic Pyrolysis

87 Slow Pyrolysis

89 Hydrothermal Liquefaction

85 Pyrolysis Overview

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Qualitative description

This chapter provides a general overview of pyrolysis as a process, including key principles, reactor types, and its relevance for climate mitigation through negative emissions. The purpose of this overarching pyrolysis chapter is twofold; first, to provide a general introduction to process types as well as relationships between process temperature, duration, and yield. Secondly, it serves as a prospective technology list of pyrolysis technologies. This includes a status for upcoming technologies, listing advantages and disadvantages, though without going into economic detail. The subsequent chapters on pyrolysis in this technology catalogue focus on more established technologies, namely fast and catalytic pyrolysis and slow pyrolysis.

Brief technology description

Pyrolysis is the thermal decomposition of biomass occurring in the absence of oxygen. It is the fundamental chemical reaction that is the precursor of both the combustion and gasification processes. The products of biomass pyrolysis include biochar and gases, including methane, hydrogen, carbon monoxide, and carbon dioxide. The condensable share of gasses can be condensed into pyrolysis oil. While biochar serves as a stable form of carbon storage, the gas and oil fractions can be combusted to produce heat and electricity. Additionally, ongoing research is exploring the potential for upgrading of pyrolysis oil into advanced liquid fuels, which holds particular promise for maritime transport due to the need for energy-dense and sustainable fuel alternatives.

Pyrolysis processes can be categorised as slow, fast or flash pyrolysis, differentiated by the heating rate and residence time. Slow pyrolysis completion time typically ranges from several minutes to multiple hours and results in biochar as the main product. Fast pyrolysis completes the conversion in seconds and is the main industrial route when the primary product is bio-oil. Reported oil yields for fast pyrolysis can reach up to 60% pyrolysis oil under optimal conditions with woody feedstock. Flash pyrolysis is performed at very high temperatures with residence times of less than a second.

The technologies involved for slow pyrolysis, fast pyrolysis, and gasification share many common features, but also have differences that define them. Fast pyrolysis is the most efficient method to produce bio-oil, while gasification is the most efficient method to produce syngas, and slow pyrolysis is the most efficient method to produce biochar. Biochar can be used as a carbon dioxide removal (CDR) technology and has therefore been identified as a potential tool for Denmark in terms of meeting its climate targets.

There are many different types of pyrolysis reactors and processes, which have varying attributes. Pneumatic bed reactors, such as entrained flow and circulating fluidized bed reactors, are suitable for fast pyrolysis of pre-treated biomass. Gravity reactors have a simple construction and are primarily used at lab-scale. Stationary bed reactors are used for both slow and fast pyrolysis. Within this category, fixed bed reactors have a small capacity and require manual loading and unloading, making them suitable for lab-scale. Semi-batch reactors allow partial feeding or removal of materials during operation, providing better control than batch systems and are commonly used at lab- and pilot-scale. For larger-scale slow pyrolysis, semi-continuous or

continuous configurations, such as auger or rotary kiln systems with continuous biomass feeding and continuous char removal are typically used. These configurations operate with constant feed and product removal, thus enabling steady-state operation. Mechanical reactors include mechanisms to facilitate contact between biomass and the heating agent, whereas other reactor types rely on non-mechanical mechanisms such as gas flow, gravity, or static heat transfer [4].

Figure 1: Simplified flowsheet for the pyrolysis process main process stages following the biomass from input to end products. Note that the pretreatment steps, dewatering and drying, will be needed if the biomass is wet, e.g. for digestate from biogas production. Electricity will be needed for the operation of machinery used in the process.

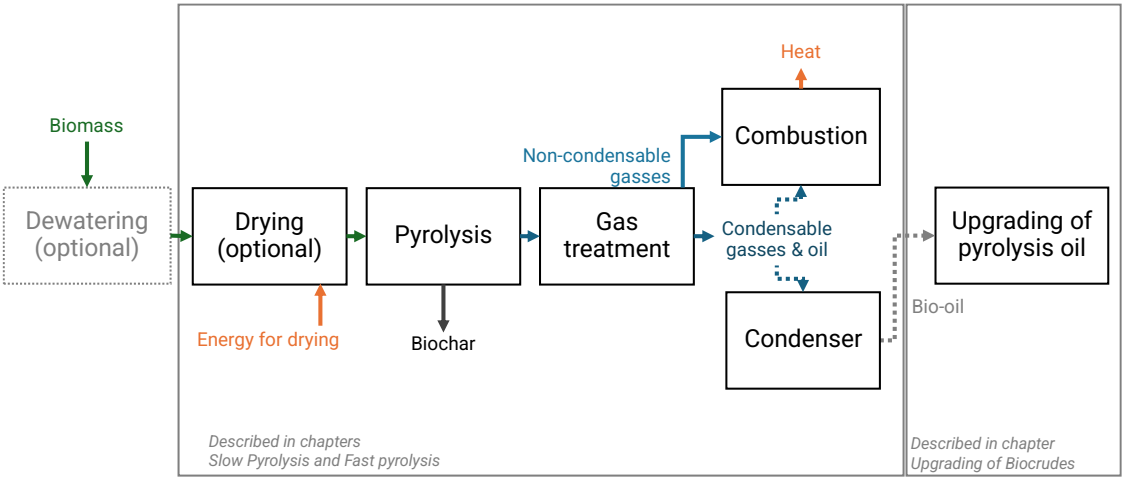
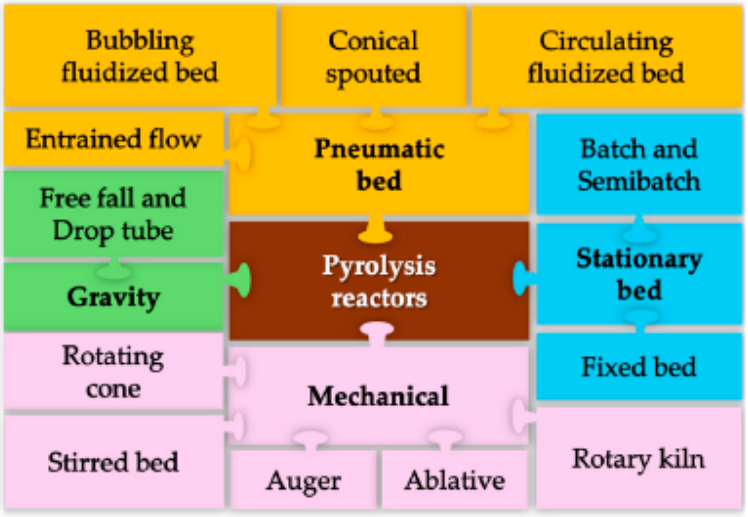


Figure 2: Overview of types of pyrolysis reactors from Jerzak et al., 2024 [4].



Inputs

The pyrolysis process can utilise a wide range of carbon-based feedstocks. However, the feedstock must often be pretreated to ensure that the size distribution and moisture content fit the process. The demands vary depending on the selected type of process and supplier of the pyrolysis technology. While fossil-based or mixed input (such as plastics and tires) can be used, the chapters on pyrolysis in this technology catalogue are limited to use of biogenic inputs.

Biomass is a mixture of hemicellulose, cellulose, lignin and minor amounts of other organics, which each pyrolyze or degrade at different rates and by different mechanisms and pathways. Lignin decomposes over a wider temperature range compared to cellulose and hemicellulose, which rapidly degrade over narrower temperature ranges. The rate and extent of decomposition

of each of these components depends on the process parameters of the pyrolysis reactor's temperature, biomass heating rate, and pressure. The degree of secondary reaction (and hence the product yields) of the gas/vapour products depends on the time-temperature history to which they are subjected before collection, which includes the influence of the reactor configuration.

The requirements for biomass input in pyrolysis systems vary significantly depending on the chosen technology. Fast pyrolysis, for example, is more sensitive to feedstock properties than slow pyrolysis and typically demands a lower moisture content and a more homogeneous feedstock.

To meet these requirements, various pre-treatment steps may be applied, such as mechanical dewatering, drying, and pelletising. Lower moisture content generally improves the efficiency of the pyrolysis process. However, reducing moisture content requires energy-intensive drying, which can add significant cost. The pre-treatment steps help reduce moisture content to a range typically between 10% and 25% and improve feedstock consistency. The specific combination of pre-treatment methods depends on the feedstock composition and the technical specifications set by the pyrolysis technology provider. An overview of the composition and energy content of selected feedstocks is provided in Table 1.

Table 1: Data for various raw materials for pyrolysis. Typical moisture content after drying is 10 to 20 w/w %.
*In this context ash refers to the non-combustible mineral matter in the feedstock. While it does not contribute chemical energy to the process, it adds sensible heat load and can influence reactions and operability.
**The data varies depending on digestate origin and type of process used for biogas production.
***The figures vary a lot depending on origin.

Raw material	Moisture content, before treatment Weight %	Ash content* Weight % on dry substance	Lower heating value H_{eff} (dry/ash free) MJ/kg	Reference
Straw	10 – 15	4 – 10	17 – 18	[2]
Digestate**	55 – 75	15 – 22	16 – 17	[1]
Wood	8 – 60	4 – 6	19 – 20	[2]
Manure***	55	7	18 – 19	[3], [4]

Digestate from biogas production as feedstock must typically be dewatered and then dried to ensure that the moisture content reaches an acceptable level for the pyrolysis plant. The digestate can be reduced to 10% moisture content by using a screw press or a decanter centrifuge, followed by a drying step, and finally a pellet press [1]. As drying is often integrated into the pyrolysis plant itself, this catalogue includes drying as part of the data on pyrolysis systems. Mechanical dewatering, on the other hand, is not included, as it is considered a separate pre-treatment step that must be performed prior to feeding the material into the pyrolysis system.

The difference between manure and digestate from manure is that some carbon has already been converted in the biogas process and is not available in the pyrolysis process. This leads to a comparably higher ash-content and lower biochar-output of the digestate after pyrolysis.

Woody biomass is the easiest feedstock to process in pyrolysis systems due to its relatively uniform structure and predictable behaviour during thermal conversion. The price of woody biomass is however often relatively high due to competing uses, such as energy production,

material recycling, and industrial applications. [5] Woody biomass is particularly relevant for fast pyrolysis, and its use is described in more detail in the chapter on fast pyrolysis.

In a Danish context, pyrolysis of garden waste has been assessed and discussed as a tool reduce climate impacts [5] Garden wood waste is typically available locally in smaller quantities and is therefore suitable for small pyrolysis plants. Woody biomass typically requires pre-treatment, such as chipping/crushing and possibly drying. Mixing wood into the digestate will increase the carbon content and decrease the ash content, resulting in a biochar that stores more carbon. Garden waste, wood, woody waste material, and similar, can be used for this purpose. Note that treated wood, e.g. impregnated or painted, should not be used in case the biochar is intended for agricultural use, as contaminants from the paint can pollute the farmland. In this technology catalogue, examples of small-scale pyrolysis plants suitable for inputs such as garden waste are provided in chapter 87 Slow Pyrolysis..

In addition to pretreatment and the feedstock itself, heat is essential for the pyrolysis process; this can be supplied either internally or externally. Many systems generate heat by combusting a portion of the pyrolysis gas produced during the process (see section regarding pyrolysis gas below). There are also technologies that use electricity as the heat source, which in certain cases can be favourable. If the electricity used is produced from renewable energy, it could be a sustainable way to increase product yields. In such systems, electricity could replace the use of generated pyrolysis gas to run the process. In addition, electricity is needed for the plant operations such as conveyers, pumps, and other equipment.

Outputs

Two main output streams are produced in the pyrolysis process: biochar and pyrolysis gas. The raw pyrolysis gas from the reactor is a mixture of condensable and non-condensable gases. This gas can be cooled, and the condensable part of the gas taken out as pyrolysis oil. The biochar is removed continuously from the pyrolysis reactor by means of a hopper or similar equipment. Slow pyrolysis tends to yield a higher proportion of biochar owing to the slow heating rates and longer residence time compared to fast pyrolysis.

The distribution of products depends on many factors besides the temperature, including, but not limited to, the type of feedstock, the type of reactor, and the residence time in the reactor.

Biochar

Biochar is a solid carbon-rich output from pyrolysis. Due to a high share of stable recalcitrant carbon in biochar, the IPCC has recognised biochar as a carbon dioxide removal (CDR) method. A basis for including carbon storage in biochar applied to mineral soils in national greenhouse gas accounts has been issued [6], but the methodology is not yet finalised. A central topic in the ongoing methodological discussions is how to characterise biochar permanence, as this is key in issuing carbon credits [7].

Biochar will contain nearly all the ash fed into the process with the feedstock, i.e. the higher amount of carbon taken out from the process as pyrolysis oil and gases, the higher the ash content and lower the carbon content will be in the biochar. Nearly all phosphorus included in the feedstock will be found in the biochar, also oxygen and hydrogen will be present in certain amounts. The exact content distribution primarily depends on the raw material [8]. Trials show that feedstock content of cadmium, mercury, polycyclic aromatic hydrocarbons (PAH), polychlorinated biphenyl (PCB), and dioxins is not directly transferred to the biochar. (Paulsson M. , 2020) In Denmark, the Danish government’s pyrolysis strategy includes work to be carried out by the Danish Environmental Protection Agency setting out a clear and simple legislative framework for the use of biochar based on initiated research trials [10].

Furthermore, in Denmark, research initiatives by GEUS and DCE are underway to investigate long-term carbon storage in biochar. The findings will be essential for the development and scaling of

pyrolysis technologies, particularly for slow pyrolysis, where biochar's role as a long-term carbon storage solution is considered one of the main drivers. Since the climate benefit of biochar relies heavily on the stability and permanence of carbon in biochar applied to soil, the lack of robust data on its long-term effects and interactions with soil ecosystems remains one of the most critical barriers to broader deployment.

While biochar has long been considered a soil-improving agent, its actual effects on agricultural soils remain uncertain. Current studies are investigating both the stability of biochar in soil over time and its impact on soil health, including potential effects on microbial life. Initial findings suggest that biochar may have limited or no significant effect on crop productivity in some cases, but more data is needed. In Denmark, the Danish government's pyrolysis strategy includes work to be carried out by the Danish Agricultural Agency through multi-year field trials to investigate the scale at which biochar can be used in agriculture without unacceptable risks to the environment, nature and agronomic conditions such as soil biology, nutrient availability and impact on yields. This also includes studies of biochar's fertilizer value.

Certified biochar has a higher market value compared to non-certified. The European Biochar Certificate Initiative (EBC) certifies biochar in Europe and has established a list of certification requirements for biochar. Several different classes are described in the EBC guidelines [12] and biochar can be certified under different classes. Each class has specific requirements regarding feedstock, production conditions, and chemical properties. For example, biochar intended for agricultural use must meet strict limits for heavy metals and polycyclic aromatic hydrocarbons (PAHs), and must be produced from clean, untreated biomass. The H/C_org ratio must be below 0.7, and the organic carbon content must be declared. The European Commission is preparing a similar methodology named CRCF, though it is not yet finalised or published.

The market for biochar is growing and new process and equipment suppliers, as well as retailers for products, are continuously entering the market. In Germany, where the market is well developed, the market value still depends heavily on local production and demand. Skånefrö in southern Sweden was marketing biochar sold in big bags for 540 Euro/m³ in September of 2025 [14]. However, this price is based on small-scale operations targeted at private customers and should not be considered representative of the general market value of biochar, as it likely reflects limited volumes. Biochar producers can also earn income from the voluntary carbon removal market by selling carbon credits to large companies such as Microsoft [12]. Voluntary carbon credits can be issued when biochar is applied to soils or blended in cement, concrete or asphalt whereby permanent carbon storage is achieved [7].

Biochar can also be used for other applications, e.g. as addition to animal fodder to improve animal health and increase growth rates, as filter material, as replacement of coke in the steel industry, as filling material in concrete, as feedstock for the production of activated carbon and carbon black [14].

Pyrolysis gas

The output pyrolysis gas is often combusted to provide heat to the main pyrolysis process. The gas (either the complete raw gas or the remaining gas after condensation to pyrolysis oil) is combusted and the released heat is partly used in the pyrolysis reactor and partly for other purposes such as drying. The gas or excess heat can also be in used other in processes with thermal demand, for example, in co-location with a biogas plant, where heat integration may improve overall system efficiency, or be used for district heating.

In industrial-size pyrolysis plants, a small fraction of the pyrolysis gas may be used for process heating, unless electric heating frees up the entire amount of pyrolysis gas for external use.

On a dry basis, the main components of the gas are: CO_2 (35 - 50%), CO (15 - 30%), CH_4 (8 - 15%), H_2 (4 - 14%), N_2 (0 - 12%). In addition, the gas consists of a variety of hydrocarbons and other

gaseous components. The composition is highly dependent on whether an inert sweep gas is used and whether there is air ingress. Other factors having an impact are pyrolysis temperature, type of feedstock and type of reactor. Note that this gas composition differs significantly from natural gas, which mainly consist of methane, ethane and other components with a higher heating value than the components in pyrolysis gas.

In case a surplus of gas is produced, which, e.g., could be the case if the pyrolysis process is heated by electricity, the gas can be utilised for various purposes. The hot pyrolysis gas can be used directly in industrial processes, as boiler fuel and used for steam production. Cooled pyrolysis gas can be used for electricity production via specially designed gas turbines or gas motors, or it can be upgraded to other products via different process steps. However, this would require a rather large-scale unit to be economically feasible. One option is to heat the gas to around 1000°C, thereby cracking all long hydrocarbon chains and delivering a synthesis gas that can be used for subsequent synthesis of e.g. methanol. Other pathways are under development, including bio-methanation of the non-condensable part of the pyrolysis gas after bio-oil condensation, currently being developed by Frichs and Stiesdal in collaboration with SDU, DTU, and Biogasclean.

Pyrolysis oil

Pyrolysis oil differs significantly from crude oil in its chemical composition, particularly due to its content of oxygen and nitrogen compounds. It contains water, typically ranging from 15% to 30% [13, 14], has a comparably low heating value, high oxygen content, and a pH-value of less than 3, compared to a pH value of 5-7 for fuel oil. This low pH is a technical challenge due to corrosion. The chemical composition of pyrolysis oil changes due to chemical ageing reactions during storage (for example polymerisation, condensation, phase separation), which change viscosity and stability [16]. To reduce these issues and produce a useful fuel from the crude pyrolysis oil, upgrading of the pyrolysis oil is typically needed. This is expanded on in chapter 107 Upgrading of Biocrudes.

Energy balance

The amount of energy (heat) needed for the pyrolysis depends on many factors, such as feedstock moisture content. Chapter 86 Fast and Catalytic Pyrolysis and chapter 87 Slow Pyrolysis. contain specific plant examples based on input.

Advantages/disadvantages

Pyrolysis can convert biomass residues into storable energy products, which are valuable as they can displace use of fossil fuels. Meanwhile, biochar provides value in terms of CDR and soil improvement. However, pyrolysis is also associated with high investment costs and requires stable feedstock availability, which can be challenging when relying on residues.

Research and development perspectives

There is a need for more research and development in pyrolysis technologies when applied to biogenic feedstocks. Current systems in Denmark are operating at pilot or demonstration scale and scaling up remains a challenge. A significant part of the R&D needs resides in testing and improving the end products. For biochar, further research is needed to understand its long-term stability in soil and its effectiveness as a carbon removal solution under different conditions. This includes the abovementioned ongoing Danish studies on environmental performance, carbon permanence, interactions with soil ecosystems, and certification frameworks. For pyrolysis oil, upgrading remains a barrier to broader use. Although it shows potential as a renewable fuel, especially in maritime applications, current refinery integration is limited, and further testing and development are needed to improve compatibility and economic viability.

In this catalogue, the status of slow, fast and catalytic pyrolysis technologies along with the upgrading of bio-oil and the research and development perspectives of the technologies are explored in more detail.

In addition, advanced pyrolysis technologies such as catalytic hydropyrolysis, microwave pyrolysis, and flash pyrolysis are being explored to improve energy efficiency, product quality, and feedstock flexibility. These approaches show promise but require further testing and optimisation before they can be widely adopted.

Methane pyrolysis

Methane pyrolysis is also known as methane cracking or methane decomposition. This is the process of converting methane (CH_4) into gaseous hydrogen (H_2) and solid carbon (e.g., carbon black). The carbon can also form other structures, such as nanotubes, and research is exploring its use for soil improvement due to its high surface area.

The reaction is endothermic and requires relatively high temperatures to occur (≥800 °C). The reaction energy can be provided either through conventional means, e.g. electrical heaters, or alternatively using plasma. Traditionally, fossil natural gas is used, but some recent studies [17] [18] have explored possibilities of using biogas or biomethane as input.

Catalytical hydropyrolysis

Catalytic hydropyrolysis (CHyP) of biomass is the thermochemical conversion of biomass into bio-oil or liquid fuels by processing with hydrogen and a catalyst to break down the solid biomass structure to liquid, gaseous and solid components. Typical catalytic hydropyrolysis processing conditions are 375-475 °C in temperature with operating pressures from 10 to 30 bar. CHyP is still in the research and pilot phases. While it shows promise as a single-step method for producing high-quality hydrocarbons from biomass, challenges remain with catalyst stability, hydrogen demand, and process scalability. Current studies emphasise the need for continuous-flow reactor designs and realistic feedstock testing to move toward industrial application [19].

Microwave pyrolysis / microwave cracking

Microwave pyrolysis is an advanced thermal conversion process that uses electromagnetic radiation to uniformly heat biomass, enabling efficient breakdown into products such as biochar and biooil. Unlike conventional pyrolysis, which often relies on combustion of syngas for heat, microwave systems can achieve precise temperature control and high carbon retention using electricity. The Danish company Organic Fuel Technologies (OFT) applies a patented form of microwave cracking that operates at 350°C, minimising hotspots and enabling continuous or semi-continuous processing. The system currently treats activated sludge, removing PFAS and other contaminants, and produces stable biochar suitable for soil carbon storage and biooil under development for maritime fuel use. Facilities range from a pilot plant near Aarhus to a modular full-scale, 30 tonne, setup currently under construction at GreenLab.

Flash pyrolysis

In flash pyrolysis, small biomass particles are decomposed in the reactor at high temperatures with short duration time to produce pyrolysis gas and biochar. In Denmark, Frichs Pyrolysis use a patented flash pyrolysis technology to produce gas and biochar in a modular system. A full-scale plant in Vrejlev is treating 30 tonne dry matter from digestate daily.

The above-mentioned technologies might be considered for detailed descriptions in future updates of the technology catalogue, subject to their technology readiness levels.

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86 Fast and Catalytic Pyrolysis

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Date	Ref.	Description
02/2026	-	Major update with new qualitative text and revised technical and economic data
12/2018	-	Datasheet revised

Quantitative description

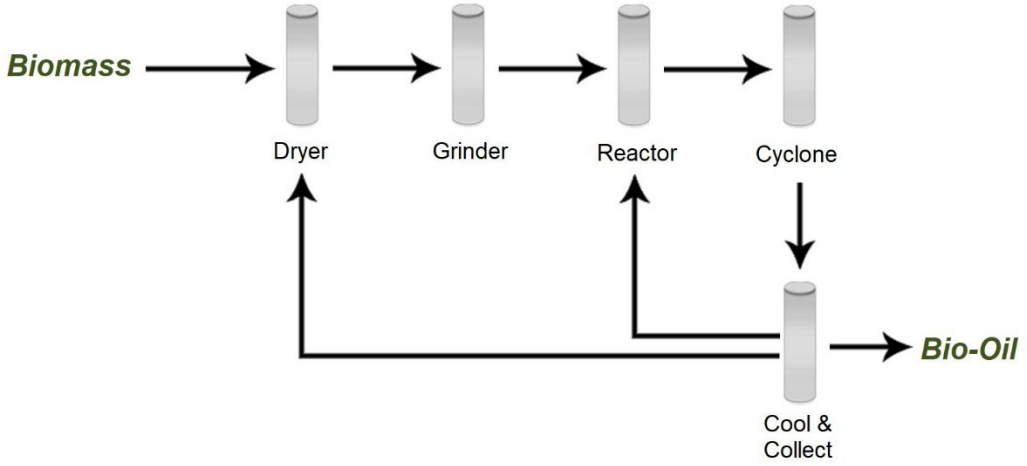
Brief technology description

Fast pyrolysis is a thermochemical process in which biomass is rapidly heated to 450-600°C in the absence of oxygen, typically using sand as a heat carrier. This process produces pyrolysis oil as the main product, along with light gases and biochar. The biochar and gases are generally combusted to provide the necessary process heat. [1]

Key characteristics of fast pyrolysis include very high heat transfer rates, which usually require finely ground biomass, and extremely short vapor residence times, typically less than two seconds, to maximise liquid yield. In practice, vapor residence times of about one second are ideal, although acceptable yields can still be achieved at up to five seconds if the vapor temperature remains below 400 °C. [2]

Fast pyrolysis places stricter requirements on the feedstock compared to slow pyrolysis. This is due to the need for rapid heating, and because the input material largely determines the quality of the oil output. Therefore, wood-based biomasses are the most relevant feedstocks for fast pyrolysis. The moisture content of the feedstock is also critical. After cooling and condensation, a pyrolysis oil is formed which has an energy content of roughly half that of conventional fuel oil.

Figure 1: Fast pyrolysis process flow



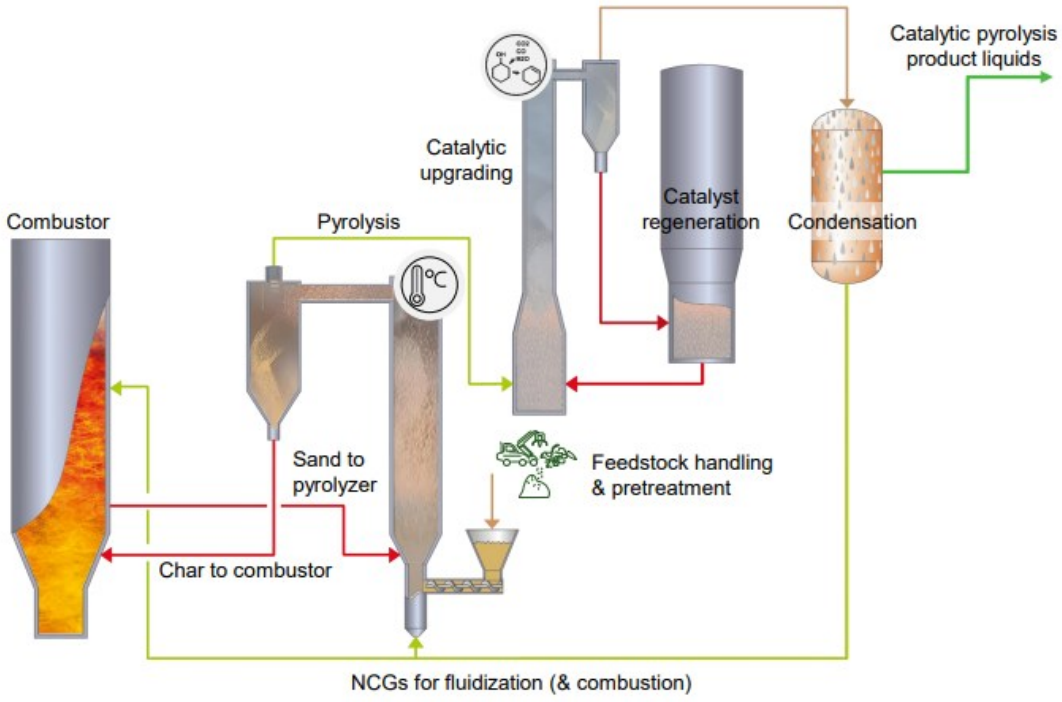
A variation of this process is catalytic fast pyrolysis, where an acid-based catalyst is introduced in situ (during thermal decomposition) or ex-situ (after the pyrolysis process). This approach removes oxygen in the form of CO, CO_2, and H_2O. However, some hydrocarbons are lost as light gases. The oxygen level in the produced oil can be regulated by adjusting the process intensity, a lower oxygen percentage leads to increased gas formation and thus a lower liquid yield. Figure 2 shows an example of ex-situ catalytic fast pyrolysis [3].

The quality of the oil produced from fast pyrolysis, with and without catalysis, respectively, compared to petroleum, is displayed in Table 1.

Table 1: Comparison of oil quality from fast pyrolysis (with and without catalysis) and petroleum [4, 3]

	Fast pyrolysis oil – Woody feedstock	Catalytic fast pyrolysis oil – Woody feedstock	Petroleum
H/C	1.4-1.6	1.0-1.1	1.5-1.8
O	36-52	10-36	0.1-1.0
S	~0	~0	0.1-6
N	<0.2	<0.2	0.1-2
H_2O	17-30	~5-10	0.02-0.1

Figure 2 Valmet Pyrolyzer with catalytic upgrading [3]



This chapter describes large-scale fast pyrolysis plants with different inputs, and catalytic fast pyrolysis, as illustrated in Table 2. The data is specified in the associated datasheets.

Table 2: Overview of Fast Pyrolysis Plants Included in This Chapter

Input	Large scale fast pyrolysis (25MW)	Large scale catalytic fast pyrolysis (25MW)
Wood	X	X
Straw	X	

Inputs

Existing commercial fast pyrolysis plants use clean wood as feedstock. However, some facilities are investigating the use of straw sunflower husks, and EFB (Empty Fruit Bunch from palm oil production) as alternative inputs. This introduces challenges e.g., due to the higher chloride content in straw, which can lead to corrosion and operational issues. A potential solution is washing the straw to reduce chloride levels, but this approach significantly increases operational expenditures (OPEX) because of the additional water consumption and handling requirements. Furthermore, the oil yield is lower compared to wood inputs. Therefore, for these alternatives to be economically viable, their price must be significantly lower than that of woody biomass.

The moisture content of the biomass should be approx. 5-10% after drying, as water ends up in the final product, reducing its quality. The moisture in the raw feedstock (clean wood), is up to 45–50%, meaning that drying is needed. Drying to lower levels is energy-intensive, so 5-10% is generally considered an acceptable compromise. The drying process requires some electricity for operation, while the heating demand is typically met by combusting the produced gases and char. Additional support energy may be required, depending on the feedstock's moisture content coming into the dryer.

Outputs

The primary output from the system is pyrolysis oil, and potentially some biochar and gas depending on the design of the system, operating conditions and feedstock used.

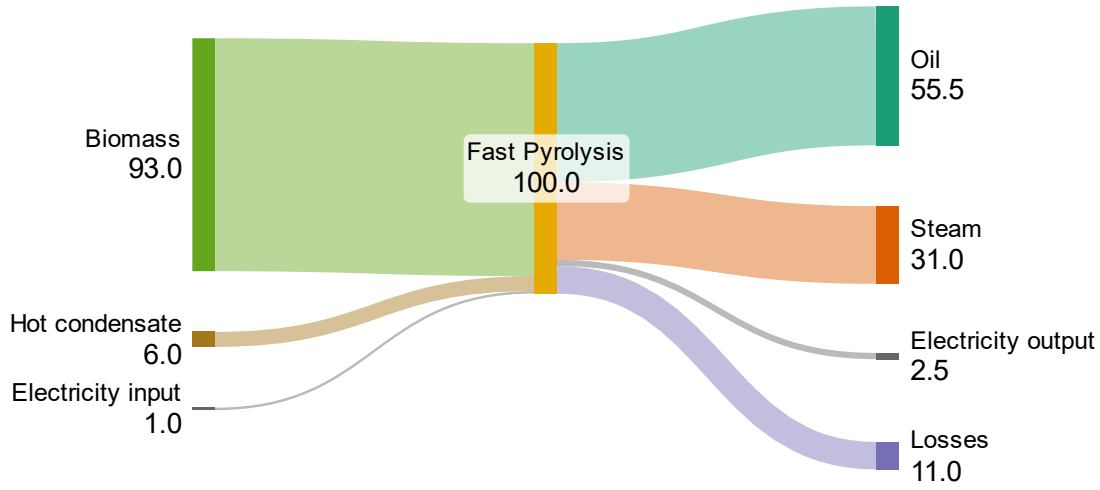
The oil produced through fast pyrolysis can be utilised in several ways. Today it is most commonly combusted for heat generation. Alternatively, it can be upgraded, an option described in detail in chapter 107 Upgrading of Biocrudes, to e.g., produce renewable “drop-in” transportation fuels such as gasoline and diesel. In addition, fast pyrolysis oil can serve as a chemical feedstock.

Energy Balance

The energy balance of the process is illustrated in Figure 3 and is based on operational data from BTG’s Empyro plant in Hengelo, the Netherlands. At Empyro, the biochar produced during pyrolysis is combusted to supply the necessary process heat, enabling energy self-sufficiency. The plant primarily produces bio-oil and steam, with the potential for co-generation of electricity. The steam can be used for heating purposes, and at Empyro, the heat is used both internally for drying, and sold externally as industrial process steam [5]. The moisture content of the feedstock

determines how much of the generated steam is consumed internally and how much can be exported.

Figure 3: Energy balance based on BTG's Empyro plant in Hengelo. [5]



The oil yield from catalytic fast pyrolysis is lower than that of non-catalytic fast pyrolysis. This is due to enhanced cracking and deoxygenation reactions promoted by the catalyst, which convert a larger portion of the feedstock into gaseous products and char.

The catalytic fast pyrolysis oil consists of:

- 90% biocrude, which can be further upgraded and refined into biofuels, including drop-in marine fuels and Sustainable Aviation Fuel (SAF).
- 10% bio-aromatics, which can be used as chemical feedstocks.

The carbon distribution outputs from the catalytic fast pyrolysis process is roughly equal shares of liquid, solid, and gaseous output; With 27-32% catalytic pyrolysis oil, 30-34% non-condensable gas and 32-34% biochar and coke.

Typical Capacities

Currently, there are three commercial-scale fast pyrolysis plants operating in Northern Europe, all utilising BTG Bioliquids technology. In addition, three commercial plants are in operation in North America. BTG's plants have a capacity of approx. 25 MW. The plant capacity is not expected to be scaled up. The benefit of this plant size is that the feedstock is often available at a single location.

Empyro in Hengelo, the Netherlands, began production in 2015 and was the first commercial-scale fast pyrolysis oil plant. It has a capacity of 5 tonnes per hour, producing approximately 24,000 tonnes of fast pyrolysis oil per year, primarily used in a steam boiler. The feedstock is clean woody biomass (e.g. sawdust) [6] [7] [8].

Green Fuel Nordic in Finland started production in 2020 and produces 24,000 tonnes of fast pyrolysis oil annually. The feedstock includes sawmill byproducts and forest residues. One of the primary off takers is a local heating plant (IEA Bioenergy, 2023).

Pyrocell in Sweden started production in 2021 and produces 25,000 tonnes of fast pyrolysis oil annually from sawdust.

The largest operational plant to date is located in Port-Cartier, Quebec, Canada. Commissioned in 2018, it produces 38 million litres of fast pyrolysis oil per year from wood residues sourced from the Arbec sawmill. The feedstock input is 65,000 tonnes/year of dried cellulosic woody biomass. The fast pyrolysis oil is sold to local industrial users and for heating applications [10].

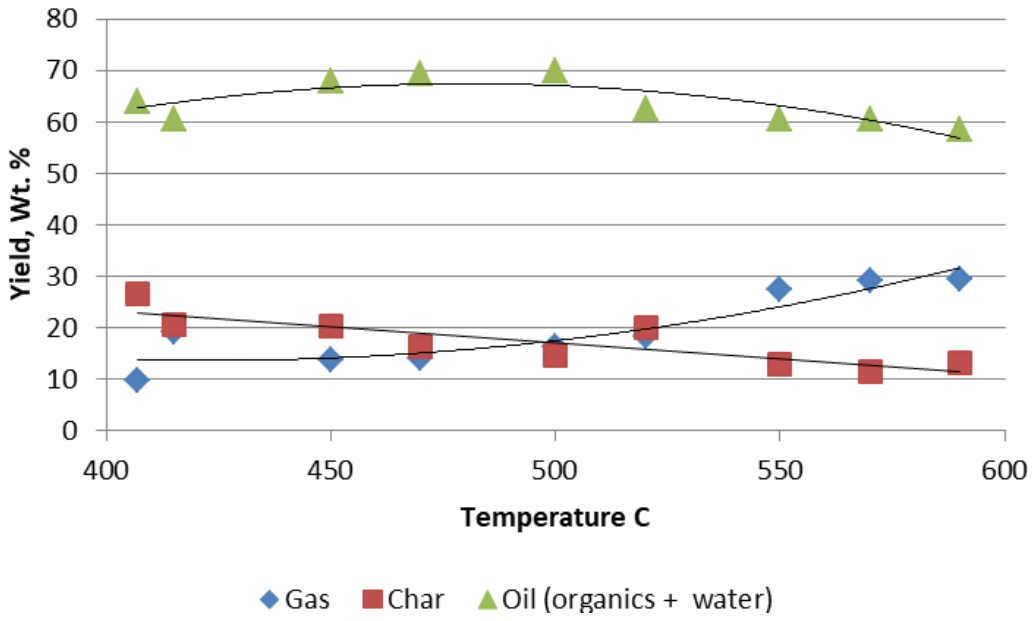
The commercial plants currently in operation have annual production capacities ranging from 24,000 litres to 38 million litres of fast pyrolysis oil.

The largest known planned commercial project is Ensyn's Aracruz project in Brazil, which is currently under development. It has a nameplate capacity of 83 million litres per year.

Regulation Ability

The operating temperature of the reaction zone of the systems drives product yields and ratios of liquids to gas and biochar [11], as displayed in Figure 4.

Figure 4: Typical product Yields vs. Temperature



The regulation capacity of the systems is therefore limited. Operating outside of the optimal temperature zone will result in a loss of oil yield.

Space Requirements

Ensyn operates a 10 million litre per year facility in Renfrew, Ontario. The plant is located on 6.4 ha of land, which thus equates to 0.64 ha/million litres. BTG has an indicative dimension of approximately 2,385 m². This area includes the fast pyrolysis production plant and 250 m^3 of oil storage [5].

Advantages/Disadvantages

Advantages

A key advantage of fast pyrolysis is the production of liquid fuel, which can be used as a renewable energy carrier or as a feedstock for further upgrading into transportation fuels and chemicals. The process also allows for relatively high liquid yields compared to other thermochemical conversion technologies, and it can be integrated into decentralized biomass valorization systems.

For catalytic fast pyrolysis, an advantage is the reduction in oxygen content in the resulting oil output. This leads to improved stability, energy density, and compatibility with existing refining infrastructure.

Disadvantages
Fast pyrolysis also presents several challenges. Compared to slow pyrolysis, fast pyrolysis currently relies almost exclusively on wood-based feedstocks. This limits the flexibility of the process, as it cannot easily accommodate more heterogeneous or moist biomass types. Additionally, the technology faces competition for woody biomass, which is also in demand for other uses, e.g., gasification or combustion for power production, thereby limiting feedstock availability.

Pyrolysis oil differs significantly from fossil crude oil, as it is acidic, oxygen-rich, and thermally unstable, which makes it difficult to store, transport, and use directly. In most applications, the pyrolysis oil must therefore undergo upgrading before it can replace conventional fuels or chemicals [12]. For catalytic fast pyrolysis, a disadvantage is that it tends to break down hydrocarbons, resulting in shorter molecular chains and increased formation of solids.

Environment
The GHG emission performance of pyrolysis oil is good, assuming that the biomass is sustainably sourced, as most of the energy to drive the process comes from the biomass feedstock.

Research and Development Perspective
Fast pyrolysis of biomass for fast pyrolysis oil production is considered to be at Technology Readiness Level (TRL) 9, as the core process has been demonstrated in full-scale and is commercially deployed. However, the current market diffusion of the technology is still in an early phase. In the context of the technology catalogue, this places fast pyrolysis in category 2 (pioneer stage), reflecting that only a limited number of commercial plants are operating, and that uncertainties regarding long-term performance, scalability, and cost competitiveness remain. Recent developments nevertheless indicate continued progress toward broader commercial deployment.

Catalytic fast pyrolysis has been assessed at Technology Readiness Level (TRL) 7, indicating that the technology has reached pilot-scale demonstration. The next phase of development focuses on continued research and establishing stable long-term production at the pilot scale. Over the coming years, the development goal is to gradually scale up the technology. Another central goal is to adapt the technology to agricultural feedstocks, such as straw. [3]

BTG Bioliquids, a Dutch technology provider referred to above, has supplied fast pyrolysis bio-oil (FPBO) technology to several commercial-scale plants in Northern Europe. These include the Empyro plant in Hengelo, the Netherlands, the Green Fuel Nordic plant in Lieksa, Finland, and the Pyrocell plant in Gävle, Sweden, which is a joint venture between Setra and Preem. In Finland, Valmet and Fortum had a fast pyrolysis plant in operation between 2013-2020. Today Valmet has catalytic fast pyrolysis units on pilot scale.

In North America, Ensyn has developed and operated several fast pyrolysis facilities primarily for renewable heating oil applications. These include plants in Renfrew, Ontario, and Côte Nord, Quebec, with additional projects under development in Maine (USA), Nova Scotia (Canada), and Georgia (USA).

These installations represent important steps toward scaling the technology, although fast pyrolysis, particularly in its catalytic form, still requires further development to optimize process efficiency, catalyst durability, and the quality and upgrading of the bio-oil product. Continued

research and operational feedback from existing plants will be essential to reduce costs and support wider adoption.

Examples of Market Standard Technology

While fast pyrolysis of biomass is considered commercial (TRL 9), its deployment remains limited. As of today, only four plants are in operation across Europe.

Valmet and BTG have demonstrated stable, continuous operation and integration with industrial steam users. These systems rely on wood-based feedstocks and with an input between 30-50.000 tons of feedstock a year.

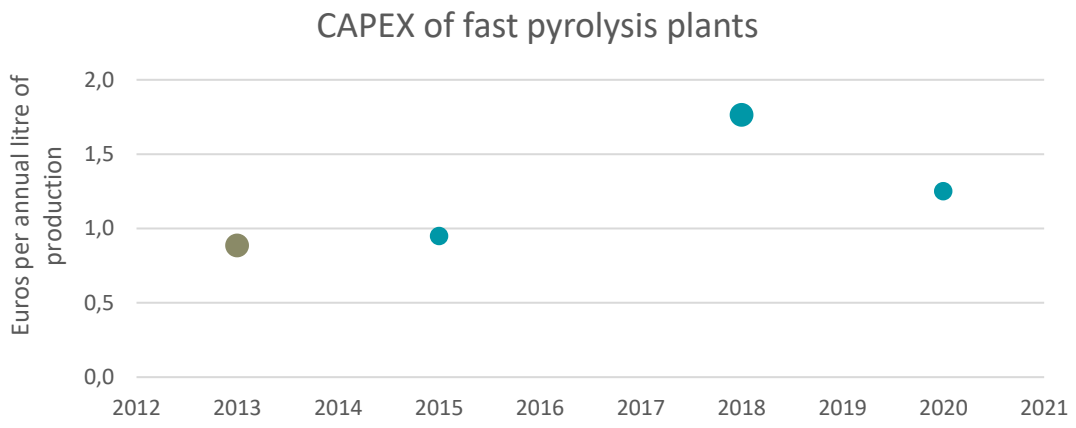
Catalytic fast pyrolysis is still under development (TRL 7), there are therefore no examples of market standard technology as of today.

Quantitative description

Predication of Performance and Cost

Despite the relatively high Technology Readiness Level (TRL) of fast pyrolysis, few commercial-scale plants have been built, indicating the technology is still early in its deployment. The figure below presents CAPEX data for four fast pyrolysis plants commissioned between 2013 and 2020, showing investment costs per annual litre of bio-oil ranging from 0.95 to 1.8 €/litre. Each data point represents a different technology provider, except for the 2015 and 2020 plants, which share the same provider. The facilities are located in Europe and Canada, and the size of each point reflects annual bio-oil output, varying between 20 and 43 million litres. No clear decrease in CAPEX is observed from the available data, which may be explained by the plants being first-of-a-kind, differing in size, technology, location, and influenced by site-specific factors and evolving market conditions.

Figure 5: CAPEX of fast pyrolysis plants commissioned between 2013 and 2020. (the one marked as gray is no longer in operation) (IEA Bioenergy, 2023)



With few new projects currently under construction, it is assumed that CAPEX values will not start to decrease until after 2030. Under the assumption that an increasing number of new plants initiate construction in the upcoming years, a 15% price reduction is anticipated between 2030 and 2040. This is driven by technological improvements, increased deployment, and learning effects across the value chain. After 2040, the pace of cost reduction is expected to slow down, with a further 10% decrease projected by 2050, as the technology approaches maturity and the most substantial efficiency gains have been realised. These CAPEX development rates are based on estimates provided by market actors for slow pyrolysis, though taking into account differences between the two technologies.

Uncertainty

It should be noted that the above CAPEX estimates are quite uncertain, as they are based on relatively few projects, and that interviewed market actors deemed future CAPEX estimates for fast pyrolysis to be sensitive proprietary information. In addition, there is uncertainty with respect to performance and other economic aspects, particularly related to scaling issues due to the early developmental stage of the technology.

Additional Remarks

There is significant interest in this technology pathway from petroleum refiners. The potential of pyrolysis oils to be co-processed in a petroleum refinery to produce gasoline and diesel blending components without oxygen is attractive to refiners who are mandated to lower the carbon intensity of the products used for transportation applications.

Typical Plant Size

The commercial plants currently in operation have annual production capacities ranging from 24,000 to 38 million litres of bio-oil. BTG in Europe have plants at approx. 25 MW. The benefit of this plant size is that feedstock is typically sourced from a single source. There are larger projects currently being developed, e.g., Ensyn's Aracruz project in Brazil with a nameplate capacity of 83 million litres per year.

Input and Output

The input and output volumes for different plant sizes are shown in the following table, based on known projects with different approximate equivalent energy metrics. (IEA Bioenergy, 2023)

Table 3: Input and Output Volumes for different Plant Sizes

Volume Based, Million litres per year	Input Mass Based, thousand tons per year	Energy Output Based, GJ/day
11	25,5	700
20	44	1,300
38	65	2,500
83	150	5,400

Forced and Planned Outage

The plants operate continuously. BTG have stated that they have reached nameplate capacity in 2017 and have reportedly operated on a continuous 24/7 basis ever since. [13]

Compared to non-catalytic fast pyrolysis, catalytic pyrolysis is expected to involve higher downtime, primarily due to the need for catalyst replacement. This issue is particularly pronounced in in-situ catalytic pyrolysis systems.

Technical Lifetime

Plants of this type would normally be designed for a lifetime of 25 years.

Construction Time

Construction times of 18 to 24 months can be expected. For example, the Ensyn, Quebec project started construction in June of 2016 and was in the commission phase in the first quarter of 2018.

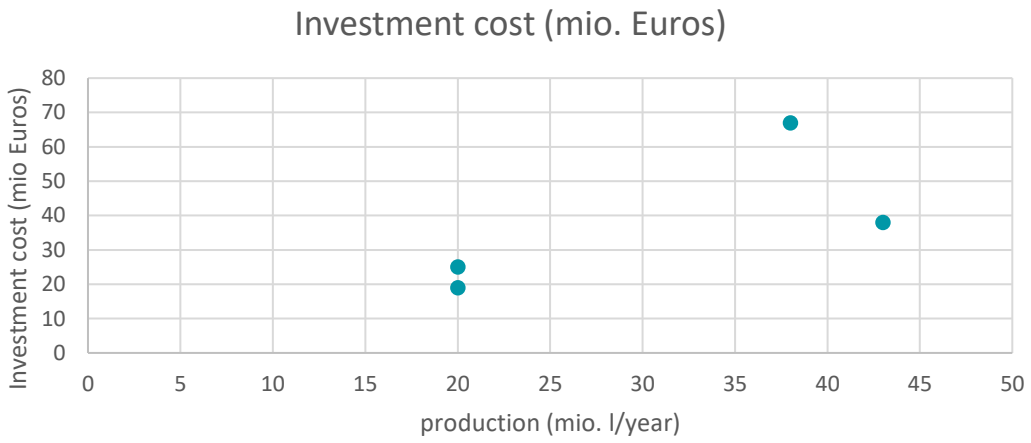
Financial Data

In addition to the publicly available information on the capital costs for the recent BTG and Ensyn facilities, there are several sources of detailed information (Mark M. Wright, 2010) [15] [16] on capital costs and operating costs in the public domain. However, these have been generated by independent third parties and not by the leading process developers.

Investment Costs

The capital investment for fast pyrolysis plants varies significantly depending on scale, location, and integration. The BTG plant in Hengelo, commissioned in 2015, had a total project cost of approximately €19 million, corresponding to an investment of around €0.95 per annual litre of bio-oil production. Similarly, Green Fuel Nordic's facility, commissioned in 2020, required a capital investment of €25 million, or about €1.25 per annual litre. The Ensyn project in Quebec, commissioned in 2018, had a reported capital cost of CAD 103 million (approximately €67 million), equivalent to €1.75 per annual litre of production.

Figure 6: Investment Costs of Plants and corresponding Annual Production Capacities



For catalytic pyrolysis, the best available estimate suggests a plant designed to produce 100,000 tonnes of biocrude per year (equivalent to approximately 85 million litres) would require an investment of around €200 million (2.35 euro/l). This figure is based on input from technology developers and should be considered indicative. As catalytic pyrolysis is still at an early development stage, cost estimates remain uncertain and subject to change as the technology matures and scales.

Operating and Maintenance Costs

OPEX vary depending on the process configuration and technology applied. For fast pyrolysis, OPEX is estimated at around 8% of the plant's CAPEX [17].

In the case of catalytic pyrolysis, OPEX tends to be higher. When the catalyst is applied in-situ, OPEX can reach up to 20% of CAPEX, primarily due to the frequent replacement of catalysts. [17] For ex-situ catalytic pyrolysis, OPEX is generally lower, ranging between 10–15% of CAPEX.

Start-up Costs

The start-up costs are included in the operating cost estimates.

Technology Specific Data

The properties of fast pyrolysis oil are shown in the following table [18] and compared to typical values for heavy fuel oil [19]. The fast pyrolysis oil contains oxygen which reduces the energy

content and increases the density. Due to the differences between fast pyrolysis oil and heavy fuel oil, fast pyrolysis oil cannot be assumed to substitute heavy fuel oil without any upgrading process. Upgrading of fast pyrolysis oil is discussed in chapter 107 Upgrading of Biocrudes.

Table 4: Fast Pyrolysis Bio-Oil Properties

Parameter	Fast Pyrolysis Oil	Heavy Fuel Oil
Energy Content, MJ/kg (LHV)	16	39
Water Content, wt. %	25	0.1
Density, kg/m^3	1.20	0.98
Oxygen Content, wt. %	47	0
Pour Point, C	-36	15
Flash Point, C	50	100
Kinematic viscosity at 40 °C, mm^2/s	13	
Kinematic viscosity at 50 °C, mm^2/s		200-600

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87 Slow pyrolysis

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Author: Ea Energyanalyse

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Amendments after publication date

Date	Ref.	Description
02/2026	-	Major update with new qualitative text and revised technical and economic data
02/2019	-	Financial data added for configuration 1

Quantitative description

Brief technology description

Slow pyrolysis is a thermal conversion process where organic material is heated in absence or near-absence of oxygen. Compared to fast pyrolysis, slow pyrolysis operates with lower heating rates typically below 10 °C/min (with many systems operating at 1–5 °C/min) and longer residence times, typically between 5 and 30 minutes, depending on the technology and desired product distribution. Solids residence time in slow pyrolysis, especially in biochar-focused systems, can extend to hours, whereas vapour residence time is usually kept short to control secondary reactions.

The process is primarily used for the production of biochar, especially in small-scale plants, although several suppliers are now planning larger-scale installations (see section Typical Capacities). Table provides an overview of input configurations for small (1–2 MW) and large-scale (20 MW) systems included in the data sheets associated with this chapter.

Table 1: Overview of technology and input configurations exemplified in the current chapter.

Input	Small scale (1-2MW)	Large scale (20MW)
Straw	X	X
Digestate	X	X
Garden waste		X

Although the primary driver for installing slow pyrolysis systems today is the production of biochar (*except for sludge treatment, where the main objective is waste management rather than biochar production*), the process also generates a gas and oil fraction. In most systems, these co-products are used internally, for example, to support drying or to provide heat for the pyrolysis process itself. However, depending on the energy balance of the system, there is potential for surplus energy that could be upgraded and used for other purposes, such as industrial heat or transport fuels.

Slow pyrolysis is generally more tolerant of feedstocks with higher moisture content and lower purity than fast pyrolysis. This is primarily because the process is often designed for biochar production, which is less sensitive to feedstock composition. In contrast, pyrolysis oil requires a more controlled input quality, as its chemical composition must meet standards with respect to oxygen content, heating value, acidity, etc., to allow for upgrading. Slow pyrolysis systems can accommodate inputs such as sludge, digestate, and garden waste, which would typically be unsuitable for oil-focused processes. Some systems can switch relatively freely between feedstocks, provided they meet certain requirements such as appropriate moisture content and, in many cases, pelletisation or similar pre-treatment.

Reactor types are generally divided into directly heated and indirectly heated systems:

[1] In directly heated reactors, a portion of the feedstock is combusted with a small amount of air to generate the required process heat.
[2] In indirectly heated reactors, heat is supplied externally, often by combusting pyrolysis gas.

The operating temperature typically ranges between 300 and 700°C, with most systems applying 500–600°C. The temperature, along with feedstock characteristics and reactor design, influences the distribution of end products. Lower temperatures favour biochar production, while higher temperatures increase the yield of pyrolysis gas. Additionally, higher pyrolysis temperatures typically result in higher carbon stability in the produced biochar. [1] Biochar produced at very low temperatures often does not meet the stability criteria defined by the European Biochar Certificate (EBC). Typically, process temperatures above 550°C are required to comply with these standards and ensure long-term carbon sequestration. [2]

Energy demand for the pyrolysis process varies depending on feedstock moisture and system design. This is further elaborated on in the Energy balance section. Some systems use electric heating, while others rely on self-generated gas or autothermal combustion, where no external energy source is needed.

Figure 1: Approximative typical distribution of products depending on pyrolysis temperature. Other factors than temperature also have an impact of this distribution, e.g. particle size, raw material and type of technology used (Chowdhury Z. , 2017). It should be noted that biochar yields at low temperatures (~300 °C) typically do not meet EBC stability criteria (e.g., H/C org), whereas durable carbon-removal applications generally require higher process temperatures (often ≥ 550°C) and sufficient residence time.

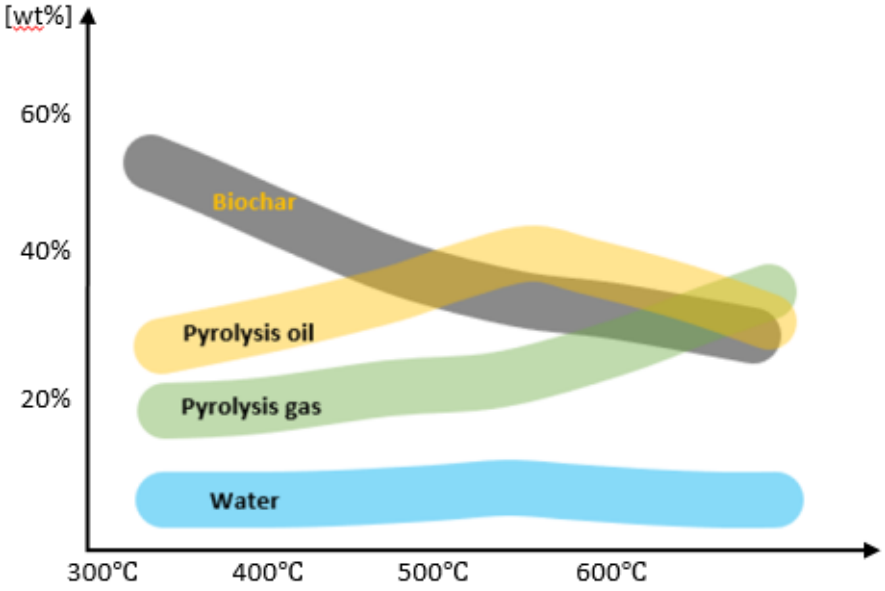
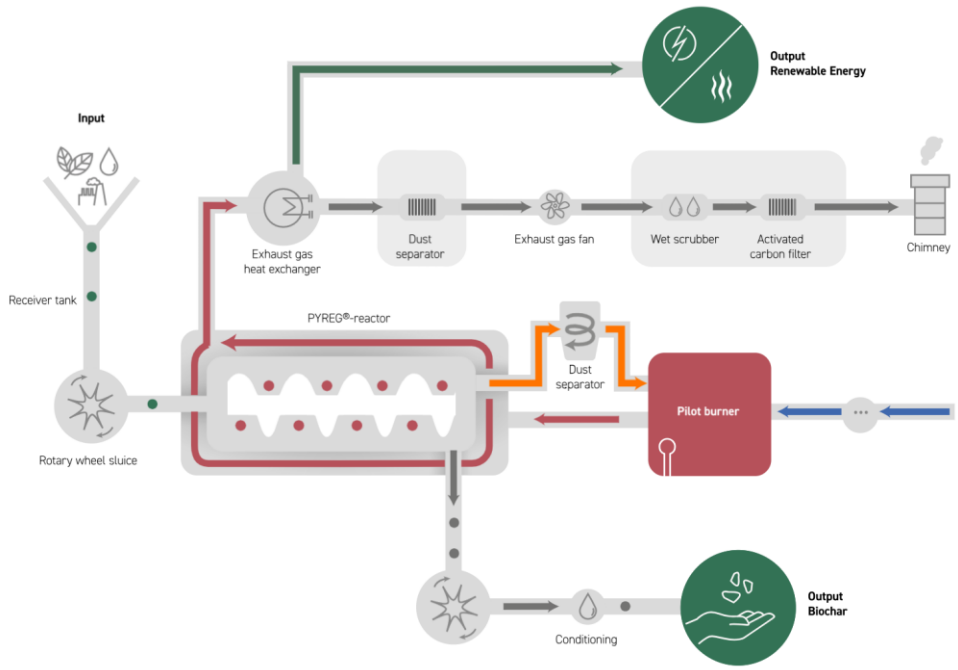


Figure 23: Illustration of a small-scale slow pyrolysis plant configuration (PYREG). [3]



Inputs

The feedstock used in the examples specified in this chapter are straw, digestate from biogas production, and garden waste (for large scale pyrolysis only). Other feedstocks which are used in slow pyrolysis processes include manure, wood residues, saw dust, sludge, and other biogenic waste material.

Slow pyrolysis can tolerate higher moisture content than fast pyrolysis and the configuration of slow pyrolysis plants highly depend on the input used. For feedstocks with high moisture content, drying is needed. In addition to the feedstock input, electricity is needed to run equipment such as conveyers, pumps, etc.

The data presented for slow pyrolysis does not include mechanical dewatering. However, for feedstocks with high water content, such as sludge or digestate, mechanical dewatering is a necessary pre-treatment step prior to drying. Two common methods for dewatering wet feedstocks are decanter centrifuges and screw presses. While both aim to reduce moisture content before drying and pyrolysis, they differ in performance, energy use, and suitability for different feedstocks [4]. The overall assessment is that screw presses are more widely used for pre-treatment of wet biomass input to slow pyrolysis.

Decanter centrifuges use high-speed rotation to separate solids from liquids and are generally more effective at achieving a lower final moisture content. They are well-suited to fine or homogeneous sludges but tend to have higher energy consumption and maintenance requirements. [4]

Screw presses, on the other hand, apply mechanical pressure through a rotating screw. They are simpler, more energy-efficient, and often preferred for fibrous materials. However, they typically leave a higher residual moisture content and may also lead to loss of organic matter and carbon, as part of the dissolved and fine particulate material is expelled with the liquid phase. This can reduce the overall carbon yield of the pyrolysis process, depending on the feedstock and press configuration. [4]

Large pyrolysis plants operating on digestate from biogas facilities require substantial amounts of digestate, making co-location with biogas plants advantageous. Furthermore, co-location can enable heat recirculation and improve process efficiency. However, this principle applies to all feedstocks; the availability of the input resource is critical when assessing the feasibility and scale of pyrolysis plants. Limited or dispersed feedstock supply can significantly impact logistics and costs.

Outputs

The main output is biochar, as well as significant amounts of pyrolysis gas, which often is used for heat production. While the pyrolysis gas is commonly used for heat generation in industrial processes or district heating, in some process configurations part of the pyrolysis gas is utilised by condensing the condensable fractions into pyrolysis oil. The prolonged residence time of vapours in the hot zone promotes secondary cracking reactions, leading to increased formation of char and non-condensable gases at the expense of bio-oil yield. As a result, slow pyrolysis typically produces lower yields of bio-oil. In addition to residence time, several other process parameters, such as heating rate, pyrolysis temperature, and reactor design, affect bio-oil yield, quality, and stability. [6]

Approximately 35 to 60% of the carbon in the feedstock will end up in the biochar (see section regarding carbon balance below). There are several possibilities and challenges related to the utilisation of biochar for soil improvement and as a CDR technology.

Energy balance

Heating the feedstock to pyrolysis temperature requires an energy input in the form of heat. Furthermore, heat is also required for vaporization of the residual water in the feedstock. Therefore, from a pyrolysis point of view, it is important to feed the process with as dry material as possible. However, energy for drying will have to be added in a separate drying step in case the feedstock is wet. With low-cost feedstock such as digestate or sludge, both fractions have a high moisture content, thus drying of the feedstock is an important part of the energy balance. Drying is itself an energy-intensive process, and there is therefore a trade-off between reducing moisture

content and minimising energy consumption. In practice, this means that the feedstock is typically dried to a moisture content of roughly 8–15%, which balances energy consumption in pre-treatment drying and heat demand in the pyrolysis process.

In the pyrolysis process itself, both exothermic and endothermic reactions occur, i.e. both reactions release and consume energy. However, the overall process is almost thermoneutral. For design purposes, one may neglect the heat of reaction for the pyrolysis process, but it is necessary to calculate the energy required for vaporization of products and for heating of feedstock gases and char to the pyrolysis temperature [5].

The specific energy balance is process-dependent, and here it is illustrated using examples from the data sheets for this chapter. Most slow pyrolysis plants utilise part of the energy produced internally. If the system generates excess energy, the gas/heat can be used for district heating, or in co-located processes with a heat demand, such as biogas production. While some excess gas from the process can be condensed, the quality of the resulting liquid fraction is typically low, and doing so complicates the process. Consequently, only in large-scale plants with straw as input is the process shown with a pyrolysis oil output.

The energy balances are largely based on input from pyrolysis producers using data from commercial plants. In certain cases, it has been necessary to supplement with information derived from literature, including for example a large plant with garden waste as feedstock, since it has not been possible to obtain data from large-scale facilities using garden waste as feedstock.

The energy balance for each configuration has been included as a process flow diagram, thus providing a visual overview of energy inputs and outputs. The diagrams are all based on Lower Heating Value (LHV).

Before displaying the input specific process flow diagrams, a simplified Sankey diagram is included (Figure 3) to illustrate the core pyrolysis process from a ~20 MW biomass input, with the direct outputs (gas and char), without showing subsequent utilisation of the gas. This provides a clear overview of the energy distribution within the pyrolysis step itself. The diagrams thereafter display process designs from developers, including the full process for different inputs and outputs based on the data provided.

Figure 3: Sankey diagram displaying the energy distribution in the core pyrolysis process for a 20 MW biomass input. The diagram illustrates the conversion of biomass into gas and biochar. The pelletized input biomass has a water content of 10 wt%.

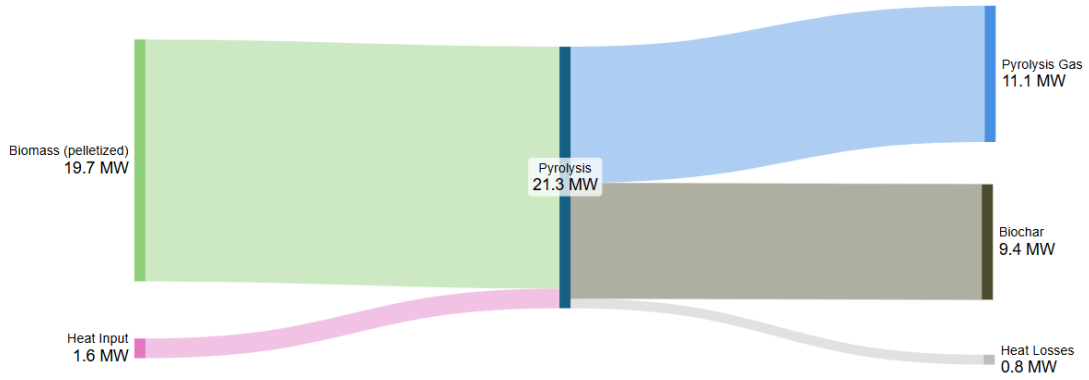


Figure 4: Process configuration for a large plant based on pelletized digestate input. The LHV of the input digestate is 16.4 MW which corresponds to 20 MW on a dry basis. The steam from the drying process can be condensed to a liquid form and one can recover some of the drying heat.

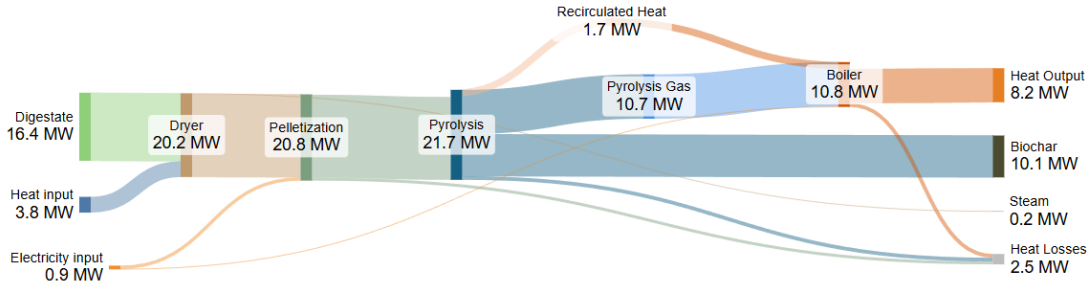


Figure 5: Process configuration for a large plant based on pelletized straw input. The water content in the straw is approx. 10 wt%.

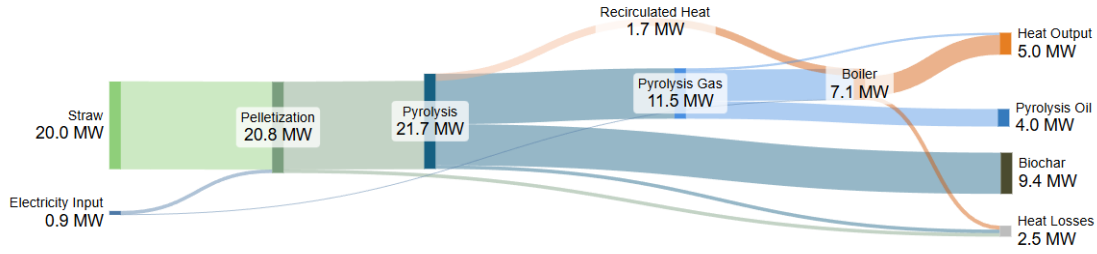


Figure 6: Large-scale garden waste configuration based on model results: [6, 7]. The LHV of the input Garden Waste is 18.2 MW which corresponds to 20 MW on a dry basis.

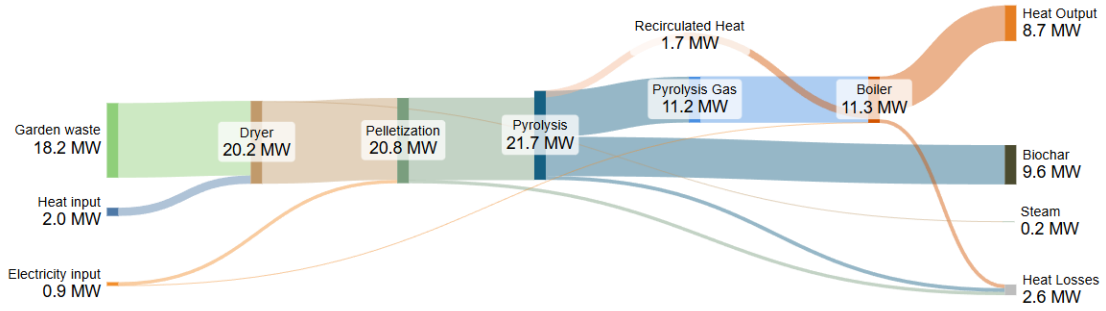


Figure 7: Process configuration for a small plant based on straw input. The information is based on input from Pyreg and Biomacon.

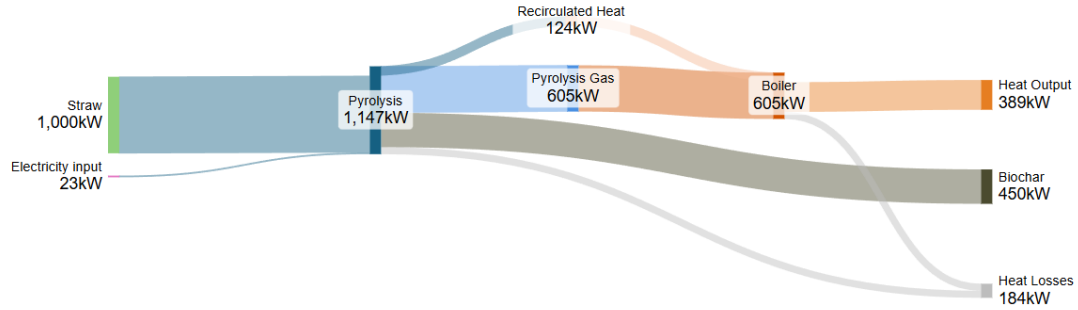


Figure 8: Process configuration for a small plant based on digestate input. The information is based on input from Pyreg and Biomacon. The energy demand for drying the digestate is expected to match the heat demand of the process itself, meaning that all available heat must be reused internally. Otherwise, additional external heat input is required

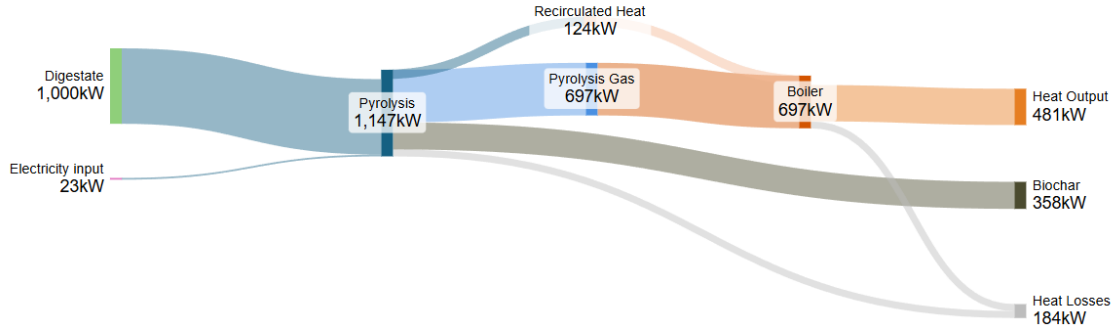


Table 2: Energy balance for exemplified large scale slow pyrolysis plants.

Large scale (20MW)	Straw*	Digestate*	Garden waste*
Pyrolysis gas/heat	24%	40%	46%
Pyrolysis oil	20%		
Biochar	47%	48%	39%
Losses	9%	12%	15%

*apart from biomass input there is an additional 5% electricity input to the process.

Table 3: Energy balance for exemplified small scale slow pyrolysis plants.

Small scale (0.5-2MW)	Straw*	Digestate*	Sludge
Pyrolysis gas/heat	38%	47%	53%
Pyrolysis oil			
Biochar	44%	35%	29%
Losses	18%	18%	18%

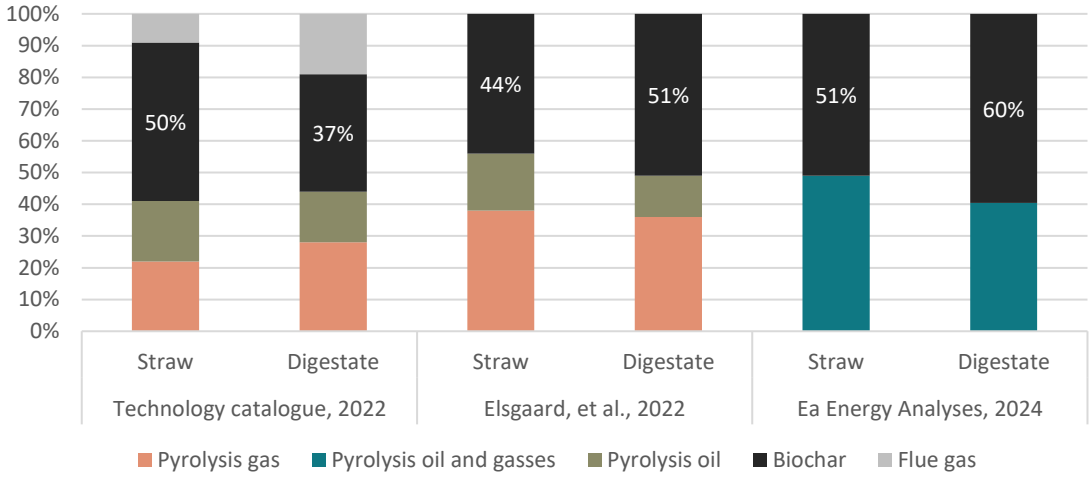
*apart from biomass input there is an additional 2% electricity input to the process.

Carbon balance

Slow pyrolysis results in three main product streams: biochar, pyrolysis gas, and pyrolysis oil (condensable fraction of the gas). The distribution of carbon between these outputs depends on feedstock characteristics, reactor design, and operating temperature. Typically, approximately 35–60% of the carbon in the feedstock ends up in the biochar fraction, while the remainder is released as gas and oil. The carbon in the gas and oil fractions is often used internally for process heat but may also be upgraded for external use depending on system configuration.

A comparison of carbon balances from different sources highlights the influence of process conditions, or system design. According to Ea Energy Analysis 2024 [8], 51% of the carbon from straw ends up in biochar, while 60% of the carbon from digestate is retained in the biochar fraction. In contrast, data from Elsgaard et al. [11], shows a different distribution: 51% of the carbon from digestate goes to biochar, but only 44% from straw. Differences can result from several factors: (i) operating conditions such as temperature and residence time, (ii) feedstock properties, including moisture, particle size, and ash/mineral content; (iii) reactor design and heating rate, and (iv) measurement approach and system boundaries.

Figure 9: Comparison of carbon balance data from different sources per unit of biomass carbon input into the slow pyrolysis process.



Typical Capacities

Most slow pyrolysis plants in Europe are small-scale facilities. In 2023, over half of the 171 operational plants had an annual capacity below 1 MW thermal input. [9] In the Nordic region, most installations are pilot or small-scale, though several larger plants are emerging, with capacities up to 20 MW thermal / 15,000 tonnes of biochar annually. The achievable scale of a pyrolysis plant is influenced by several site-specific factors, including feedstock availability and characteristics, transport logistics, and opportunities for integration with surrounding industrial activities or local energy systems.

Regulation Ability

Pyrolysis is considered a flexible process since the distribution of products changes as the operational parameters are changed. The most important parameter for the yield and quality of the biochar and for the yield of gas and pyrolysis oil is the pyrolysis temperature, defined as the maximum temperature in the reactor (see Figure 23 above). Longer residence times and low heating rate (°C/min) also increase the biochar yield and lower the yield of gasses [10]. A low heating rate could be achieved by decreasing the flow of biomass and increasing the particle size; however, this would be dependent of the reactor configuration and other technology-specific parameters.

A slow pyrolysis plant should be operated in continuous mode apart from maintenance shutdowns, for the following reasons:

- Many stops and start-up sequences increase the wear of the plant and thereby the maintenance need.
- The quality of the products will be affected by the changed operating conditions during start and stop, and during heat-up, the reactor operates under non-steady-state conditions, resulting in lower conversion efficiency and deviations in product distribution until stable operating temperature is reached.
- For some technologies, a fossil fuel is needed to heat the process prior to start-up.
- The emissions will usually increase during start up and shutdown.

Depending on technology, the regulation ability varies. While not described further in this chapter, for electric input pyrolysis, the Biogreen plants report a very broad operating window, nearly 0 to 100% (Chowdhury Z. , 2017), whereas Carbofex report that there is a minimum capacity, however, quite low (Tukainen, 2021).

There might be other equipment in the total system than the pyrolysis core process that sets minimum capacity, however such limitations could be mitigated by doubling of certain equipment, introducing recycle streams, etc.

Space Requirement

Stiesdal informs that a commercial 20 MW plant will have a total footprint of approximately 45 x 50 meters, including equipment for drying and pelletising [12]. For smaller-scale systems, Pyreg reports that its 500 kW unit measures 12 x 6 meters, while the 1,500-kW unit has dimensions of 13 x 7 meters [3]. In addition to the core pyrolysis equipment, additional space is typically required for feedstock and product storage, pre-treatment, infrastructure, and facilities for personnel.

It is advantageous to locate plants for slow pyrolysis close to the biomass source, e.g. at biogas plants, since transportation of the feedstock is more costly than transportation of pyrolysis output products. However, this is very dependent on the type of feedstock, as straw can for example be transported longer than digestate.

Advantages/disadvantages

Slow pyrolysis has several characteristic strengths and limitations that relate to its process design and operational behaviour. One of the key advantages is its ability to produce a high yield of biochar, which is central when carbon sequestration or solid carbon products are prioritised.

The slow pyrolysis process itself can be considered a robust technology, at least for the types proven in larger scale. However, the practical execution of the technology is not easy and new technology suppliers delivering these kinds of processes often run into operation and maintenance problems needed to be solved prior to commercialising. Such challenges can be tar formation/clogging, material problems/corrosion, and issues with control and automatic safety systems, mainly in the systems handling the produced gas.

Tar is a black liquid with a high viscosity, found in colder parts of the process, consisting of condensed hydrocarbons together with oxygen-, nitrogen-, and sulphur-rich compounds. . Because of its complexity and touling tendency, systems required to separate tars from gas are generally more advanced and costly than standard cleaning systems [10].

As for all plants handling combustible, and potentially explosive, material, the risks must be taken seriously and be mitigated as far as possible. Corrosion, plugging, leakages and other sources of risks must be handled in a structured manner. This is especially important for new technologies and when scaling up from e.g., a pilot plant to a commercial scale.

Environment

From a greenhouse gas (GHG) perspective, the net emissions from pyrolysis can be negative if the resulting biochar is used for carbon storage. The long-term stability of biochar in soil, and thus its actual climate mitigation potential, is currently the subject of ongoing research. While biochar is generally considered a stable form of carbon storage, factors such as feedstock type, pyrolysis conditions, and soil characteristics can influence its degradation rate over time. Similarly, the impact of biochar incorporation on soil microbial life is not yet fully understood. Some studies suggest potential benefits for soil health, while others raise concerns about possible disruptions to microbial communities. Given these uncertainties, the environmental benefits of biochar depend on the outcomes of current and future research. The topic is not elaborated further here.

Polycyclic aromatic hydrocarbons (PAHs), which are considered environmentally hazardous, can form during the pyrolysis process and remain in the biochar. Their concentration is influenced by several factors, particularly the peak pyrolysis temperature, with higher temperatures generally resulting in lower PAH levels. [13] Other relevant factors include reactor pressure, gas residence time, and feedstock type. According to the European Biochar Initiative, the maximum PAH

content for basic-grade biochar is 12 mg/kg [14], though actual concentrations can range from 0.1 to over 10,000 mg/kg depending on process conditions. [15]

Interviewed stakeholders have highlighted that one effective strategy to minimize PAH concentrations is to prevent the condensation of pyrolysis gases onto the biochar. In practice, this approach has yielded consistently low PAH levels in the final product.

Air emissions are an inherent part of pyrolysis operations, particularly from the combustion of pyrolysis gas used to heat the reactor or other parts of the plant. These emissions typically include sulphur dioxide (SO_2), nitrogen oxides (NO_X), particulate matter, volatile organic compounds (VOCs), and hydrogen chloride (HCl). Operators highlighted that maintaining a slight negative pressure within the pyrolysis system is a practical approach that has shown good results in reducing air emissions. By ensuring that the system operates under mild vacuum conditions, it becomes easier to contain volatile gases and guide them through appropriate treatment processes, thereby improving environmental performance. [16]

Water will be required for cooling and wetting, or in some cases quenching, of the biochar. Furthermore, water will be used for cooling of the pyrolysis gas and for other parts of the process, depending on the setup of the plant. During quenching the steam formed can contain several contaminants such as H_2S, NO_X, CO_2, CO and H_2 and leave the plant as a contaminated water stream. [18]

Research and development perspectives

As noted in previous sections, slow pyrolysis is an emerging technology. The knowledge and practical use of pyrolysis can be considered as a mature technology; however, all new suppliers will have a learning curve when it comes to scaling up the process. Ensuring that a pyrolysis plant can operate continuously without unplanned shutdowns remains a technical challenge, particularly in larger systems and represents an important focus area for ongoing research and development.

In the coming years, research and development activities within slow pyrolysis and biochar will be strongly influenced by regulatory clarification and the need for knowledge-building.

Ongoing research and regulatory framework development will in part determine the future of slow pyrolysis. Three central themes stand out:

[1] Soil health and the biochar regulation
There is a need for comprehensive documentation on how biochar applied to soil affects soil health, including nutrient retention, microbial activity, and long-term impacts on crop yields. The establishment of a national biochar regulation depends on closing this knowledge gap.

[2] Carbon stability and long-term storage
The climate relevance of biochar depends on its ability to store carbon over extended periods. Research is needed to clarify how stable biochar is in different soil types and under varying environmental conditions, and how this is influenced by feedstock and pyrolysis parameters.

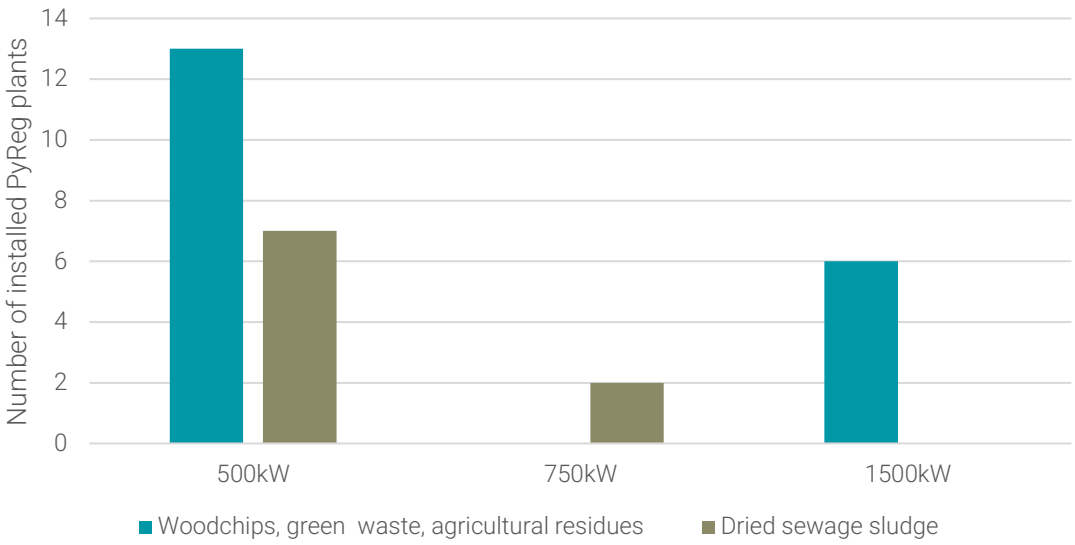
[3] Clarification of economic incentives and support mechanisms
The future development of slow pyrolysis will be influenced by how economic incentives are structured. This includes the role of voluntary carbon credits, potential public investment support, and the conditions under which biochar can be integrated into corporate climate strategies. Clarification in this area is essential for enabling broader market participation.

Examples of market standard technology

The first full-scale Stiesdal SkyClean plant has an input capacity of 20 MW. This is located in Vrå, Danmark, where it was built next to a biogas plant, and the input for the pyrolysis process is digestate from that biogas plant. Leading up to this full-scale facility, Stiesdal built pilot plants of 200 kW and 2 MW, respectively.

For production of biochar as a single product there are numerous technology providers providing proven technology with numerous reference plants. These suppliers have equipment and plants ranging from small-scale, comparably simple units, used e.g. at single farms, up to larger industrial scale facilities. Examples of suppliers of small-scale slow pyrolysis plants are the German producers Pyreg and Biomacon. Pyreg report to have installed close to 30 units and Biomacon over 20. Most of these are around 500 kW.

Figure 10: Overview of installed Pyreg slow pyrolysis units [17].



For slow pyrolysis of sludge, Aquagreen's technology is commercial, with three full-scale plants currently in operation and two additional units under construction. Each unit has a capacity of approximately 400 kW. The first two plants were commissioned in 2022, marking an important step in the commercialisation of sludge pyrolysis. Pyrolysis of sludge is primarily applied as a waste management technology which is the rationale behind it not being detailed in this technology catalogue. The energy balance depends on the dry matter and energy content of the incoming sludge, and in some cases, the system can be energy neutral. Technology developers indicate that gas-fired auxiliary heating is commonly needed in practice, usually around 5 hours each week. In commercial solutions that typically include steam dryers, district heating can be produced through condensation of steam, provided the heat is not reused within the process.

Prediction of performance and costs

Looking ahead, the costs are expected to gradually decline after 2030, driven by continued establishment of new plants and technological optimisation. This anticipated cost reduction aligns with typical trends observed in emerging clean technologies, where early-stage investments are followed by efficiency gains and lower unit costs as deployment increases (learning curve effects). Detailed analysis of potential cost developments has not been conducted. However, based on historical trends observed in other renewable technologies and inputs from developers, it is reasonable to assume a potential annual reduction in capital expenditure (CAPEX) of approximately 3%. If 10 plants are constructed within five years, this would correspond to a CAPEX index of 86, relative to the baseline (index 100).

Uncertainty
There is a degree of uncertainty associated with several aspects of pyrolysis technology and its implementation. This includes variability in cost estimates, which are influenced by fluctuating component prices, evolving regulatory requirements, and differences in site-specific conditions. For example, recent CAPEX assessments, such as those provided by Stiesdal, are higher than previously published figures, partly due to unforeseeable conditions affecting the market.

Quantitative description

Typical Plant Size
The examples in this chapter are for large-scale plants with 20 MW inputs of straw, digestate, or garden waste and small scale 1MW plants for straw and digestate input. Costs for pretreatment is included in the calculations.

Inputs and Outputs
The inputs and outputs in terms of energy quantities are described in more detail under the energy balance section. For small plants, the output in terms of weight corresponds to between 19% and 32% of the input, depending on the plant [18]. For large pyrolysis plants, the share is higher, at around 50% of the input, while approximately 30–40% of the input is converted into gas, which can be partially condensed into oil [19]. The difference between the mass input and output is mainly due to the water content (both in the feedstock and formed during the process) and the flue gas components resulting from the combustion used to supply heat to the process.

Forced and Planned Outage
For pyrolysis plants operating under standard conditions, typical maintenance schedules include short, planned stops approximately every 30 days, as well as two longer scheduled shutdowns per year, typically lasting 2 to 5 days [20]

Technical Lifetime
Small pyrolysis plants are designed with a lifetime of over 15 years. For large-scale pyrolysis plants, a typical design lifetime is between 20 and 25 years. In contrast, the buildings housing the plant are expected to last up to 60 years.

Construction Time
For large pyrolysis plants, after all permits have been approved, Danish developers estimate a construction time of 16–20 months.

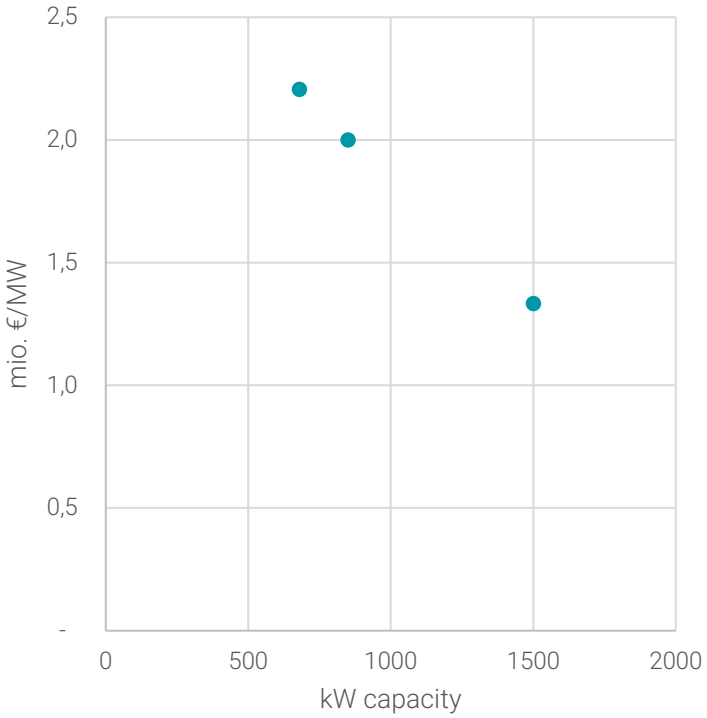
Financial Data

Investment Costs
Current CAPEX estimates for large-scale pyrolysis facilities are based on Stiesdal's commercial assessments and reflect increased market prices due to inflation and rising component costs. Investment costs are approximately €1 million/MW for straw-based plants and €1.5 million/MW for digestate-based plants.

For small-scale slow pyrolysis, cost data is derived from Pyreg and Biomacon [18]. Specific investment figures for sewage sludge treatment plants are unavailable, but payback time is estimated at 5–10 years, depending on PFAS content, as municipalities are willing to pay more for handling contaminated sludge.

Preliminary data on CAPEX for small scale plants indicate a linear relation between capacity and CAPEX. For a 1 MW plant, the investment cost is therefore approximately 1.8 million euros.

Figure 11: Cost data derived from Pyreg and Biomacon [18]



Operation and Maintenance (O&M) Costs

Operating expenses (OPEX) are expected to be in the range of 10–15% of CAPEX annually, depending on plant type and feedstock.

Start-up Costs

The start-up costs are included in the operating cost estimates.

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89 Hydrothermal Liquefaction

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Publication date

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Amendments after publication date

Date	Ref.	Description
12/2018	-	Datasheet revised

Qualitative Description

Hydrothermal liquefaction of biomass is the thermochemical conversion of biomass into liquid fuels by processing in a hot, pressurized water environment for sufficient time to break down the solid biomass structure to mainly liquid components. Typical hydrothermal processing conditions are 250–420°C of temperature and operating pressures from 40 to 350 bar of pressure.

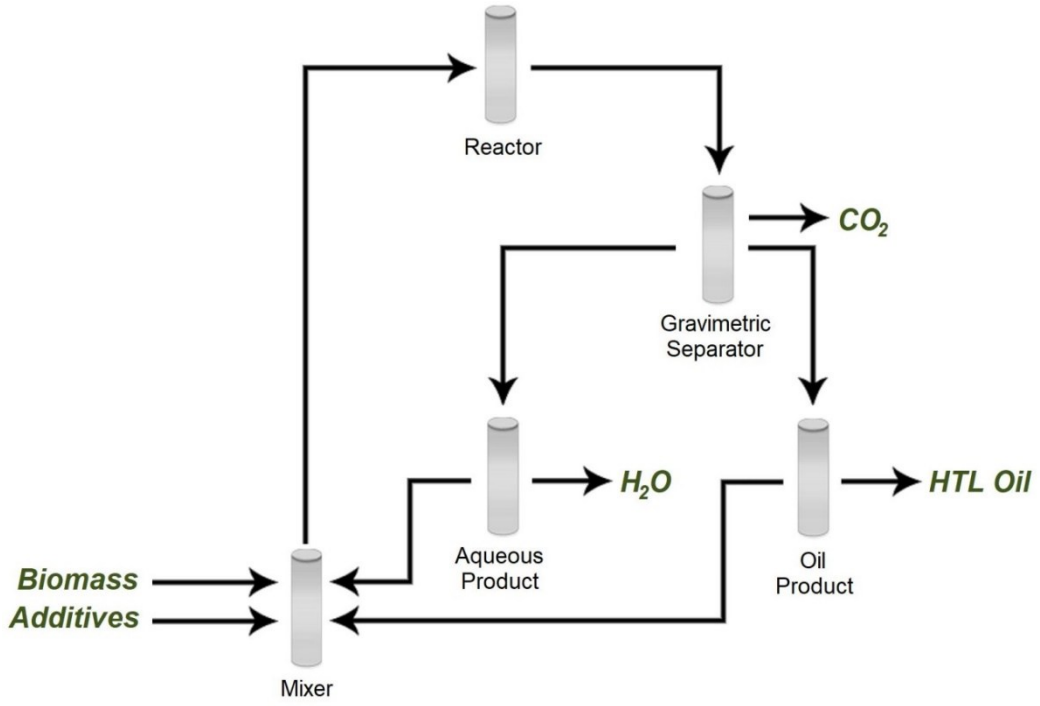
Alkali catalysts are often used to facilitate the degradation of macromolecules by hydrolysis, decarboxylation, and depolymerisation type of reactions, as well as inhibit formation of tar, char, and coke.

The process can handle most types of biomasses and unlike some other thermochemical process it does not require dry biomass.

Brief Technology Description

A biomass slurry is fed into reactors at high pressure and high temperature. The slurry contains water, biomass and the alkali catalysts. The reactors provide sufficient residence time for the solid biomass to breakdown. After leaving the reactor the product stream is degassed and the liquid is separated into an aqueous phase and an oil phase. The aqueous phase is recirculated through the reactors and oil is available for market or further processing. The gas is used internally in the process to provide the thermal energy. The basic process flow is shown in the following figure.

Figure 1: Hydrothermal Liquefaction Process Flow



Input

The process inputs are biomass, water, and an alkali catalyst. Some electricity is required to operate the process.

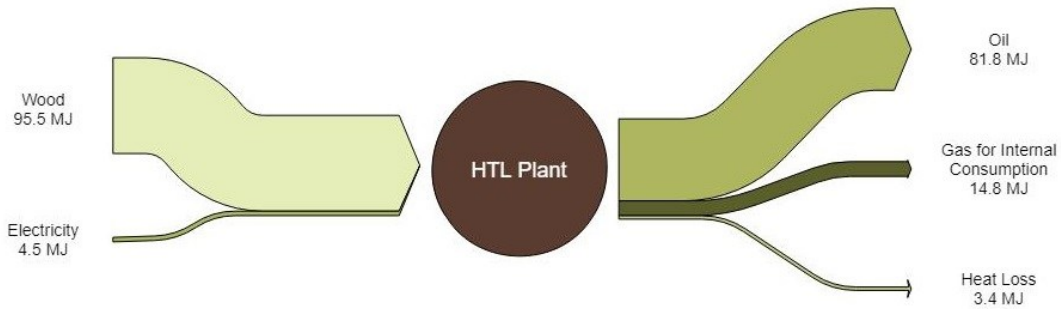
Output

The primary output from the system is the bio-oil and water that is recycled through the process. There is some gas produced in the reactors but this is consumed in the process to provide the process heat to maintain the reactor operating temperature.

Energy Balance

The energy balance of the process is shown in the following figure. This energy balance is based on laboratory scale operations and could change with feedstock and operating conditions. A similar energy balance but without the electrical input was reported by Jensen [5].

Figure 2: Hydrothermal Liquefaction Energy Balance



Typical Capacities

Steeper Energy, one of the companies developing the technology is proposing that its first commercial plants will produce 2,000 bbls or oil per day (140 MW) and consume 240,000 odt (oven dry tonne) of feedstock per year [6].

Zhu [7] undertook a techno-economic analysis of an HTL plant and upgrader that processed 2,000 odt per day and produced 3,780 to 4,900 bbls of oil per day. The low end of the production range represents the state of the art according to Zhu and the high end of the range, the ultimate design goal.

Regulation Ability

There is little published on the performance of continuously operated plants. Given the high pressure and temperatures required in the reactors and the required reactor residence time it is likely that the performance will be altered when the process is operated at rates below the design capacity.

Space Requirements

The space requirements are likely to be dominated by the area required for feedstock storage. None of the techno economic analyses found in the literature report the area required for the plants. There has been no large scale, similar technology plants built. The best estimate is that the area required would be similar to pulp mills that have the same feedstock inputs.

Advantages/Disadvantages

The primary attractiveness of the technology is that biomass can be converted into a liquid fuel. The HTL process produces a liquid with a low oxygen content compared to pyrolysis processes and may be suitable for use as a fuel that can substitute for heavy fuel oil in applications such as the marine sector. The low oxygen content will also make the fuel easier to hydrotreat to remove the oxygen and upgrade the fuel so that it is suitable for more demanding applications such as the use in transport diesel applications. The product could also be used as a bio crude oil for use in existing petroleum refineries to produce bio-gasoline and diesel blending components that are available after fractionation.

Environment

The GHG emission performance of the product is expected to be very good due to the limited fossil fuel inputs into the system and the high oil yield. The high oil yield is also attractive where biomass availability is limited. The low oxygen content of the product will also mean that if it is upgraded through hydrotreating, the hydrogen requirements will be modest and the GHG emission profile will still be good.

Research and Development Perspective

Hydrothermal liquefaction of biomass to produce a bio-oil is a category 1 technology, a technology in the research, development and deployment stage. There is significant uncertainty with respect to the performance and costs of the technology. There is potential to improve yields and reduce costs as more experience with the technology is gained from a demonstration facility and then the technology is scaled to commercial plants.

Examples of Market Standard Technology

Since this is a Category 1 technology there are no market standard technologies. One of the leading development companies with this technology is the Danish-Canadian company Steeper Energy. They have recently announced plans [1] for an industrial scale demonstration plant at a former pulp mill located in Tofte, Norway with their partner Silva Green Fuel, a Norwegian-Swedish joint venture.

Steeper Energy

Sandbjergvej 11
DK 2970 Hørsholm
Denmark

http://steeperenergy.com/

The other major technology supplier is the Australian company, Licella. They have formed a joint venture with the Canadian pulp and paper company Canfor and announced plans to build a 400,000 bbl/day facility in Prince George, BC, Canada [8].

Licella Pty Ltd

Level 7, 140 Arthur Street
North Sydney NSW 2060 Australia

http://www.licella.com.au/contact/

Predication of Performance and Cost

Given the early stage of development predictions of performance and cost are uncertain. The announcement of an industrial scale demonstration facility is a significant step in the development of the technology.

There is some discussion in the literature [7] of the current state of the art with respect to performance, capital costs and operating costs. These estimates have been developed by independent third parties and not by process developers.

Uncertainty

There is significant uncertainty with respect to performance and economics due to the stage of the development of the technology.

Additional Remarks

There is significant interest in this pathway from petroleum refiners. Due to the low oxygen content of the product, there is a potential for the product to be co-processed in existing refineries with minimal modifications.

HTL oil is very viscous which will limit the potential applications of the oil to a replacement for other heavy viscous oil. The physical properties of HTL can be improved by hydrotreating the oil.

Jensen et al [9] have reported that about 2% wt. hydrogen is consumed when the oil is hydrotreated. Hydrotreating yields were about 80% on a mass basis, and over 90% on an energy basis. Hydrotreating also produced 9% water and 5-6% gas. The properties of the HTL and the hydrotreated HTL are shown in the following table. The hydrotreated HTL properties will vary with the severity of the hydrotreating.

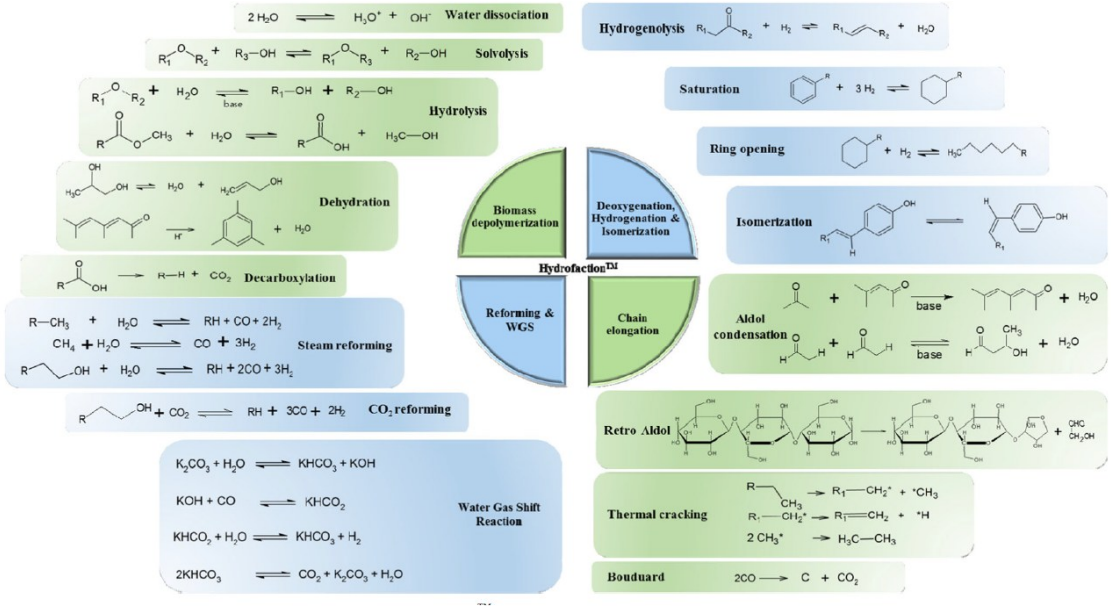
Table 1: Hydrotreated HTL Properties

	HTL	Hydrotreated HTL
Density (kg/m3)	1103	989
Viscosity at 20°C (cP)	80,432	297
HHV (MJ/kg)	37.2	42.1
Carbon (wt. %)	80.6	88.1
Hydrogen (wt. %)	9.1	11.9
Oxygen (wt. %)	10.1	0.0
Nitrogen (ppm)	1500	1175
Sulphur (ppm)	309	389
TAN (mg/g)	55.7	0.0

Quantitative Description

The reaction chemistry of hydrothermal liquefaction is complex, and many different chemical reactions may proceed depending of the specific operating conditions. Jensen et al [2] have proposed a number of the major chemical reactions involved in the process and these are shown in the following figure.

Figure 3: Major Chemical Reactions in the Steeper Energy Process



Typical Plant Size

The technology has not yet moved beyond the pilot plant stage. Steeper and Licella are planning commercial plants in the range of 1,000 to 2,000 bbls per day. Techno-economic assessments [7]

have considered plants of 5,000 bbls/day as long-term design objective. These different plant sizes are shown in the following table with different approximate equivalent metrics.

Table 2: Typical Plant Sized

Volume Based, bbl/day	Input Mass Based, tpy	Energy Output Based, GJ/day
1,000	120,000	6,100
2,000	240,000	12,200
5,000	600,000	30,500

Input and Output

The input and output for a 1,000 bbl/day plant has been extrapolated from Steeper published information [4] and summarized in the following table. In addition to the parameters shown in the table there would be 420 GJ of produced gas that is utilized in the process.

Table 3: Typical Input and Output

	Input	Output
Wood, tonnes/day	350	
Power, kWh	93,325	
Bio-oil, bbls/day		1,000 Bbls

Forced and Planned Outage

Harris Group prepared a report on HTL reactor design for the National Renewable Energy Laboratory (NREL) and used a 90% on stream design factor. That allows 36 days per year for system maintenance.

The combination of a solid feedstock and the high pressures involved have been problematic for systems like pressurized biomass gasification systems. Harris reported that primary challenges associated with the reactor section design were (1) maximizing heat integration, (2) managing the potential for poor heat transfer from the reactor effluent to the reactor feed due to the potential for high viscosities in the feed streams, and (3) minimizing cost associated with the reactor system itself, given the very high required pressures. It is possible that the first large scale demonstration plants will experience some forced outages as these challenges are addressed.

Technical Lifetime

Plants of this type would normally be designed for at least a 20-year lifetime.

Construction Time

Construction periods of 2 years are likely due in part to the fabrication time of the specialty high pressure reactors involved in the process.

Financial Data

Several sources of detailed information [7] [10] [11] [12] on capital costs and operating costs are in the public domain. These have mostly been generated by independent third parties and not by the leading process developers. However, Pedersen et al [12] are involved with Steeper Energy so

the information from this paper has been used in the data sheet. The information from of the reports is discussed below.

Investment Costs

Zhu et al [7] and Pedersen et al [12] provided information for both the production of HTL and for upgrading the HTL to a pure hydrocarbon. The capital costs are compared in the following table. The Harris Group [10] investigated five different designs with widely varying capital costs due to trying to address the design challenges identified above.

Table 4: Capital Cost Comparison

Parameters	Zhu	Harris	Pedersen
Base Year for Costs	2007	2011	2016
Feedstock, tpd	2000	2000	500
Total Capital Investment	$512 Million	$222 to $1,646 million	$225
Best case		$364 million	

The cost estimate by Zhu et al was prepared using Aspen Process Economic Analyzer includes an allowance for potential missing equipment in the design. The estimate by Harris Group is an engineering estimate based on vendor quotes for the equipment. Both estimate approaches use factors for the installation costs and indirect construction costs.

Using the estimate by Pedersen et al and an output of 1150 bbl/day the capital costs are $3.40/annual litre (€3.5/annual litre for a plant in Europe).

Operating and Maintenance Costs

The operating and maintenance costs were estimated in the studies. The results are compared in the following table. The report by Zhu et al does not separate the operating costs for the upgrading from the HTL production.

Table 5: Operating and Maintenance Costs

Parameters	Zhu	Harris
Feedstock	$46 million	Not reported
Waste disposal	$25 million	Not reported
Utilities	$8 million	$22 million
Fixed costs	$24 million	$10 million

The operating and maintenance costs from the two reports have significantly different profiles. Pedersen provided a more detailed breakdown of the variable operating costs. The data was presented per litre of gasoline equivalent (LGE). Here, the data is converted to a per-litre of HTL basis by assuming that 1 LGE is equal to 0.90 litres of HTL. The information is shown in the following table.

Table 6: Operating and Maintenance Costs

Parameter	US$ per Litre HTL
Feedstock	0.13
Hydrogen	0.11
Thermal energy	0.24
Power	0.05
Wood Grinding	0.058
Water disposal	0.014
Fixed O&M	0.10

The variable O&M costs less feedstock, power and thermal energy is 0.18 US$/litre (0.15 €/litre) and the fixed O&M is 0.10 US$/litre (0.08 €/litre).

Start-up Costs

The start-up costs are included in the operating cost estimates.

Technology Specific Data

The properties of the Hydrofaction™ oil produced by the Steeper Energy process are shown in the following table [2] and compared to typical values for heavy fuel oil [3]. The bio-oil has some oxygen which reduces the energy content and increases the density but the properties are broadly similar. The bio-oil viscosity is very sensitive to the temperature and is more viscous that the petroleum HFO at the same temperature.

Table 7: 55 HTL Bio-Oil Properties

Parameter	Hydrofaction™ Oil	Heavy Fuel Oil
Energy Content, MJ/kg (LHV)	36.7	39
Water Content, wt. %	0.8	0.1
Density, kg/m^3	1.05	0.98
Oxygen Content, wt. %	9.8	0
Pour Point, C	24	15
Flash Point, C	29	100
Kinematic viscosity at 40 °C, mm^2/s	17,360	
Kinematic viscosity at 50 °C, mm^2/s		200-600
Kinematic viscosity at 60 °C, mm^2/s	1,545	

Data sheets

The information on HTL production is summarized in the datasheets. Since there are no operating plants in 2015, the data is presented for the years 2020 to 2050.

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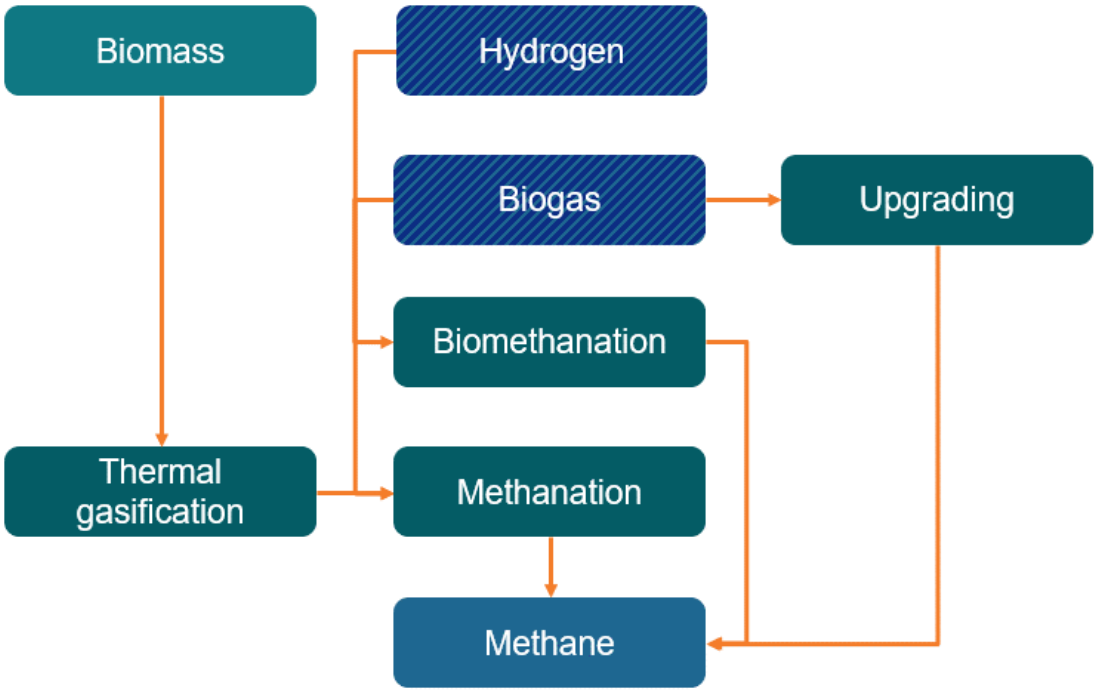
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Methane



This section deals with the production of methane. The inputs are the intermediate outputs from biogas plants, hydrogen from electrolysis or biomass through thermal gasification. The produced methane can be used directly or loaded on the gas grid.

The process of upgrading biogas refers to the removal of CO_2 from the biogas, while in methanation and biomethanation hydrogen is added to react with the CO2 in biogas. Finally, the syngas from thermal gasification can also be methanated with or without hydrogen. The subchapter here deals with methanation without hydrogen addition.

The section has the following chapters:

91 Biogas Upgrading (CO2 Removal)

92 Biogas Methanation (Hydrogen Addition)

93 Biogas Biomethanation (Hydrogen Addition)

94 Syngas Methanation (from Thermal Gasification)

91 Biogas Upgrading (CO_2 Removal)

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Publication date

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Amendments after publication date

Date	Ref.	Description
08/2023		Parameter name correction in datasheets for Biogas Upgrading (no change of values)
06/2023		Extensive update of chapter on data, large- and small-scale plants

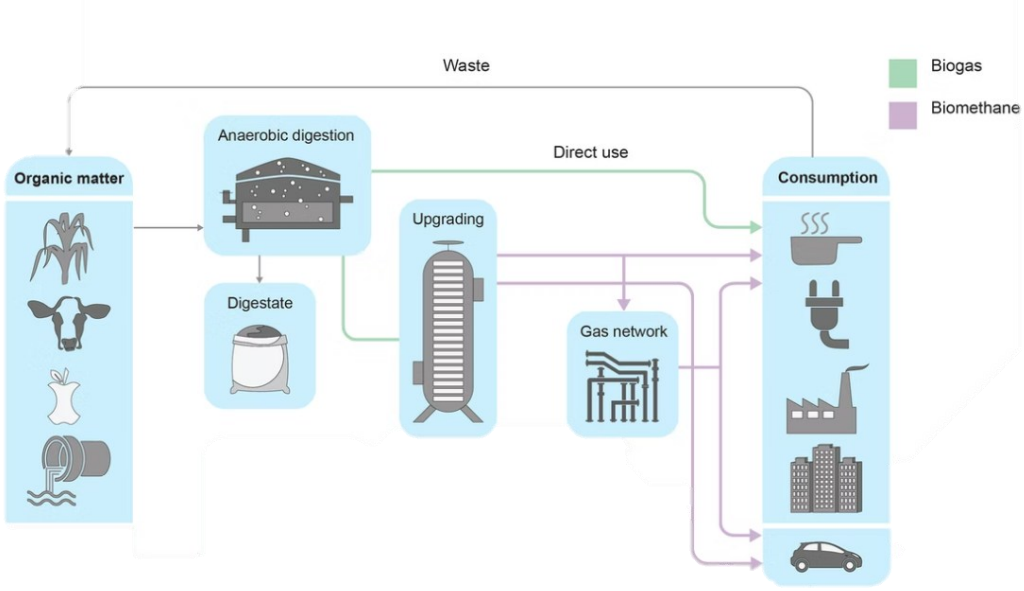
Qualitative description

Brief technology description

Biogas can be utilized as a renewable energy source. For instance, to produce heat and electricity or as vehicle fuel. For some applications where it is important to have a high energy content in the gas, e.g., as vehicle fuel or for grid injection, the gas needs to be upgraded. Upgrading biogas refers to the process of removing CO_2 to obtain a gas with a high methane content, known as biomethane. Today, most Danish biogas is upgraded to biomethane and distributed via the national gas grid. Almost all new biogas plants are expected to upgrade the biogas, resulting in almost 80% of all biogas produced being upgraded by 2030 [1]. It is likely that in the future, all or a portion of the biogas at some biogas plants, will be used for producing liquefied biogas (LBG), methanol or other Power-to-X (PtX) fuels, however, these conversion technologies are not covered by this technology chapter.

Figure 1 shows how biogas can either be upgraded through an upgrading plant to be fed into the gas grid or be directly applied for consumptions needing gas with high energy content.

Figure 1: Biogas uses



Source: Illustration from IEA. Ann.: Biogas may either be used directly as an energy source or be upgraded through an upgrading plant to be fed into the gas network or be directly applied for consumption needing gas with high energy content.

Biogas is converted into biomethane through a purification process at an upgrading facility. Biomethane has similar properties to conventional natural gas [3]. The input for upgrading facilities is raw biogas from an anaerobic digester, which typically contains 50-75% methane (CH_4) and 25-50% carbon dioxide (CO_2), plus a minor content of hydrogen (H), nitrogen (N), oxygen (O), hydrogen sulfide (H_2S) and ammonia (NH_3). The composition of the biogas varies based on the specific mix of the input.

Before injecting the gas into the gas grid, it is necessary to remove the content of CO_2, thereby increasing ("upgrading") the heating value of the gas. Depending on the composition of the raw biogas, it may also be necessary to remove water moisture, particles, H_2S, NH_3 and N_2. As it is rather expensive to remove N_2, this is rarely done. H_2S needs to be removed before further use as it is a corrosive gas.

Upgrading can also take place by catalytic conversion of CO_2 to CH_4 by adding hydrogen or through bio-methanation. Biomethanation for upgrading can be done by two methods, in-situ or ex-situ processes. The in-situ process involves adding H_2 to the processes in the biogas digester to obtain an output gas from the biogas process with higher CH_4 content. However, this method produces an inadequate concentration of methane in the off gas for addition to the gas grid combined with a slip of hydrogen. In the ex-situ process, the off gas from the biogas digester is further processed in a subsequent (trickle bed) reactor in which very high methane concentration (> 95%) can be achieved with low hydrogen slip. These technologies, which are less mature than traditional upgrading through CO_2 removal, are not addressed in this technology chapter.

During upgrading, a stream of CO_2 is produced. Up until recently this CO_2 has usually been vented into the air but is now increasingly being sold for the purpose of storage or utilization, thus representing an additional source of income. This should be considered when the economy of the plant is assessed.

Danish quality requirements for biomethane are described in BEK no 230 of 21/03/2018 under the Danish gas legislation [4]. It should be noted that gas quality in European countries is regulated at the national level, meaning that some countries have more strict gas quality

specifications than others. This can hinder the physical trading of gas between countries' gas systems, or a need for treatment between borders. This chapter assumes that the biogas is upgraded to a quality that is compliant with Danish quality requirements.

The heating value of natural gas depends on the content of propane in the natural gas. Natural gas with a high content of propane has a higher heating value (about +10%) than biomethane, which causes challenges for example concerning proper billing of customers. One approach to solving the problem is to add propane to the upgraded gas and thereby increase the heating value. This method is e.g., used in the UK and Germany [5, 6]. Propane addition is associated with considerable costs. The Danish gas distribution companies have therefore decided to solve the problem through measurements of the gas quality rather than adding propane. By connecting the upgrading plants to the gas grid at metering and regulating (MR) stations, gas companies can keep track of the gas quality in different parts of the distribution network and thus ensure proper billing of customers. Therefore, costs related to propane addition are not considered in the technology data sheets.

Typically, the investment costs for a complete upgrading system connected to a gas grid can be categorized using the following main components excluding the biogas plant itself,

The upgrading plants
Additional equipment to treat methane slip (where necessary)
Compressor units (where necessary)
Grid connection plant
Facilities for carbon storage or utilization

The upgrading plants

The main purpose of the upgrading plant is the removal of CO_2, to end up with a gas with a higher methane content (biomethane). The capacity of the upgrading plant is usually stated in Nm^3 CH4/year.

All Danish upgrading plants are based on one of the three following technologies:

Amine scrubbing
Water scrubbing
Membrane separation

Besides these technologies, pressure swing adsorption (PSA) is a widely used method worldwide. Other technologies – such as cryolitic separation and enzymatic upgrading – are under development. However, there is currently little data available on these technologies.

Today, the most common upgrading technology in Denmark is amine scrubbing, which is used for ¾ of all biomethane production (by volume), followed by water scrubbing and membrane separation [7]. The other technologies are not currently in operation in Denmark. Membrane separation is only used for small biogas plants, while water scrubbing is expected only to be used for new biogas plants to a limited extent. Accordingly, amine scrubbing is expected to continue as the most applied technology, especially for large biogas plants, which are and will be producing the majority of the biogas looking forward. Therefore, the calculations in the data sheets in this technology catalogue are based on the amine scrubbing technology.

The following sections give a brief introduction to the different upgrading technologies.

Amine scrubbing use amines that chemically bind to the CO_2 and H_2S molecules, removing them from the gas. The amine is regenerated in a stripper where the CO_2 is removed from the amine solution by adding heat. This process has the highest efficiency in terms of methane conservation.

To regenerate the amines, the process uses temperatures between 120-150 °C, which typically is produced by the combustion of natural gas[3] or biomass. Cheap heat sources and reuse of the heat are therefore key parameters for amine scrubbing to be economically competitive. Today amine scrubbing plants reuse at least 40% and up to more than 80% of the heat. Further, as the process happens under a pressure between 1-3 bar, there is a demand for electricity to run a compressor ensuring sufficient pressure for the gas to be injected into the distribution gas grid (4/7 bar).

Water scrubbing was the first upgrading technology used in Denmark. In a water scrubber, the absorption process is purely physical. The biogas is put in contact with water by spraying or bubbling through to wash out the CO_2 but also H_2S since these gases are more soluble in water than methane. When injecting biomethane into the gas grid, the gas needs to be under the same pressure as the grid gas. The pressure in water scrubber plants is high enough (around 6 bar) for the gas to be directly injected into the distribution grid meaning that no further compression is necessary for grid injection of the biomethane. The major reason why water scrubbing is opted out is the methane loss, which is approx. 1% of the biomethane production.

Membrane separation is a process where membranes, which consist of hollow fibres, separate the carbon from the biogas. Components such as water and H_2S are likewise separated through the process. The membranes are permeable to ammonia, water, and CO_2. Nitrogen and methane only pass through the membrane to a very low extent while oxygen and hydrogen sulphide pass through the membrane to some extent. Typically, the process is carried out in two stages. In the first step, before reaching the membranes the gas passes through a filter that catches water and oil droplets that would otherwise affect the efficiency of the membranes. Besides that, H_2S is typically removed using activated carbon. The CO_2 is removed from the gas in the membrane in the second step. The advantages usually presented for membrane technology are the lack of demand for water or chemicals and the ability to scale down the process without large efficiency losses. Based on number of plants, membrane separation is the most widely applied technology in Europe [8]. However, the technology has, relative to the scrubbing technologies, a high OPEX, making it expensive especially for larger plants.

PSA is among the most widely practised upgrading methods worldwide but is not used in Denmark, mainly because of the process' relatively high methane losses. It separates some gas components from a mixture of gases under high pressure according to the molecular characteristics of the components and the affinity for an adsorbent material (often active carbon). The process then swings to low pressure to desorb the adsorbent material.

Relative costs

Figure 2 provides an indication of the specific costs per Nm^3 biomethane production for the water scrubber, amine scrubbing, and membrane separation at plants of different sizes. It should be recognized that in reality costs will depend on the specific circumstances of the plant, electricity prices, options for energy integration with the biogas plant itself, etc. The price estimations from [9] are from 2020 and there has been a general increase in prices since then. However, it is expected that the prices of the different upgrading technologies have been affected equally, implying that relative costs remain unchanged, and that prices eventually will return to prices similar to the ones stated below.

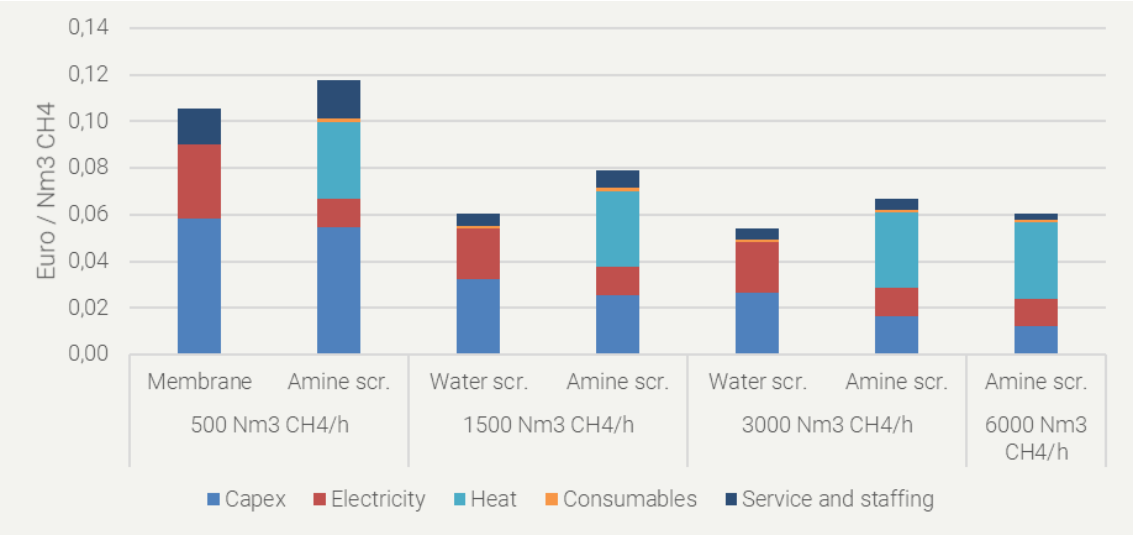
The graph in Figure 2 shows that membrane separation is cheaper than amine scrubbers for small scale plants. For larger plants (1,500-3,000 Nm^3 CH_4/h), water scrubbers are cheaper than amine scrubbers, although the costs for amine scrubbers are seen to decrease compared to small scale plants. Comparing the costs across the different upgrading technologies there are

[3] Natural gas is often preferred over biogas in Denmark due to tax regulation.

seen to be significant differences in the relative importance of different cost components. The high share of costs for the amine scrubber attributable to heat implies that reusing process heat is an important determinant of the plants' economy.

The cost figures include regenerative thermal oxidation (RTO) to treat the methane loss of around 1% at the water scrubbing plants, however, the value of the loss of biogas is not considered in the comparison.

Figure 2: Comparison of the cost of upgrading biogas



Source: Calculations by Ea. Energy Analyses based on [9]. Ann.: (excluding the cost of the biogas plant) using different upgrading technologies at plants of different scales. The calculation assumes 50% reuse of the waste heat (assuming 80% reuse the amine scrubbers are the cheapest per Nm3). The results are based on a private sector stakeholder perspective assuming a 10% discount rate. In the calculations, the price for electricity is 6.3 Euro/100 kWh.

Additional equipment to treat the methane slip

The waste gases from an upgrading plant contain methane in a small concentration, but seldom enough to maintain a flame without the addition of natural gas or biogas. One way of limiting the methane slip is to oxidize the methane by regenerative thermal or catalytic oxidation.

The need for off-gas treatment depends on the methane slip from the specific plant. Plants using water scrubbing technology or PSA technology would often require tail-end solutions to decrease the methane slip – this is not included in the cost in the graph above. Therefore, water scrubbing will be costlier when the methane emission must be kept below 1%, which is the target value for the biogas industry (for the total process including the biogas plant).

For amine scrubbers, the methane loss in the process is very low (about 0.1%) and consequently, off-treatment is not required.

Compressor unit and grid connection plant

In conjunction with the gas treatment plant, a grid connection facility should be established. For larger upgrading plants the local distribution network will in many cases not be able to take all the produced gas at all seasons. In these situations, the gas needs to be further pressurised from 4-7 bar to 40 bars, to be fed into the natural gas transmission network. In addition, measurement regulation equipment is required. Further to this, but not included in the datasheet costs, is the connection pipeline to the gas grid.

Included in the datasheet

Upgrading of biogas (incl. H_2S removal)
Measurement regulation
Odorization equipment
Pressurizing to 4-7 Bar

Not included in the datasheet

Pressurizing from 4/7 to 40 Bar
Connection pipeline to the gas grid

Carbon capture

During upgrading a stream of CO_2 is produced, which can potentially be sold for carbon capture and storage (CCS) or utilization (CCU) as an additional source of income. This should be considered when the economy of a plant is calculated. According to Biogas Denmark, CO_2 is seen to be a significant new source of income from biomethane production, which is expected to cause a decrease in the price of biomethane.

To utilize the CO_2, it needs to be purified, dried, compressed, and possibly cooled so that it can be transported via pipeline infrastructure or in liquid form with a truck to storage. When the biogas is upgraded in an amine scrubber, the CO_2 flow has high purity. This means that in contrast to the other mentioned alternative upgrading technologies, additional purification steps are not required to utilize the CO_2 [10].

If the CO_2 is to be transported by trucks, an intermediate CO_2 storage facility will be required at the biogas facility. At Korskro biogas plant, the first in Denmark to utilize the CO_2, the intermediate CO_2 storages have been designed to store approx. 1% of the annual amount of CO_2 generated at the plant. The intermediate storage consists of three large tanks, which each has a capacity of 50,000 liters [11].

It has been a common assumption that if large amounts of CO_2 are to be transported, it must be done at high pressure and thus in liquid form. However, screening and analyses carried out by Evida indicate that transporting CO_2 in gaseous form and thus at lower pressures is a realistic alternative – also for larger amounts of CO_2 of, for example, several million tons per year [12]. Transporting CO_2 in gaseous form will reduce the costs for pressurization. At the same time, it will technically be possible to convert existing pipelines to transport CO_2, if they are not used for methane or hydrogen, which will further lower the cost per ton of CO_2 transported. CO_2 assumes gaseous form at pressure up to about 40 bar(g). At pressures above 80 bar(g), CO_2 is liquid. In the range between 40 and 80 bar(g), CO_2 will be partly liquid and partly gaseous [12].

It is expected that many upgrading plants will invest in facilities to capture and store CO_2 as it is seen as a source of income. CO_2 can be used in PtX, CCS, the food industry, etc. According to the biogas industry, currently, CO_2 is in short supply because of increasing demand. With large expectations for producing carbonaceous e-fuels such as methanol towards 2030 and onwards in which CO_2 is a key component for production, the demand is expected to continuously increase.

In addition to CCS and CCU, other options for utilizing the CO_2 stream from biogas plants are developing. One way is methanation, where H_2 is added and reacts with the CO_2 in the biogas to convert it to more CH_4. This can substantially increase the amount of CH_4 produced per unit of biomass and reduces the cost of amine scrubbing used in conventional upgrading. It is traditionally done using a catalyst, described in chapter 92 Biogas Methanation (Hydrogen Addition) in this catalogue. An alternative method that is currently being investigated is biological methanation, or biomethanation, where microorganisms suspended in a liquid or fixed to a

reactor bed consume CO_2 and H_2 as part of their metabolism, resulting in the production of CH4 as a by-product.

Input

Raw biogas from a biogas plant.
Heat (or electricity depending on the technology) for the upgrading process.
Electricity for compression.
Smaller amounts of water and various chemicals.

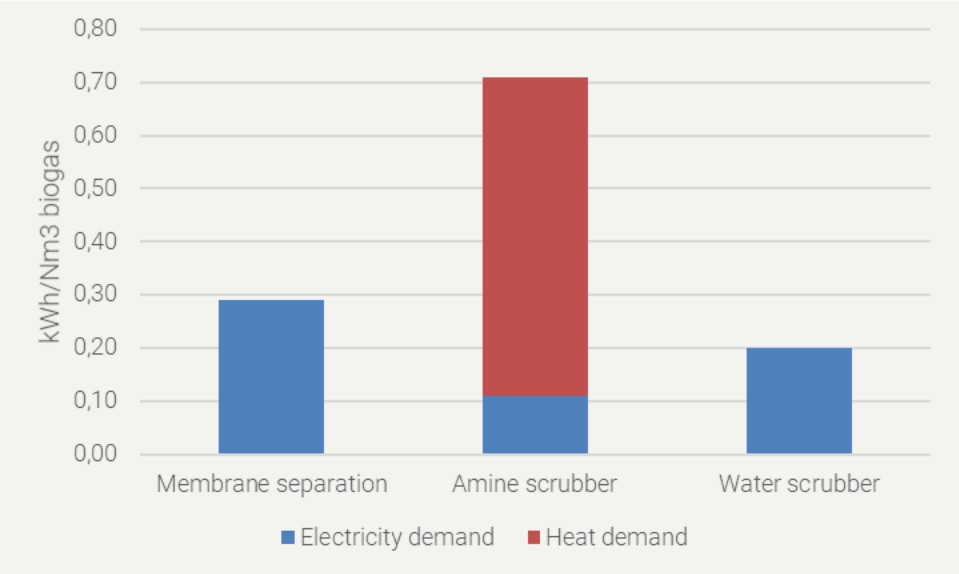
Output

Upgraded biogas with 95-99 vol. % methane, CO_2, nitrogen, and oxygen [7].
Waste gas containing mostly CO_2

Energy balance

As shown in Figure 3, the power consumption of the upgrading processes varies, but it ranges from approximately 0.12 to 0.3 kWh/Nm^3 biogas. The amine scrubber has a heat demand of around 0.6 kWh/Nm^3 biogas, but a lower electricity consumption than the other upgrading technologies. The heat for the amine scrubber should be supplied at 120- 150°C and 80% of the heat can be reused. A third of the reusable heat can be used for high-temperature purposes, while the rest only can be used in low-temperature (65°C or below) applications unless a heat pump or broiler is used to boost the temperature.

Figure 3: Energy consumption (kWh/Nm3 raw biogas) for the upgrading technologies, water scrubbing, amine scrubber and membrane



In the upgrading process, there is typically a methane slip between 0.1-1%; thus, depending on the upgrading technology, 99-99.9% of the inlet methane exits as a product [13].

Typical capacities

Different upgrading facilities are available from several suppliers in a broad range of capacities.

As of October 2022, 58 biogas plants supply biomethane to the gas grid in Denmark. The capacities of upgrading plants follow the development of the size of biogas plants, where the size has increased over the last years. Upgrading facilities for new state-of-the-art plants are expected to have capacities of 25-50 mil. Nm^3 CH_4 per year (equivalent to 3,000-6,000 Nm^3 CH_4/h).

Membrane-based upgrading facilities are generally used for smaller plants (the largest plant has a capacity of 1,300 Nm3 CH_4/h), while water scrubbing is primarily used for medium-sized plants. Amine scrubbers are available in all sizes and are especially prevalent in the market for large-scale biogas plants [9, 13].

Regulation ability

Biogas upgrading plants can down-regulate to 20-30 % of the full load. Cold start-up time is about 2-4 hours, whereas hot standby (maintaining the plant at operating temperature) takes less than 0.5 hours.

Advantages/disadvantages

Upgrading of biogas to biomethane and injection in the gas grid makes it possible to decouple demand and production, as the upgraded biomethane can be transported and stored in the central gas grid and used where and when needed throughout Europe.

A disadvantage is the energy requirement and the relatively large investments connected with the upgrading.

The different upgrading technologies each have advantages and disadvantages respective to each other. For the most common technology, amine upgrading plants, the following are the most significant advantages and disadvantages.

Advantages:

This process has the highest efficiency in terms of methane conservation, which is important from an economic, but also environmental and climatic, perspective.
Low operational expenses, OPEX, making it especially economically attractive for larger plants
The technology is offered commercially by multiple vendors in a large capacity range

Disadvantages:

The process uses temperatures between 120-150°C. The relatively high energy demand for heating is therefore the biggest disadvantage of amine scrubbing. Cheap heat sources and reuse of the heat in the process are key parameters for amine scrubbing to be economically competitive
The process happens under a pressure between 1-3 bar, implying that there is a demand for electricity to run a compressor ensuring sufficient pressure for the gas to be injected into the gas grid (4/7 bar).

Environment

Besides the energy consumption for operation, biogas-upgrading technologies have two other major environmental issues depending on the technology: the consumption of water and chemicals and methane leak/emission.

Only the water scrubber and the amine scrubber use water – respectively 0.4-4 l/Nm^3 and 0.03 l/Nm^3 raw biogas. The chemical consumption for the water scrubber and amine mainly consists of anti-foaming. The removal of hydrogen sulphide requires activated carbon for both PSA, membrane separation and amine scrubbing. The highest reported chemical requirement of activated carbon is 0.08 grams/Nm^3 of raw biogas. Furthermore, the amine scrubber has a demand for amine to account for the loss of amines in the process. During normal operation, only minor amounts of amine are lost [2]. High concentrations of amines disposed to the environment may cause issues for the aquatic environment [14]. However, the risk can be reduced or eliminated by using the right techniques. It is therefore not assessed to be a noteworthy environmental issue for upgrading plants.

The highest methane slip among the technologies is reported from PSA (1.8%-2%), followed by the water scrubber (1%), membrane separation (0.5%) and the lowest slip is from amine scrubbing (0.1%). In principle, psychical scrubbers have a higher slip than the other technologies, but the methane is utilized internally. Methane is the second most important GHG contributor to climate change following CO_2. On a 100-year timescale, methane has 28 times greater global warming potential than CO_2 per kg. [15]. Keeping the methane emission from biogas upgrading plants to a minimum is therefore of great importance for the climate impact of biomethane production. The methane slip can be eliminated if the off-gas from the upgrading plant is treated in a regenerative thermal oxidation (RTO) plant. Other solutions for reducing methane leaks and increasing the output of upgraded gas are developed. This is especially relevant for water scrubber plants, as it will increase the competitiveness of the technology.

Research and development

As noted above it is expected that the research and development and the competition between the different upgrading technologies will lead to incremental continuous improvements in the technology and, to some extent, a reduction of costs.

Cryogenic upgrading

There may be a future potential for the development of cryogenic treatment for upgrading biogas and for the condensation of upgraded biomethane to liquefied biogas. However, today the technology deployment is limited by operational problems.

Compared to other upgrading technologies cryogenic upgrading may have a lower energy demand, no contact between gas and chemicals, production of pure CO_2 as a side product, the possibility to produce liquefied biogas (LBG) and removing nitrogen from the gas stream.

Enzymatic upgrading

Enzymatic upgrading process is anticipated to be more energy-efficient and cost-effective than commercially available upgrading technologies. In enzymatic upgrading, the CO_2 is captured in a non-volatile solvent with a biocatalyst in an absorber column. The biocatalyst accelerates CO_2 absorption using enzymes. Afterwards, the CO_2 is removed from the solvent in a stripper column. The technology integrates enzymes to create an industrial biocatalyst that can be readily incorporated into conventional chemical absorption processes for CO_2 removal.

Due to the anticipated potential, both Energiforsk [16] and The Danish Energy Technology Development and Demonstration Program (EUDP) [17] have supported the development and testing of enzymatic upgrading technology Both projects were however closed down prematurely, in 2015 and 2017, and have therefore not demonstrated upgrading of biogas using the enzyme-based upgrading technology.

Examples of market standard technology

Vinkel Bioenergi, 2019, 52 mill. Nm^3 biomethane per year – appx. 6,000 Nm^3 per hour. Applies amine upgrading technology. https://lundsbybiogas.dk/da/case/vinkel-bioenergi/

Vesthimmerlands Biogas, 2020, 35 mill. Nm^3 biomethane per year – appx. 4,000 Nm^3 per hour. Applies amine upgrading technology. https://www.vhbiogas.dk/fakta

Linko Gas, 2016 + 2019, 19 mill. Nm^3 biomethane per year – appx. 2,200 Nm^3 per hour. Applies water scrubbing upgrading technology. https://www.linkogas.dk/hoveddata/

Assumptions and perspectives for further development

This section mainly focuses on the learning perspective for the amine scrubber technology since this technology dominates the market today. However, it cannot be ruled out that one or more of the other upgrading technologies will become more competitive. The development potential within these technologies is therefore also touched upon.

The technical options for improving the amine scrubber technology lie within the following areas [18]:

- development of new and improved amine solvents that:
    - require less energy for regeneration,
    - have a higher cyclic capacity (smaller equipment),
    - are more resistant to degradation
    - can provide a CO_2 stream at a higher pressure and thus reduce the need for subsequent compression
    - improved thermodynamical integration at the facility
    - use of high-temperature heat pumps for the supply of hot water for amine upgrading
    - development of radically different solvents e.g. non-aqueous solvents, yet this is very uncertain
    - development of process equipment optimized for upgrading facilities

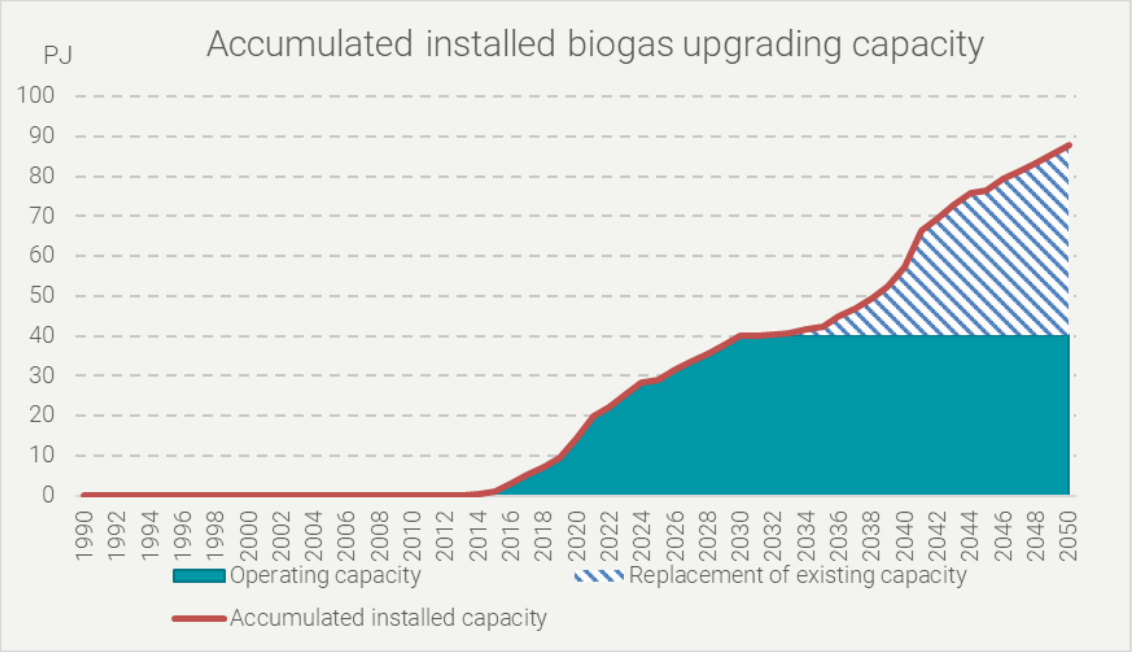
Amine scrubbing is a mature technology, which has been used since 1930 to separate CO_2 from natural gas and hydrogen [19]. However, as suggested above the technical opportunities for improving the technology are plentiful. In the future, the application of amine scrubbers for CO_2 capture at power plants and large industrial point sources could provide a major driver for improving the technology.

In the Danish Energy Agency's catalogues for Carbon Capture, Transport and Storage the investment cost and fixed O&M of post-combustion amine-based scrubbers are expected to drop 18% by 2030 and 39% by 2050. Disregarding differences in plant sizes and input gas composition, it is reasonable to assume that a cost reduction of a similar level can be expected for amine scrubber plants used for upgrading biogas.

The cost reductions can be expected to take place through technological innovation that will affect global markets, for example, the development of new and more cost-efficient amine solvents that may be used at CCS plants and for biogas upgrading, as well as through learning process that relate to the actual application and installation of technologies in the Danish market for upgraded biogas.

In the Danish Energy Agency's Climate Programme 2021 four long-term scenarios are established exploring how Denmark could become climate neutral by 2050 focusing on different mitigation options. Depending on the scenario, biogas production reaches somewhere between 35 and 52 PJ by 2050, i.e. the same level as 2030 or lower. All new biogas production is assumed to be connected to an upgrading facility. Assuming that the installed production capacity remains stable after 2030 there will however be a need for replacement capacity as existing biogas plants reach their technical lifetime of around 20 years, which leads to additional learning. Therefore, by 2050 the accumulated installed (not operating) upgrading capacity would reach around 90 PJ.

Figure 4: Forecast of accumulated installed biogas upgrading production capacity in Denmark



Ann.: Based on Denmark's Climate Status and Outlook 2022 and assuming the need for replacement capacity beyond 2030.

Learning rates for energy technologies typically vary between 5% and 25%. In 2015, Rubin et al. published "A review of learning rates for electricity supply technologies", which provides a comprehensive and up-to-date overview of learning rates for a range of relevant technologies. 10-15% seems to be the typical level for many technologies, with solar PV being an exception demonstrating learning rates well above 20% [20].

Considering the wide range of technical options for the improvement of the technology, it is estimated that a learning rate of approximately 15%, thus at the higher end of the interval, is applicable to amine-based upgrading. This implies a 15% cost reduction for every doubling of the installed biogas upgrading capacity. When applying this learning rate to the forecast of accumulated installed biogas upgrading production capacity in Denmark, the cost reduction compared to 2020 is estimated to 22% and 39% by 2030 and 2050 respectively. These potentials resemble the corresponding cost reduction potentials identified for post-combustion amine-based scrubbers in the Danish Energy Agency's catalogues for Carbon Capture, Transport and Storage.

The cost reduction potentials derived above are applied to both investment costs and fixed O&M in the data sheets. Additionally, in accordance with the Danish Energy Agency's catalogues for Carbon Capture, Transport and Storage, we assume a 14% and 21% reduction in heat input by 2030 and 2050 respectively.

Biogas upgrading by amine scrubbing is a commercial technology with large deployment in Denmark. It is therefore assessed to be a category 4 technology, meaning that the price and performance of the technology today is well known, and only incremental improvements are expected.

Other upgrading technologies

Water scrubbing and PSA have been mature technologies for many years, and only incremental technology development is expected, while cryogenic upgrading is a technology under development and demonstration. One advantage of cryogenic upgrading is the integrated

liquefaction, which is particularly attractive if the CO_2 is to be stored locally for later sale or for the, in Denmark, rare cases where the biomethane is not injected into the grid but offered as LBG. Improving the energy efficiency of the liquefaction process will be key to improving the overall performance of the process [21].

Datasheets

The datasheest contains data for upgrading plants of varying sizes, one of which is capable of upgrading approx. 25 million Nm3 methane while the other, a larger plant, can upgrade approx. 50 million Nm^3 methane (equivalent to approx. 3,000 and 6,000 Nm^3 CH_4/hour). It is important to note that the data provided is not directly applicable to smaller plants. For all projected years (2020, 2025, 2030, 2040 and 2050) the datasheets are based on an amine scrubber plant.

The datasheet does not include data for producing biogas, as this information can be found in chapter 81 Biogas Plants. The biogas plants are of similar sizes as the upgrading plants, thereby enabling the linkage of data to provide a comprehensive overview of the entire process from feedstock to upgraded biogas.

In the data sheets methane emissions from the upgrading plant are set to 0.1%, reflecting the upper-bound estimate of methane emissions from upgrading using amine scrubbing technology.

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92 Biogas Methanation (Hydrogen Addition)

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August 2018

Amendments after publication date

Date	Ref.	Description
12/2018		Datasheet revised

Qualitative Description

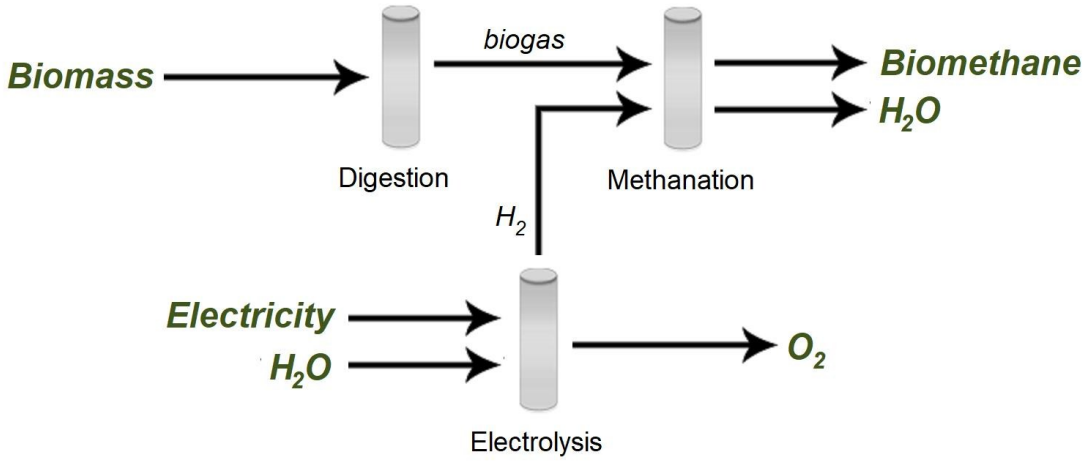
Methane can be produced through the methanation of biogas. The main components in biogas are methane and CO_2. The content of CO_2 may vary between about 35-50 vol. % depending on the actual biogas production technology. The carbon dioxide in the biogas is reacted with hydrogen to produce additional methane. Alternatively, methane can be produced via gasification and methanation of biomass which is described in and the subsequent chapters.

This chapter does not consider the production of the hydrogen as there are multiple production options available. Biogas, hydrogen production, and methanation are all commercial technologies but the combination of the three processes is a concept that is still at the research, development and deployment stage.

Brief Technology Description

The core unit in a methanation plant is the hydrogenation/methanation unit in which CO_2 is converted to methane by reacting with hydrogen. A catalyst is usually used for this process. The methanation reaction is exothermic and will generate large amount of heat. There is also some research and development work being undertaken on biological conversion routes [1]. This section is based on the chemical route.

Figure 1: Methanation Process



The theoretical chemical reaction that occurs with methanation of biogas is shown below.

$$CO_2 + 4\ H_2 \rightarrow CH_4 + 2\ H_2O$$

The reaction is exothermic (releases heat) and the reactors are operated at temperatures below 200°C (at 1 bar) or 300°C (20 bar) to reach conversion rates of greater than 98%.

The catalytic methanation reaction has been known since 1902. The technology has been developed for large scale coal to gas plants. The application of the technology may require some different concepts for the smaller scale and potentially intermittent or dynamic operation.

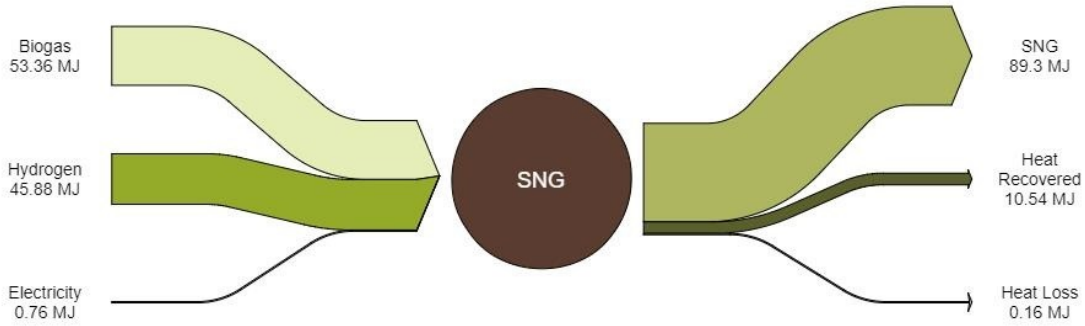
Input and Output

The systems require biogas, hydrogen and electricity to operate. They produce methane suitable for injection into the gas distribution grids, a small amount of water and some recoverable heat.

Energy Balance

The energy balance is presented as a Sankey diagram in the following figure. Note that this is based on hydrogen and the energy required to produce the hydrogen is not included. The data in the following table is based on "process concept 1" in the Swiss Federal Office of Energy report [2]. This is based on the biogas having 65% methane; gas with less methane will require more hydrogen and have a different energy balance.

Figure 2: Energy Balance



The methanation stage is relatively efficient with some high temperature (>100°C) heat recoverable from the process.

Typical Capacities

The capacities of the systems will likely be limited by the size of the biogas plants. Gotz et al [1] reports that in 2016 the largest demonstration plant was the Audi e-gas plant in Wertle, Germany with a hydrogen production capacity of 6 MW. The plant output was 325 NM^3/hr (88,000 GJ methane/year) (3 MW) [3]. Gotz et al undertook their techno-economic analysis on a plant that produced 1000 NM^3/hr (270,000 GJ/year) (9 MW).

The concept could also be applied to other sources of carbon dioxide such as biomass gasification or industrial processes that produce high quality CO_2 such as ethanol production, some natural gas processing plants and processes such as ethylene oxide production [1]. Larger plants may be possible with these systems.

Regulation Ability

Without storage capacity of the biogas and the hydrogen the overall system will have very limited regulation capacity, as both inputs are required for the methanation stage.

The methanation process itself operates at elevated temperatures and load changes can induce unwanted temperature changes in the catalyst beds. Steady state operation will provide the optimum performance.

Space Requirements

The COSYMA (Container Based System for Methanation) pilot plant in Switzerland put the entire methanation process, including compressor and gas cleaning, is integrated into a standard 20 feet shipping container [2]. This was a small system but the space requirements will be less than the space requirements of the biogas facility.

Advantages/Disadvantages

The biogas methanation concept can be considered another variation of the power to energy concept, although any source of hydrogen could be utilized in the process the most likely embodiment of the concept will produce the hydrogen from electricity. The process would essentially double the quantity of low carbon natural gas that can be produced from biogas plants and there is a ready market for the gas in the natural gas distribution systems.

The size of the plants will be limited by the capacity of biogas plants, which tend to be relatively small. Economies of scale that could reduce the production costs may be difficult to obtain. The potential to standardize and produce a number of systems that have an integrated electrolyzer and methanation unit may provide some help in reducing system costs.

Larger systems based on other sources of carbon dioxide may have more favourable economics and the potential to produce greater quantities of methane. The GoBiGas project in Göteborg Sweden was design to produce 72 GJ/hour (576,000 GJ/year) of methane in the first phase and eventually 2.3 to 2.9 million GJ/year in a final phase [4].

Environment

The overall impact on the GHG emissions of the produced methane will be dependent on how the hydrogen is produced. Hydrogen from electrolysis produced with renewable electricity will produce methane with low GHG emissions. Some of the other potential systems which utilize CO_2 from renewable sources may also have good emission profiles but CO_2 from fossil systems will not produce renewable natural gas.

The environmental impact from the biogas production system and from the hydrogen production system will be larger than from the methanation process.

Research and Development Perspective

Methane production from power is a category 2 technology, a pioneer phase technology with limited applications to date. The technology has been proven to work through demonstration facilities or semi-commercial plants. However, due to the limited application, the price and performance is still attached with high uncertainty, since development and customization is still needed. The technology still has a significant development potential.

As noted, the three sub-processes are all commercial processes although the commercial methanation processes are generally at a larger scale than would be required for biogas plants. What is required for commercialization is the system integrators who can integrate hydrogen production and methanation into a system that can be added to the biogas production unit.

Examples of Market Standard Technology

Haldor Topsøe offers methanation technology, their TREMP™ process. The applications of the technology have been in larger fossil-based industries.

The applications of the technology to smaller biogas facilities has been at the demonstration plant level, so no market standard technology has developed yet. There are at least three groups that are promoting their technology demonstrations.

The Paul Scherrer Institute in Switzerland along with their partners energie 360 and erdgas biogas have the Biosweet project that has an operating system at Zurich Biogas [2]. They have published results of their system analyses and demonstration plant performance.

The Vienna University of Technology [5] is offering a system with a single stage methanation step and membrane cleanup of the gas. They claim low capital and operating costs.

HZI Etogas develops and builds Power-to-Gas Plants. They built the Audi e-gas plant in Wertle, Germany in 2013. They claim to offer a modular structure that is suited for outdoor use. The plant consists of the following operating modules:

- power electronics with rectifiers (connection to medium voltage level)
- water treatment system
- hydrogen production (alkaline pressurized electrolysis)
- CO_2 feed gas treatment
- methanation reactor
- cooling system
- ancillary systems

The parent company is:

Hitachi Zosen Inova.

Hardturmstrasse 127, 8005 Zurich, Switzerland.

They have recently received an order for a pilot facility in Japan that will take fossil CO_2 emissions from a coal fired power plant and combine them with hydrogen to produce methane which will then be fed into an existing gas grid [6]. HZI will supply the electrolyzer and the methanation unit.

Predication of Performance and Cost

Each of the three processes involved in the concept is a commercial process at some scale. What is not well developed is the integration of the three processes and the demonstration of the methanation at a suitable scale. With very limited commercial deployment of the concept the predictions of performance and costs is developed from published reports and peer reviewed papers.

Uncertainty

Given the lack of commercial development and the potential for variations in the methanation process configuration there is a relatively high level of uncertainty with respect to the performance and cost.

Additional Remarks

The Audi e-gas plant in Wertle Germany is shown in the following photo.

Figure 3: Methanation Plant



As with the discussion of power to methanol there are multiple options for hydrogen production for methanation systems. Haldor Topsøe designed and constructed a pilot plant that became operational April 2016 [7]. The design capacity is 10 Nm^3/h of upgraded biogas. This capacity requires approx. 50 kW solid oxide electrolyzer cell, which is provided by two Fuel Cores, each consisting of 4 SOEC stacks. Haldor Topsøe A/S also designed the biogas cleaning unit and the methanation plant which is located at the Agricultural Research Centre of Aarhus University at Foulum, Jutland.

Chemical reactions at temperature and pressure can be easily upset if there are rapid changes in the temperature. However, this system has proven to be capable of withstanding power loss without having a significant negative impact on the methanation catalyst. The high efficiency of the SOEC hydrogen production system should lead to a better energy balance than shown in Figure 2. The facility is shown in the following figure.

Figure 4: Haldor Topsøe Methanation Pilot Plant



Quantitative Descrption

Typical Plant Size

It is the biogas plant that will determine the typical plant size and the hydrogen production and the methanation stages can be easily scaled to match the biogas output. Centralized Danish biogas plants range in size from about 70,000 to 700,000 GJ/year [8]. The trend is towards larger plants, driven by cost reductions related to economy-of-scale effects for the biogas plant and in particular the upgrading facilities.

Biogas methanation systems have approximately double the gas output of the biogas or 140,000 to 1.4 million GJ per year. This range is from 2 to 20 times the size of the operating Audi e-gas plant in Germany.

Input and Output

The inputs and outputs for a small system are shown in the following table [2]. The inputs to the system are biogas, hydrogen, and a small amount of power. The system output is methane with approximately 98% methane. There are a number of possible process configurations involving the number of reactor stages, the operating pressures, and gas upgrading schemes.

Table 1: Inputs and Outputs

Parameter	Input	Output
Biogas, 65% CH_4	4,777 MJ	
Hydrogen	4,108 MJ	
Electricity	19 kwh	
Methane		8,000 MJ

Forced and Planned Outage

Ten to 15 days per year of downtime should be expected for the system for routine maintenance and changing catalysts.

Technical Lifetime

The technical lifetime of the systems should be greater than 25 years.

Construction Time

The construction time should be less than one year given the relatively small size and the ability modularize the hydrogen production and methanation stages of the process.

Financial Data

The most recent and comprehensive analysis of capital and operating costs for Bio SDMG systems is the work of Gotz [1].

Investment Costs

Gotz surveyed the literature and reported a wide range in the capital costs for methanation from a low of 36,000 €/GJ/hr to 415,000 €/GJ/hr methane. A small plant producing 70,000 GJ/year could have capital costs of 3.36 million € to a high of 29 million € (€48/GJ to €414/GJ). Gotz concluded that the costs at the low end of the range were the most realistic as they were recently (2014) developed by an engineering company.

The capital costs of the methanation process were less than 5% of the capital costs of the electrolyzer, compressor and hydrogen storage requirements.

Operating and Maintenance Costs

Operating costs are dominated by the cost of hydrogen, which is excluded in this analysis. There is some electricity that is required for the operation of the methanation process but it is small compared to the hydrogen costs. Little information in the literature was identified for operating and maintenance costs. The fixed O&M costs are assumed to be 4% of the capital and the variable portion as 4% of the capital costs.

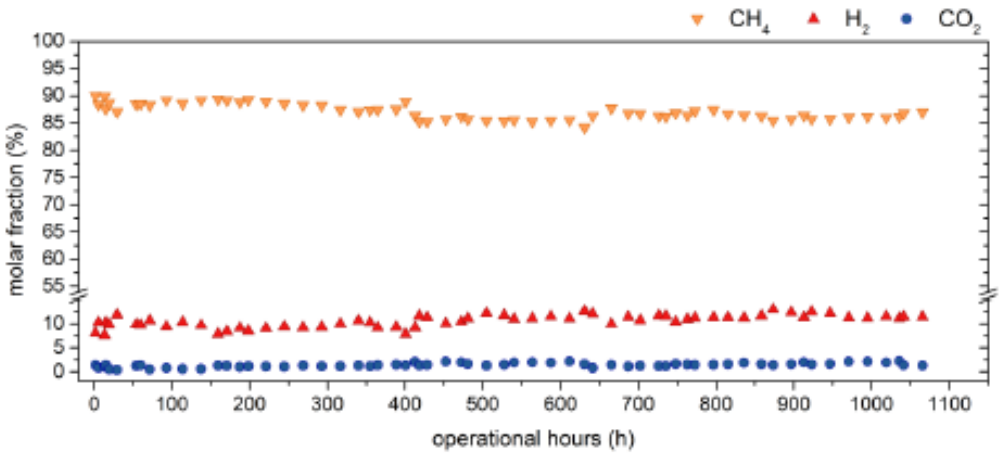
Start-up Costs

The start-up costs are expected to be low in a well-designed system. Some capital investment on intermediate storage of hydrogen and/or biogas will help to minimize the need for gas venting or flaring during start-up.

Technology Specific Data

The composition of the resulying gas will depend on the methanation design and the approach used for gas clean-up. The data from the Swiss project for the gas composition before the final clean-up are shown in the following figure.

Figure 5: Gas Composition Prior to Final Clean-up



With some gas clean-up methane levels of greater than 95% can be achieved. Unlike the composition of fossil natural gas there is little to no C2 to C4 components present. The heating value of the gas may be lower than the typical fossil natural gas as a result, however as it is expected that the gas will be injected into the grid and co-mingled with fossil natural gas at a low concentration, there will be no impact on the final users.

Data sheets

The information on the production process is summarized in the data sheets.

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93 Biogas Biomethanation (Hydrogen Addition)

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Qualitative Description

Methane is a promising energy carrier for the green transition. When produced in biogas plants, biogas consists of 50-65% (volume percent) methane (CH_4) and 35-50% (volume percent) carbon dioxide (CO_2), depending on the substrates used. To achieve grid-quality methane, the CO_2 either needs to be removed or converted to methane through methanation with hydrogen (H_2) which is also referred to as e-methane. This methanation can be achieved either through catalytic methanation or biological methanation – the latter being described in this chapter (will be referred to as biomethanation). For information on catalytic methanation see chapter 92 Biogas Methanation (Hydrogen Addition).

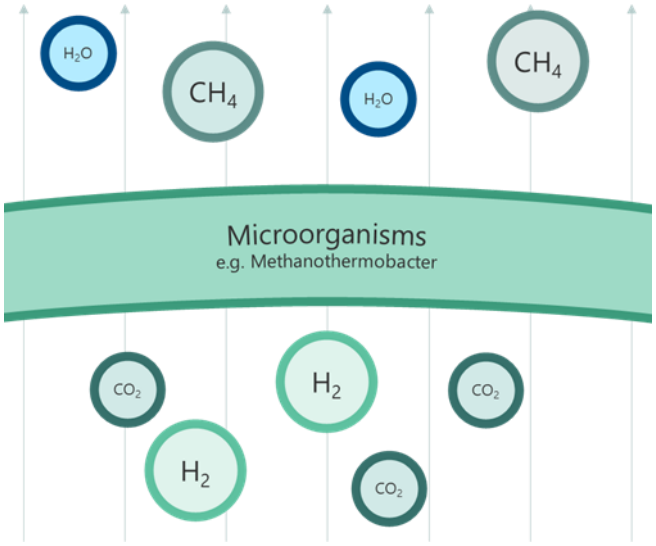
Information on biogas production cost can be found in this catalogue in the chapter 81 Biogas Plants and the cost of upgrading biogas through removal of CO_2 is described further in chapter 91 Biogas Upgrading (CO2 Removal).

This chapter does not consider the production of hydrogen. For information about hydrogen production technologies and costs see chapter 80 Electrolysis. Though biogas and hydrogen production are both commercial technologies, biological methanation has mostly been tested on a smaller scale but attempts at full scale are expected to be completed within the next few years. Today only biogas where no hydrogen is added can receive subsidies. However, biogas used for transport is already removed from the Danish subsidy scheme, but the biogas from methanation will have another certificate as it will be produced with hydrogen from a Renewable Fuel of Non-Biologic Origin (RFNBO). The exact value and certificate that can be achieved depends on the status of the electricity from which the hydrogen is produced.

Brief Technology Description

In biological methanation, carbon dioxide reacts with hydrogen to produce methane, a reaction which is facilitated by microorganism acting as a biological catalyst, as shown in Figure 1:

Figure 1: Microorganisms in a biomethanation process where hydrogen and carbon dioxide react to form methane and water.



The reaction requires four H_2 molecules to react with one CO_2 molecule, resulting in the following reaction:

$$CO_2 + 4\,H_2 \rightarrow CH_4 + 2\,H_2O \quad \Delta H = -165\,kJ/mol$$

When coupled with biogas, biological methanation can be produced in two ways:

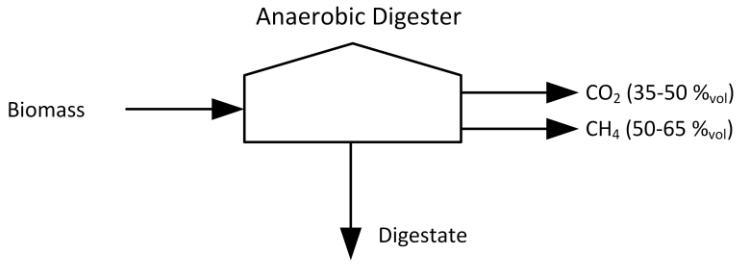
[1] Introducing hydrogen directly to the biogas digester (called in-situ) allowing microorganisms inside the digestor to react with the hydrogen and carbon dioxide.
   1. Leading either raw biogas or the carbon dioxide recovered from a biogas upgrading unit through a separate process tank where hydrogen is added to allow a separate culture of microorganisms to transform it to methane (called ex-situ). Several tank designs at different maturation levels exists for ex-situ, which are covered in "Research and development perspective."

Figure shows simplified sketches of biogas plants with and without biomethanation. For in-situ biomethanation, hydrogen is added to the anaerobic digester which reacts with CO_2 to form methane inside the tank. For ex-situ a separate reactor is installed after the anaerobic digester where the biomethanation occurs.

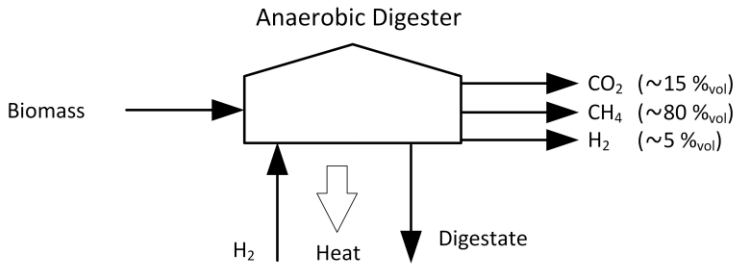
The outlet concentrations denoted in Figure 1 are included as examples and do not represent theoretical upper limits. In-situ and some ex-situ requires post-treatment for reaching a composition enabling biomethane supply to the Danish gas grid. For in-situ, outgoing hydrogen should be separated as part of post-treatment and recycled.

Figure 2: Simplified sketches of biogas plants illustrating the effect of biomethanation on energy and mass flows.

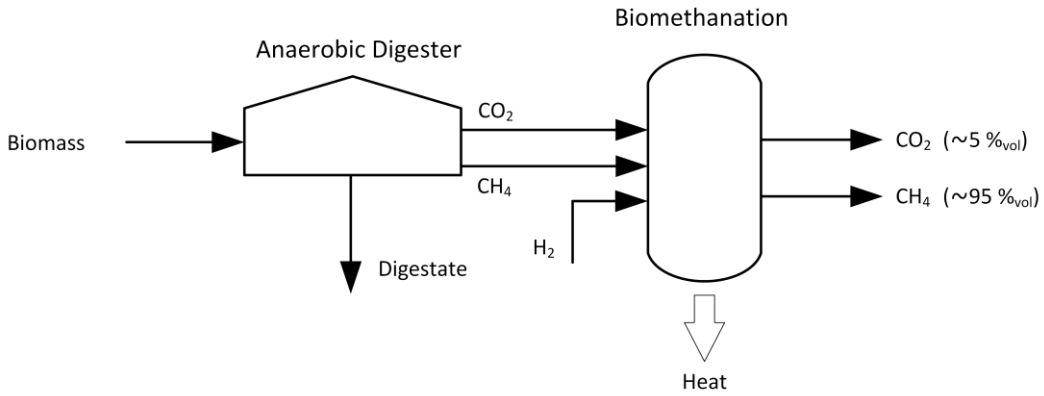
Biogas plant without methanation:



Biogas with in-situ methanation:



Biogas with ex-situ methanation:



In-situ and ex-situ have distinct advantages and disadvantages:

In-situ:

- Lower capital investment than ex-situ
- Only tested in pilot scale
- More difficult to control and optimize due to process limit considering change in pH and hydrogen in the outgoing gas
- Few suppliers

Ex-situ:

- Higher capital investment than in-situ
- Full control in separate methanation tank
- Larger selection of suppliers (see Table 2: Overview of completed and planned biological methanation plants. for examples)

The sketches in Figure 2 consider a situation where biomethanation is included in the design and construction of the biogas plant. Ex-situ biomethanation can also be considered as retrofit to existing biogas plants, and options for integration depend on the layout of the existing plant.

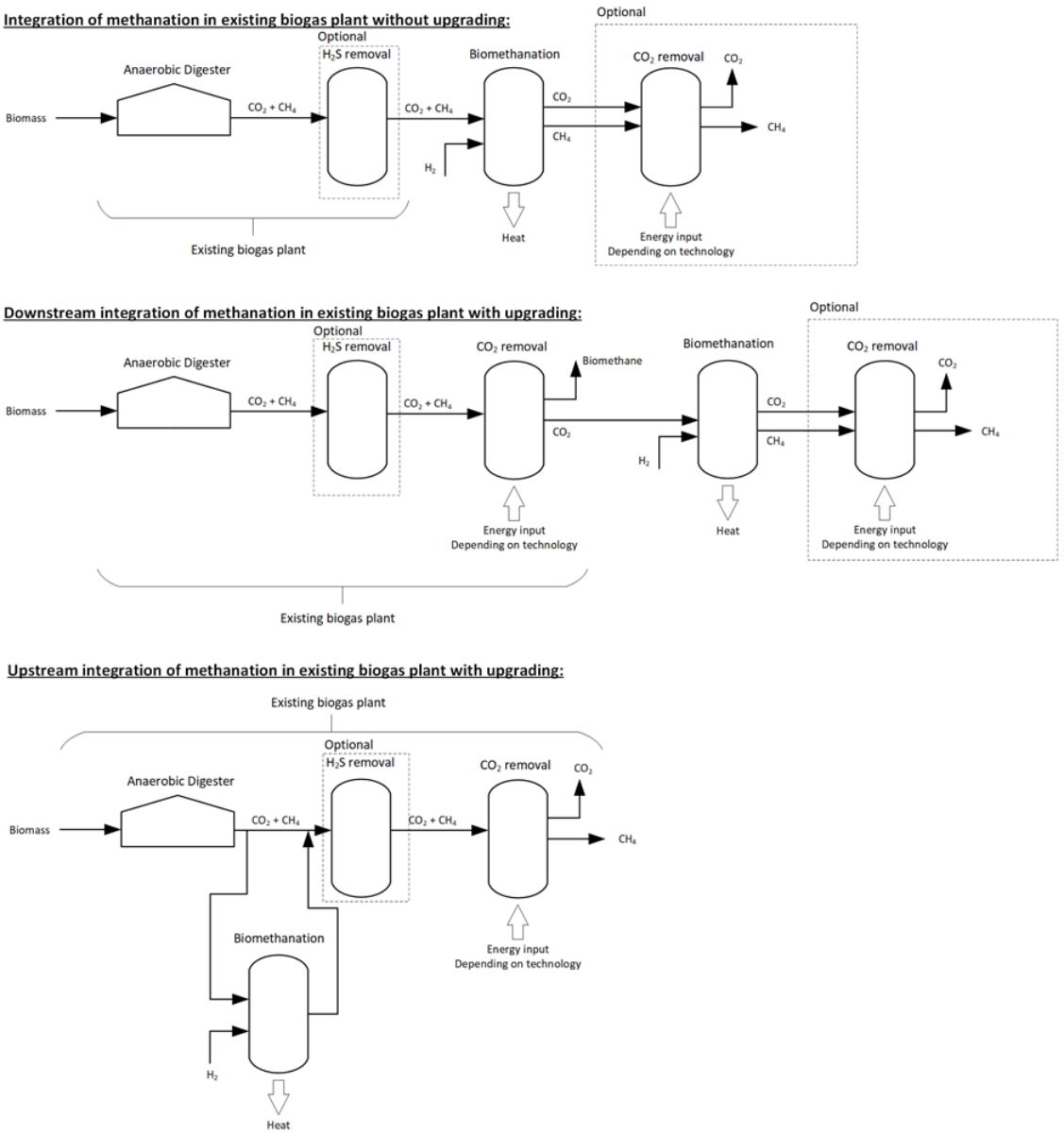
Figure 3 depicts two examples for ex-situ, one where the existing plant includes H_2S removal and one where the existing plant includes H_2S removal and CO_2 removal (biogas upgrading).

In the case without biogas upgrading, the biomethanation reactor is placed after the biogas plant and takes in the full biogas stream, either before or after the H_2S removal unit. After the biomethanation reactor, a CO_2 removal unit (for example a polishing unit) may be necessary to reach biomethane quality. Data from some suppliers indicate that grid level biomethane could be achieved directly after the biomethanation reactor, meaning the CO_2 removal unit and possibly the H_2S removal unit could be omitted, which will lower costs.

In the case with biogas upgrading, the biomethanation reactor can be placed after the CO_2 removal unit, and the reactor receives the residual CO_2 stream. Depending on technology and gas conditions, it may be necessary to polish the methane stream produced in the biomethanation reactor to remove residual CO_2 and reach grid quality biomethane.

Another option for integrating biomethanation in plants with existing upgrading facilities is to install the biomethanation reactor upstream of the upgrading unit. This can potentially remove the need for additional CO_2 removal. In all systems the H_2S can alternatively be removed from the CO_2 stream before processing or discharge.

Figure 3: Integration examples of biomethanation (ex-situ) in existing biogas plants.



Ann.: Data from some suppliers indicate that grid level biomethane could be achieved directly after the biomethanation reactor, meaning the CO2 removal unit and H2S unit could be optional and omittable to save costs. This has been proven in demonstration scale. Water vapor is also removed prior to injection the methane on the grid (not shown).

Input and Output

Biomethanation requires hydrogen and electricity (for auxiliary equipment such as pumps) as inputs and sometimes heating for start-up. In in-situ methanation, the reaction between CO_2 and H_2 occurs in the main biogas reactor/digester, while ex-situ uses either biogas after the biogas digester or CO_2 recovered from an upgrading unit as input. The ex-situ methanation will also require some heat for start-up and some suppliers add water to maintain a stable biological process. However, during normal operation there is a net heat and water output as described below.

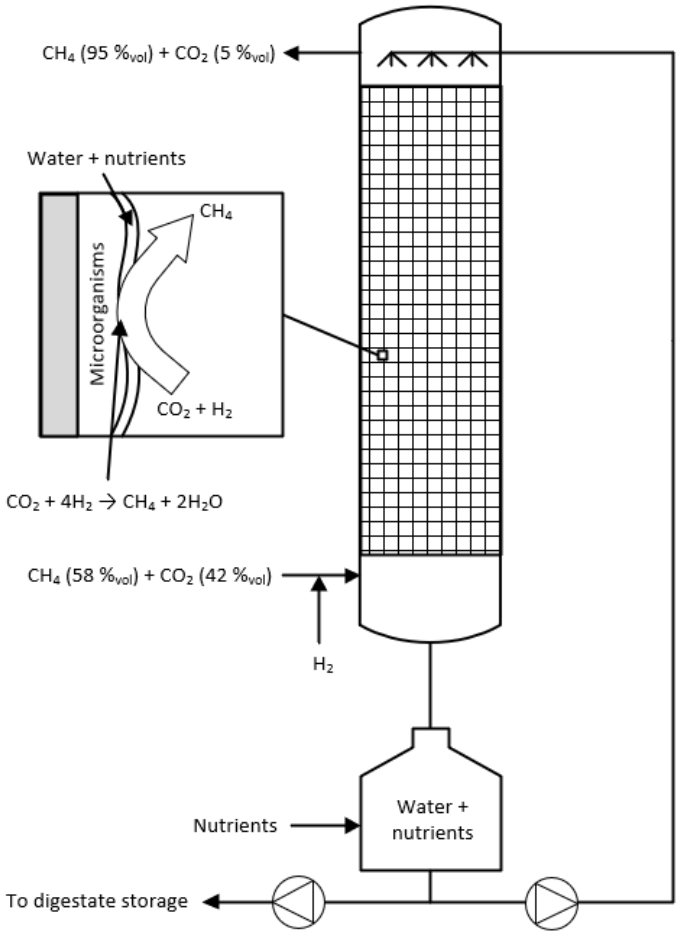
The desired output is methane suitable for injection into the gas grids, which requires some amine upgrading or polishing after the process. The methane concentration in the

raw gas leaving ex-situ reactors can be expected to be around 95-97 $\%_{vol}$, which entails some post-treatment (CO_2 removal) to reach grid injection quality. This varies between suppliers, and some have demonstrated ability to meet grid injection quality without CO_2 removal after the reactor.

For in-situ, the methane concentration is currently around 80 $\%_{vol}$, but is expected to increase as the technology is developed. In addition to CO_2, the outgoing gas from in-situ biomethanation contains H_2 in quantities (around 5 $\%_{vol}$), which makes it reasonable to include H_2 separation and recycling.

There is a net water generation due to the reaction of CO_2 with H_2, and some suppliers add water to the process. The output water will be drained from the reactor and can be pumped to the pre- or after-storage tank along with water vapor removed from the gas and be spread with the digestate. If the incoming feed gas to the reactor is dry, some water will exit the reactor with the outgoing gas due to saturation of the gas in the reactor, see Figure 4.

Figure 4: Sketch of a "trickle bed" biomethanation reactor



Ann.: CSTR reactors have also been developed to near maturity (not shown).

In addition to the gas streams entering the biomethanation reactor, the process relies on the addition of nutrients feeding the microorganisms. The nutrient mix and amount depend on the supplier, but is generally based on nitrogen, phosphorus, sulfur, potassium and some trace elements needed for enzymatic activity.

Heat recovery from ex-situ process depends on the required process temperature. The main heat output from the process will be through cooling of the process and the wastewater. The heat in the wastewater is likely to be utilized indirectly as the wastewater will be pumped to the digester or the 2nd digester tank and is not calculated as part of the input/output streams.

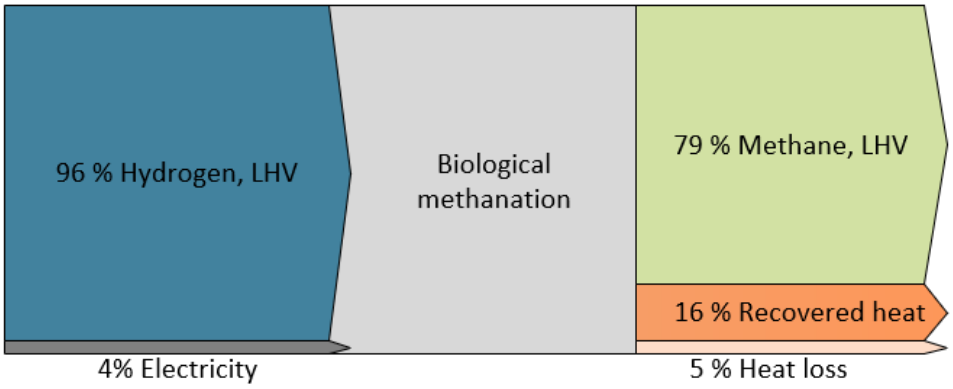
Cooling of the process is included in the battery limit and can be recovered to e.g., process heat for the biogas plant or can be utilized in a district heating network (where it might have to be raised in temperature with e.g., a heat pump). Furthermore, the following upgrading/polishing of the biomethane from the methanation will generate some extra heat compared to an amine upgrading where only biogas is upgraded and the CO_2 is emitted into the atmosphere. However, since the upgrading/polishing unit is not included in this chapter the heat recovery from this process it is not included in the battery limit either.

Energy Balance

The energy balance is shown in Figure 5. The data is based on the biogas having 58% methane content, while biogas with less methane will require more hydrogen and thus have a different energy balance with higher energy demand. The methane produced by the biogas process does not influence the energy balance of methanation and is therefore not included in the energy balance, nor is energy for CO_2-removal.

Excess heat is available from the process at temperatures around 50-60°C and can be utilized in the process or district heating networks but is likely to require the use of a heat pump to meet the local required temperature level.

Figure 5: Energy balance of biological methanation (ex-situ).



Typical Capacities

As biological methanation is approaching market maturity, the scale of current plants does not represent well the plants we will see in operation in the next 3-5 years. A list of plants is shown in Table 2.

Table 2: Overview of completed and planned biological methanation plants.

Project	Supplier	Capacity	Status	Location	Technology	Input
Bioethanol CO_2 side stream methanation pilot	Q Power	Pilot plant	Start of operation 2019	Vantaa, Finland	Ex-situ	CO_2 side stream

Landfill gas methanation pilot	Q Power	Pilot plant	Start of operation 2020	Salo, Finland	Ex-situ	Raw gas
Wastewater treatment gas methanation plant	BiON	100 Nm³/h methane	Start of operation 2022	Dietikon, Switzerland	Ex-situ	Raw gas
BioCat Project - Wastewater treatment gas methanation plant	Electrochaea	Pilot plant 50 Nm³/h methane	Operation from Apr 2016 until Sep 2019	Avedore, Denmark	Ex-situ	Raw biogas and CO_2 side stream
Store&Go Project - Wastewater treatment gas methanation plant	Electrochaea	Pilot plant 35 Nm³/h methane	Operation from May 2019 until Feb 2020	Solothurn, Switzerland	Ex-situ	CO_2 side stream
Nature Energy Holsted Biogas	Biogasclean	Pilot plant 0.8 m³ methane/h	Start of operation 2021	Holsted, Denmark	Ex-situ	Raw biogas
Nature Energy Glansager Biogas	Biogasclean	375 Nm³/h methane	Start of operation 2023	Glansager, Denmark	Ex-situ	Raw biogas
InjectMe	Landia	1.3 Nm³/h methane	Pilot plant start 2021	Foulum, Denmark	In-situ	-

In the future, significantly larger biological methanation plants are expected. Although the cost of biological methanation benefits from scale, the maximum scale is limited by the amount of biogas or biogas-derived CO_2 that can be accessed, while the economic feasibility strongly depends on the cost of the hydrogen compared with the value of the methane. Transporting CO_2 to a centralized methanation plant is a technical option, but the transportation costs are likely be too high for a feasible operation. Instead, the concepts developing in the market are biological methanation plants co-located with a biogas plant (existing or new-build plants), and the capacity is limited by the size of those plants. Capacities are therefore expected to be as follows:

co-located with a smaller biogas plant: 1,000 Nm^3 biogas per hour (8.6 M Nm³ biogas per year) = 420 Nm³ CO_2 per hour (3.6 M Nm³ CO_2)
co-located with a medium biogas plant: 2,000 Nm^3 biogas per hour (17.2 M Nm³ biogas per year)

co-located with a large biogas plant: 6,000 Nm^3 biogas per hour (51.6 M Nm^3 biogas per year)

Biogas plants are usually limited to the amount of bio-substrate that can be sourced in an economically feasible ratio, though we see a drift towards more high dry matter and energy-dense substrates for biogas. However, the biogas plant with Continuous stirred tank reactors (CSTR) has an upper limit for the dry matter that can be stirred in the digester, and high dry matter feedstock mostly requires intensive pretreatment to allow a good degradation of the volatile solid putting an upper limit both from a process and economic perspective. With the development of cost-effective pretreatment systems for high dry matter (e.g., straw) this may result in a shift towards larger biogas plants, and thus also larger scale biological methanation.

Regulation Ability

Biological methanation units can regulate loads between 20% and 100%. Ramping up and down between 20 % and 100 % can be expected to take from a few minutes to a few hours depending on condition of the process, while starting up after longer shutdowns can take from a few hours to several days.

Space Requirements

Space requirement differs between in-situ and ex-situ technologies. While in-situ mainly requires space for hydrogen production (not included in this space estimation), ex-situ furthermore requires space for a separate process tank. There might also be a need for subsequent upgrading (new or enlarged) or a polishing plant, but it is not part of this chapter, and the space for post treatment is therefore not included.

Based on information from suppliers the area requirements for ex-situ methanation plants are about 100-1000 m^2/ MW-methane for methane production capacities between 1-8 MW-methane. The area per production capacity is expected to decrease with increasing production capacity.

Advantages/Disadvantages

Electro-methane from biological methanation can be considered another variation of the power-to-X concept, for example resulting in a green fuel if the consumed electricity for hydrogen production is renewable. The process would essentially increase the quantity of gas that can be produced from biogas plants and there is a ready market for the gas as it is a drop-in fuel in the natural gas distribution systems. In 2022, 34% of the gas distributed in the Danish gas system was upgraded biogas. When compared to catalytic methanation, biological methanation is less capital intensive and the technology and process is like the biogas process in the digester. The operation of the unit will therefore not be much different from operating and optimizing the anaerobic digester process.

Retrofitting biological methanation on biogas plants with an existing upgrading plant it might be necessary to invest in an additional or supplementing system to guarantee that the methane from the methanation meets the gas quality requirements[4] in the gas network depending on the spare capacity in the existing upgrading system.

In-situ methanation in the digester can with some investments in the gas system at the biogas plant be retrofitted to existing digesters. However, the biological process in the digester is multi-staged and currently difficult to monitor in real time. The main parameter that can indicate a stable process is the biogas production. When introducing hydrogen in

[4] Bekendtgørelse om gaskvalitet

the digester there are only limited options for adjusting the process to ensure a good conversion of hydrogen to methane. Since hydrogen will be the main operating expense, this system may have some limitations although investment in a separate tank for the methanation is omitted.

Depending on the existing approval, the biogas plant may have to update the risk assessment according to the: "Directive on control of major-accidents hazards involving dangerous substances"[5] if they by adding biological methanation changes the risk category from a category 2 plant (storage of fewer than 10 tons biogas on site) to a category 3 plant (storing less then 50 tons of biogas on site). The approval process can take at least half a year to complete.

Utilizing CO_2 from the biogas with renewable hydrogen to substitute natural gas can improve biogas production's climate impact. The process does not interfere with the nutrient recycling of the waste product back to the farmland.

The amount of gas produced from the biological methanation process will be limited by the capacity of biogas plants, which is limited to the feedstock handled onsite and the availability. Seen from an energy system perspective each plant tends to be relatively small, which means that economies of scale can only be achieved to a certain extent.

Environment

The impact on the produced electro-methane's GHG emissions will mainly be dependent on how the hydrogen is produced. Hydrogen from electrolysis produced with renewable electricity will produce electro-methane with low GHG emissions. Also, the environmental impact of the biogas production system will be larger than of the methanation process. As with any methane producing or handling process, it is key to monitor and minimize leakages to avoid emitting methane which is a powerful greenhouse gas.

Both in-situ and ex-situ methanation take advantage of an already available CO_2 resource that is converted to a product already handled at the biogas plant and can be injected in the gas network. With the addition of hydrogen, the gas production can be increased with approximately 65% without finding new locations for biogas plants or increasing the transport of feedstocks.

Research and Development Perspective

Biological methanation processes have been in development for the last 10 years and is nearing full-scale operations. Several different reactor designs have been tested and are in development, some of which are shown below in Figure 6. For maturity classification, we use four categories:

Category 1. Technologies that are still in the research and development phase. The uncertainty related to price and performance today and in the future is highly significant (e.g. wave energy converters, solid oxide fuel cells).

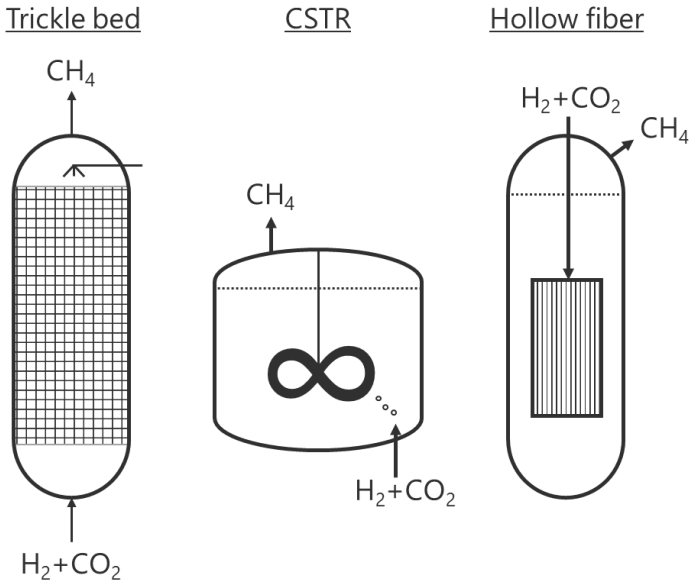
Category 2. Technologies in the pioneer phase. The technology has been proven to work through demonstration facilities or semi-commercial plants. Due to the limited application, the price and performance is still attached with high uncertainty, since development and customization is still needed. The technology still has a significant development potential (e.g. gasification of biomass).

[5] https://eur-lex.europa.eu/legal-content/EN/TXT/PDF/?uri=CELEX:31996L0082

Category 3. Commercial technologies with moderate deployment. The price and performance of the technology today is well known. These technologies are deemed to have a certain development potential and therefore there is a considerable level of uncertainty related to future price and performance (e.g. offshore wind turbines)

Category 4. Commercial technologies, with large deployment. The price and performance of the technology today is well-known, and normally only incremental improvements would be expected. Therefore, the future price and performance may also be projected with a relatively high level of certainty (e.g. coal power, gas turbine).

Figure 6: Examples of reactor design for biological methanation



Ex-situ designs

The furthest developed ex-situ technologies are continuously stirred tank reactors and trickle bed tank designs ("CSTR" and "trickle bed" in illustration above)

CSTRs and trickle beds are category 2 technologies proven demonstration- and semi-commercial plants. Since they have only been applied in a limited number of projects, uncertainties still exist on price and performance. Examples of trickle bed plants are Biogasclean's pilot plant at Holsted and QPower's plant operated on landfill biogas in Salo, Finland. For CSTRs, examples are Electrochaea's plant that has been operated on the biogas produced on the wastewater treatment facility in Avedøre. Full-scale projects are in planning, and it is expected that ex-situ trickle bed and CSTR reaches category 3 in the next 2-5 years, becoming commercial-scale technologies. The commercialization depends on whether the resulting methane is regarded as a renewable gas by the authorities, especially considering the hydrogen component.

Several lower maturity ex-situ technologies promise improved efficiency, usually through increasing the contact between microorganisms and gasses. An example of such a reactor design is a hollow fiber reactor design ("Hollow fiber" on Figure 6), where hydrogen and CO_2 are pumped into the tank through hollow ceramic membranes. The gasses are forced through small pores and diffuses into the surrounding liquid. This ensures a high gas-to-liquid transfer thus increasing efficiency of the reaction. The downside of this design is a relatively low flow rate of gasses through the ceramic membrane, and challenges with membrane fouling. This is a category 1 technology that is being tested in smaller scale with an uncertain time to market, which is not expected to be rapid.

In-situ designs

In-situ biological methanation is also a category 2 technology, which is currently being tested in a single full-scale plant in Foulum. It is expected that in-situ will reach category 3 in the next years, being a commercial-scale technology with some deployment. Especially the optimization of the methanation reaction may pose a challenge, as it is linked with the optimization of the biogas plant.

Examples of Market Standard Technology

Since biological methanation is not yet fully commercialized, it is uncertain which technology type will become dominant in future markets. However, the most developed technologies are ex-situ trickle bed style processing tanks and CSTR tanks (trickle bed and CSTR is shown in Figure 6), and in-situ hydrogen addition.

The ex-situ trickle bed tanks are used by Biogasclean[6], and QPower[7] while CSTR is used by Electrochaea[8], who have developed and tested these designs until demonstration scale. The first full-scale plants are in development. In-situ methods are being developed by several players, one of these being Landia with the technology called InjectMe. The Landia pilot plant is being tested at Aarhus University, Foulum[9].

Prediction of Performance and Cost

Though no full-scale biological methanation plants have been in long-term operation, learnings from demonstration and first full-scale have given a good indication of performance and costs.

The cost profile and performance prediction were made for an ex-situ plant based on the battery limit shown in Figure 7, and has been guided by dialog with suppliers.

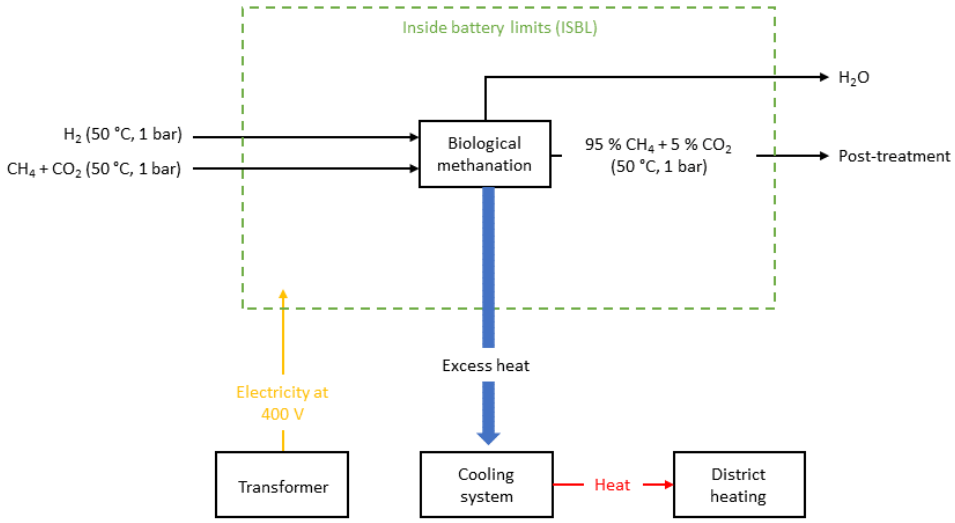
[6] https://biogasclean.com/e-fuel/

[7] https://qpower.fi/technology/

[8] https://www.electrochaea.com/technology/

[9] https://www.landia.dk/injectme

Figure 7: A sketch indicating the assumption on battery limits used for generating the performance and cost for the data sheet.



The following elements (including installation costs) are included in the CAPEX estimates:

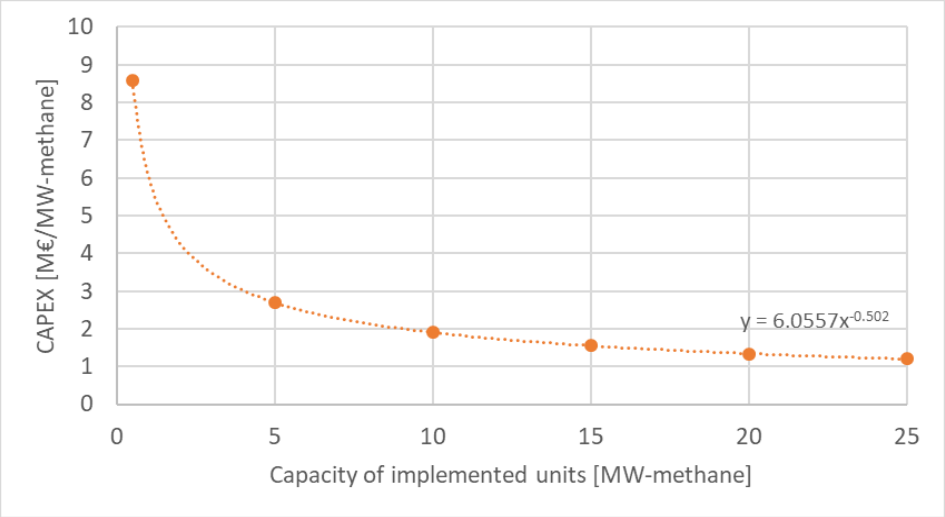
- Biological methanation reactor
- Piping between components inside the battery limits
- Electrical cabling on low voltage side
- SCADA (Supervisory Control and Data Acquisition)

The following elements are not included:

- Biogas plant
- Pre-conditioning of biogas
- Electrolysis unit
- Preconditioning of hydrogen
- High voltage electrical systems
- Fee to DSO for connecting to electrical grid
- Utility systems: cooling system
- Possible connection to district heating including heat pump
- Post-treatment of gases
- Contingencies
- Cost of land
- VAT and taxes
- Owners' costs

Future cost predictions are based on reduced specific cost (€/MW) due to increased size of the methanation units implemented, but also due to technological development. The assumed CAPEX development curve is shown in Figure 8. It is generalized across the suppliers' technologies but note that different suppliers may follow varying CAPEX development curves based on their design choices, such as modular designs.

Figure 8: CAPEX as function of capacity where the capacity of implemented units is expected to develop over time according to the data specified in the datasheet.



Though hydrogen production is outside the scope of this cost assessment, it is key to note that low production cost and availability of hydrogen is key for the financial performance of biological methanation.

Uncertainty

As biological methanation technologies are not yet fully commercialized, their financial performance and cost carry significant uncertainty. Technical performance figures are expected to be less uncertain since they are covered by bio-chemical/physical boundaries.

Acknowledgement

A special thank you to Q Power Oy, Biogasclean A/S, Electrochaea GmbH, and Landia A/S for their support with input and data to this chapter.

Quantitative Descripton

Data

The data for the plant to produce electro-methane from the biomethanation of biogas can be found in the datasheets. The datasheets have been prepared for ex-situ biomethanation with an in-feed of raw biogas according to Figure 7. Energy and investment required for downstream gas conditioning and CO_2 removal are not considered.

The plant size is expected to increase over time as the technology matures, with an upper limit governed by the biogas production capacity of large biogas plants. The specific investment cost is assumed to decrease over time according to the increase in capacity of the plants, but also due to technological development.

94 Syngas Methanation (from Thermal Gasification)

Contact Information

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Reviewer:

Publication date

March 2017

Quantitative Description

Brief technology description

Biomass can be converted to methane by gasification followed by upgrading.

After gasification, upgrading can be done by gas cleaning, CO_2 removal, drying, and methanation of the syngas, to reach a methane content of approx. 95-98%, as required for compatibility with gas in the natural gas grid.

Methanation processes can take place catalytically by conversion of syngas to methane and water. Since the methanation process produces heat it is most often an advantage to integrate the gasification and methanation processes in one plant. The methanation process can also take place by biological processes. The methanation process itself can theoretically reach an efficiency of 80%, the rest converted to heat [21], however the raw gas may contain 5-15% methane already [15]. Therefore, the highest efficiency can be obtained by starting with a gasification process that directly outputs a relatively high share of methane, which is obtained by gasification at moderate temperatures.

Figure 1: Pre-treatment and gasification



The gasification can take place by different principles using both indirect and direct gasification but aiming at a producer gas without nitrogen and a high proportion of methane which reduces the proportion of gas that needs to be methanised.

Fluidised bed gasification may offer these possibilities as they do not show some of the operational limitations seen with fixed bed gasifiers. As such, fluidised beds may be more compact and:

- have an ability to handle fuels with a high ash content and high particle size distribution as well as low bulk density
- prevent bridging, channels and hot spots in the fuel layer
- provide easier scaling up possibilities

Fluidisation is a unit operation by which solid particles through contact with a gas behave as a fluid. The bed in such a reactor may consist of more or less inert solid particles (sand) that become fluidised when a gas, such as ambient air or another agent is blown through the bed. The particles entrain possible fuel particles and the fluidisation enables efficient heat exchange between fuel, sand and fluidisation gas. Due to the fluidisation, the various steps of the gasification process (drying, pyrolysis, oxidation, reduction) that are quite separated in fixed bed gasification, are mixed in a fluidised bed reactor. This enables a uniform temperature distribution and control opportunities and thus control over the process and output. Drawbacks compared to fixed bed conversion comprise a lower carbon conversion ratio (with unburnt fuel in the ash).

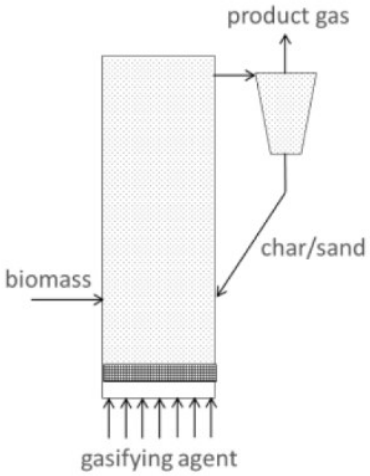
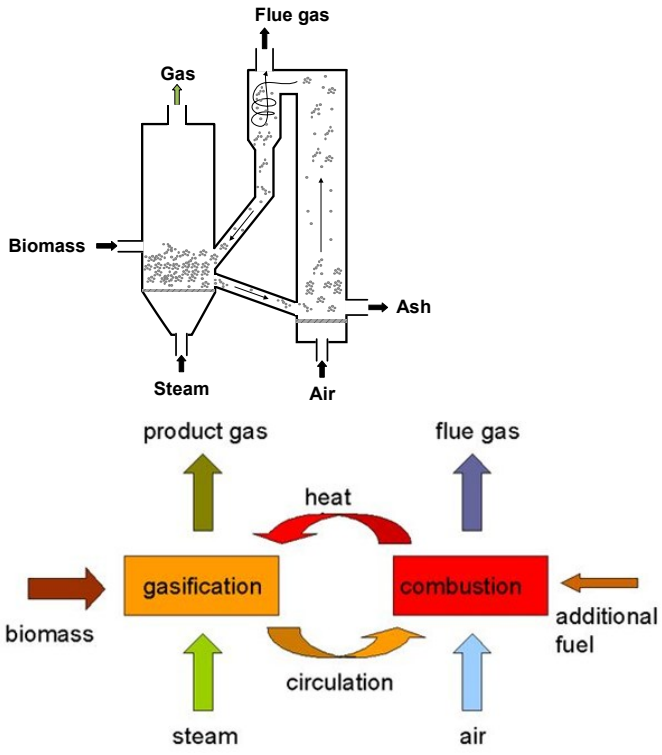


Figure 2: CFB biomass gasifier, principle [8]

Usually, a pre-treatment of the feedstock is necessary including drying, for which excess process heat can be used.

One typical design with indirect gasification uses a dual circulation fluidised bed reactor as shown below, where fast circulating bed material (for instance sand, limestone or olivine) is heated in an air blown reactor by conventional combustion and subsequently returns its heat to the gasification process in the other reactor, where the pre-dried biomass is fed in and which is typically blown by steam. The combustion is primarily fed by the char residues of the biomass feed stock that circulates to the combustor. The gasification can take place at relatively low temperatures (around 800°C) which outputs a gas with relatively high methane content which is relevant for the subsequent methanation process. Further, the low temperature prevents the ash from melting and form corrosive slag.

Figure 3: The dual CFB (circulating fluidised bed) process (Güssing type design) [14]



An alternative typical process design uses direct gasification with a pressurised CFB (circulating fluid bed) reactor blown by oxygen and steam and reaches performance data comparable with the indirect dual CFB type [14], [15].

Oxygen for the direct gasification may be produced by air separation (ASU) powered by electricity generated from process excess heat. This may account for 4% of output energy [14]. Alternatively, oxygen obtained as a by-product of electrolysis in future hydrogen generation plants could be used.

Both the direct gasification and the oxygen blown gasifier process types are able to use wood as feedstock material and can (expected) be upscaled to reach an output in the range 200 MW output. There are as well other variants of the circulating fluid bed technology intended for, or possibly useful for methane production.

A combined gasification and methanation plant may utilise some of the high temperature energy streams to generate electricity [16], [21]. However, the electricity production is not significant, and is for future plants assumed to outbalance the electricity demand for internal processes.

Further description of projects, processes, and technologies can be found in [14], [15] and [2].

Gas cleaning

Tar removal is necessary due to the relatively low gasification temperatures. Several options exist, including scrubbing with water or oil, catalytic or thermal cracking.

In addition, sour gases (primarily H_2S) and CO_2 need to be removed by chemical and/or physical absorption, and the syngas composition may be adjusted by a partial shift for obtaining the required ratio of H2 to CO as suitable for the methanation process.

Methanation

There exist several different process-designs for catalytic methanation of syngas, many of which have been demonstrated in full scale or as pilot plants. [15] [20].

Further, it is possible to convert the excess CO_2 gasses to methane by adding hydrogen gas to the process. This optional process step is not included in this technology sheet.

Input

Solid biomass such as wood chips, pellets, and agricultural waste products. ¨
Auxiliary electricity (may be generated by internal processes)

Requirements to moisture content and size of the fuel depend on the design of the reactor and the process. Fuel with high water content is usually dried prior to gasification in a CFB gasifier. In addition, many demonstration projects have aimed at using waste fractions as a feedstock for gasification [15].

Output

The output is methane
Further output is low temperature process heat, which is assumed valid for district heating
The main waste product output is ash.

Energy balance

The overall efficiency from solid fuel to methane ranges between 50-60% in present demonstration projects, and theoretically could be above 80% [8]. By integration of the

gasification and methanation processes and by use of excess heat to district heating and, possibly, electricity production to cover internal electricity demand, the overall efficiency can be high, likely up to 90%. [3], [16], [21].

In the GoBiGas 20 MW demonstration project, the following efficiencies were measured and reported, though not accounting for an electricity demand of some 3 MW and bio-oil demand of 0.5 MW [21]:

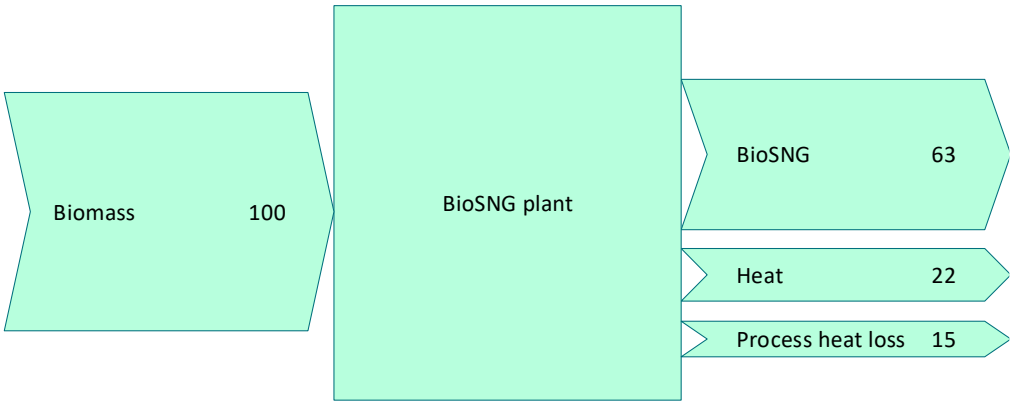
Fuel to cold gas efficiency (syngas): 76.5%

Fuel to methane efficiency: 62.7%

Total efficiency, fuel to methane and heat: 85.4%

These numbers refer to the plant fuelled with wood pellets, and the results cannot directly be transferred to fuels with higher water content, though.

Figure 4: Sankey diagram for a gasifier-methanation plant in 2030.



Typical capacities

The capacity of current (2016) demonstration plants is in the range 20 MW_{gas}.

With a further technical development and the necessary investments, it is expected that the commercial plant size will be up to 200 MW_{gas} by 2020.

Regulation ability

The CFB gasification and associated methanation process plants generally have limited regulation and part load capabilities, depending on the process types though. It is assumed that plants are in continuous operation for 8,000 hours per year.

Space requirement

The main space requirements typically relate to the storage, handling, and possibly drying of biomass feedstock, which can be assumed to correspond to what is required for biomass power plants.

Advantages/disadvantages

A major advantage methanation is the possibility to use existing natural gas infra structure for transport and storage of biomass-based energy in a form, which can be utilized for multiple purposes.

Compared with fixed bed gasification technologies, the CFB technologies with methanation are more technically complex process plants, which in turn can reach higher efficiencies and are more suitable for upscaling. This requires, however, an infrastructure

for biomass procurement, handling, and storage. A substantial heat demand from e.g. district heating systems is an advantage to reach high overall efficiencies.

The CFB gasification processes are typically relatively robust with regard to feedstock quality and can use much larger particle sizes than e.g. entrained flow gasifiers.

The relatively low temperatures of CFB gasification makes it possible to recycle the ash to forests and agricultural land, however tar content and concentration of certain heavy metals such as cadmium may be an environmental problem.

The direct, oxygen blown, CFB technologies may have an advantage over the indirect due to higher throughputs, leading to smaller relative investments, and higher methane rates.

Environment

Generally, the environmental aspects of biomass gasification are comparable to those of biomass combustion processes in general. Depending on the further processes involved in a specific plant, waste products might include condensation waste water, ash with used bed material, used catalytic material, and other waste from chemical reactors etc.

In the case of the pilot plant GoBiGas phase I, [22] mentions annual emissions of 15 tonnes of NO_x and 3.5 tonnes of sulphur as well as a small amount of methane from the methanation process. This must, however, be planning data as the plant was not yet commissioned in 2012. The environmental report from operations in 2015 [23] mentions an emission of 2,933 kg NO_x, 2,997 kg S as well as 1,516 kg NH_3 and 65 kg N_2O from production of 30,000 MWh of biomethane. This corresponds to an average emission of 0.1 kg NO_x, 0.1 kg S, 0.05 kg NH_3 and 0.002 kg N_2O per MWh gas produced. Ref [23] also mentions a number of waste streams containing harmful components such as ash, water-based streams with chemicals, active coal etc.

Research and development perspectives

Process integration and optimisations, including energy optimisations in the integration of gasification, gas treatment and methanation processes. This includes the handling and reforming of tars and preservation of methane from gasification. A specific area of R&D is the methanation process, where several proprietary technologies seem to compete [15]. Experiences from pilot and demonstration plants are expected to lead to further innovation and development that will allow upscaling. It is also expected that such optimisations can eventually lead to improved gas efficiencies compared to present technology [8].

The treatment of biomass fuels is another area for further development, as large quantities of wood and other material shall be transported, handled, stored and dried.

Examples of market standard technology

Indirect, dual CFB:

Valmet is a major supplier of CFB gasification plants in the range 100-200 MW for both combustion in power plants, and also for the GoBiGas demonstration plant.
http://www.valmet.com/products/energy-production/gasification/

GoBiGas is a 20 MW_{gas} technical demonstration project in Göteborg, Sweden, aiming at 65% conversion efficiency, 90% overall efficiency. The plant is fueled by wood pellets and has experienced campaigns of continuous operation since December 2014. [2]. The pilot plant has been retrofitted for wood chips and is in summer 2016 being recommissioned on wood chips. An informative film about the concept can be viewed at:
http://goteborgenergi.streamingbolaget.se/video/156153/link

Direct BFB:

Another major supplier of CFB biomass technology is the company Andritz Carbona which has supplied the bubbling fluidized bed gasifier to the Skive plant in Denmark, commissioned in 2009. https://www.andritz.com/products-en/group/environmental-solutions/power-generation/gasification

Direct, oxygen blown CFB:

Amec Foster Wheeler's "bio-SNG proof-of-concept" for a 12 MW_{th} oxygen/steam blown pressurized (4 bar) CFB plant in Varkaus, Finland [2], [15].

Quantitative Description

Prediction of performance and costs

As of today, the integrated biomass CFB gasification and methanation technologies are in a pioneer phase (Category 2), and the uncertainty regarding future performance and price data is high. Data for 2015 and 2020 are mainly based on demonstration projects.

Assumptions for the period 2020 to 2050

It is assumed that the present demonstration scale plants using CFB and methanation in the capacity range 10-20 MW_{th} will eventually be scaled up and can reach commercial maturity in year 2030 with a capacity in the range of 200 MW gas output. Even though the potential scaling and learning curve effects appear to be significant the estimated future values are widely based on scientific studies of process optimizations, and on industry expectations, and it is not obvious that such development will take place.

The required technical development seems feasible since some of the major elements are already widely used, such as catalytic methanation production based on fossil fuels. Furthermore, the R&D activities involve actively both universities as well as private companies and large energy companies. However, a development of large-scale biomass-to-methane technology will also require the necessary commercial drivers to be present. It is expected that the investments to ensure a further development shall be mainly made by large companies, involved in the energy sector.

Therefore, for development to take place, such investments will have to be evaluated as being overall commercially attractive, at least in the long term, taking the expected future price levels of competing fuels (natural gas and when focusing on the transport sector also of oil) as well as possible subsidies and CO_2 emission costs into account. The data for 2020 – 2050 assumes that such market demand will be present, and that investment costs can be reduced by considerable upscaling and learning curve effects.

Uncertainty

The long-term development of the technology is by nature uncertain, due to the current development stage (Pioneering phase) and the fact that positive results of larger scale deployment are not yet demonstrated. The figures in the data sheets assumes an optimistic scenario in which the described upscaling and learning curve effects will take place and therefore the uncertainty is high.

Additional remarks

Fluidized bed gasifiers may be used for a variety of purposes and can be seen in connection with small to medium scale CHP plants, larger scale co-firing plants CHP plants, methanation facilities as well as liquid biofuel facilities.

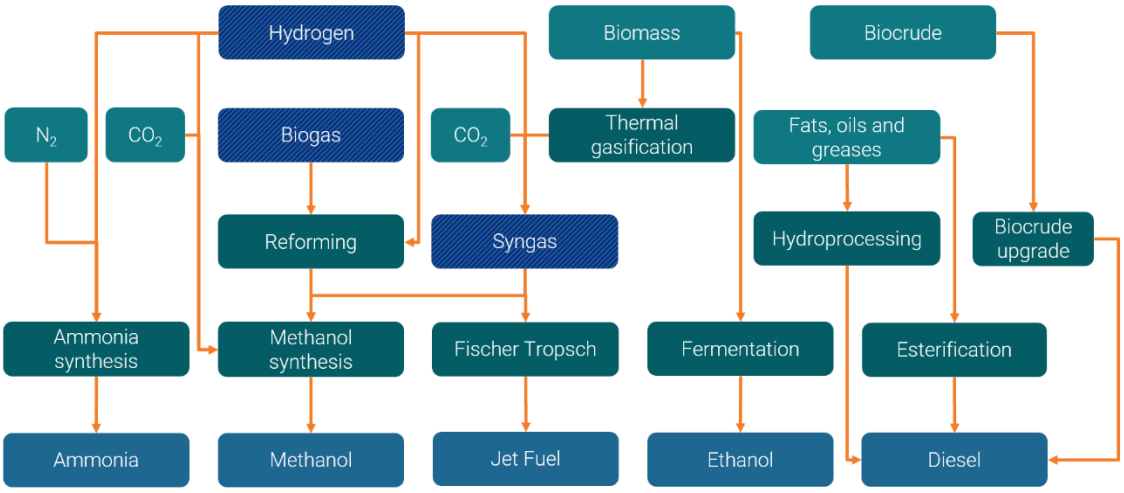
References

Please refer to chapter 99 Jet Fuel from Biomass Gasification for references.

Data sheets

The capacity of the plant is at the lower calorific value of the input biomass (MJ/s), and the output efficiencies refers to the lower calorific value of the methane and heat.

Liquid fuels



This section deals with the production of liquid fuels that are primarily intended for the transport sector. The inputs are either the intermediate outputs from other pathways described in this catalogue, as biogas, syngas or hydrogen or are primary feedstocks as N_2, CO_2, Biomass or Fats, Oils and Greases.

The section includes the following chapters:

95 Methanol from Biomass Gasification

96 Methanol from Hydrogen and Carbon Dioxide

97 Methanol from Hydrogen and Biogas

98 Ammonia from Hydrogen and Air Capture

99 Jet Fuel from Biomass Gasification

100 Jet Fuel from Hydrogen and Carbon Dioxide

101 FAME Biodiesel (Vegetable Oil)

102 FAME Biodiesel (Used Cooking Oil and Animal Fat)

103 HVO Jet Fuel

104 HVO Renewable Diesel

105 First Generation Ethanol

106 Cellulosic Ethanol

107 Upgrading of Biocrudes

95 Methanol from Biomass Gasification

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Publication date
August 2018

Amendments after publication date

Date	Ref.	Description
12/2018	-	Datasheet revised

Qualitative Description

The front end of this process is identical to the gasification process in the previous description. The production of methanol from biomass is a two-step process. In the first step the solid biomass is converted into a bio-syngas and in the second step this syngas is further converted into methanol.

Gasification is a process that converts organic or fossil-based carbonaceous materials at high temperatures (>700°C), without combustion, with a controlled amount of oxygen and/or steam into carbon monoxide, hydrogen, and carbon dioxide (syngas). Stoichiometry for methanol production of syngas requires the ratio of H_2/CO to equal 2. The H_2/CO ratio can be lowered to some extent by the reverse water-gas shift reaction

Depending on the catalyst supplier, the methanol synthesis reaction is normally carried out at about 40 to 120 bar and 200 to 300°C.

Methanol is not the only product that could be produced by this route. Dimethyl Ether (DME) could also be produced instead of methanol or in an additional process step. The methanol could also be further processed into gasoline.

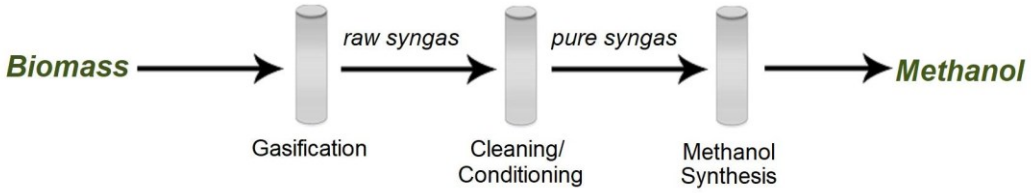
Brief Technology Description

The biomass could be agricultural or forestry residues. There is a wide range in the design of gasifiers used for biomass. Different technological solutions can be implemented in order to obtain different plant configurations; in particular, the mode of contact of the biomass with the gasification agent may be in counter-current, or co-current, or crossflow, and the heat can be transferred from the outside or directly in the reactor using a combustion agent; the residence time can be in the order of hours (static gasifiers, rotary kiln) or minutes (fluidized bed gasifiers). Different gasifier designs are better suited to different feedstocks and gas needs. Gasification is further described in and the subsequent chapters.

The syngas to methanol reaction is practised commercially mostly using natural gas to produce the syngas but there are a few plants that gasify coal to produce the syngas. While the scale of commercial plants is large there have been some small-scale methanol plants built where large natural gas reserves are not available.

The overall process is shown in the following simplified process flow diagram.

Figure 1: Biomass to Methanol Process



Input

The primary input for most process is just the biomass. The reactions are exothermic and generate enough heat for the process and in some cases also enough heat to produce the power required for the system. In other examples power is purchased for the process.

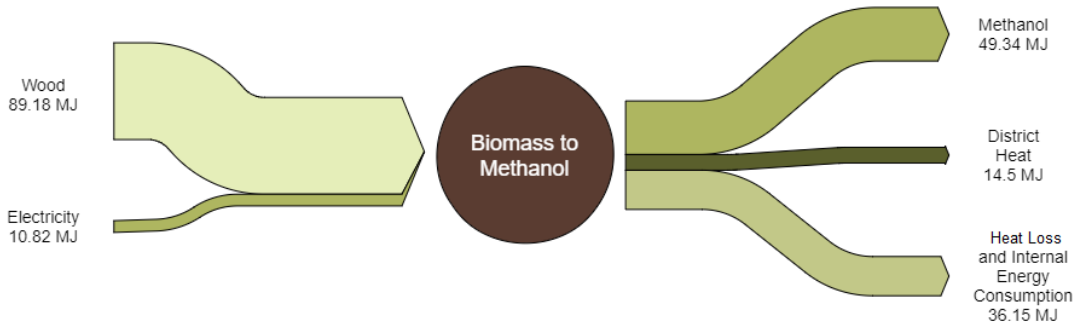
Output

The plants produce methanol and, in some cases, could produce some excess power and/or steam for sale.

Energy Balance

The energy balance for a biomass to methanol system is shown in the following figure [1].

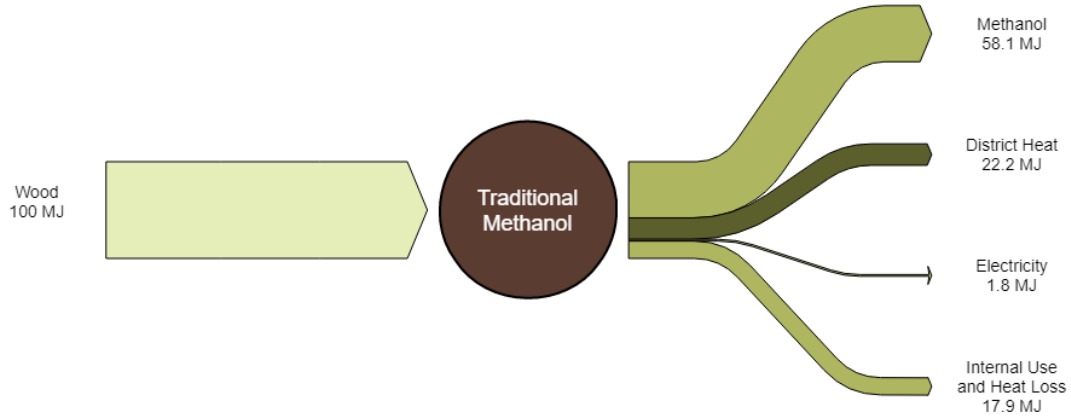
Figure 2: Bio Methanol Energy Balance



There are two potential means to recover some of the waste heat. The plants use some of the process heat to produce electricity for the plant use and potentially a small amount to be exported. Steam from the exit of the final steam turbine would be available for other uses. This could have a temperature between 150 and 185C depending on the design. There may also be some opportunity to recover some lower grade heat as the syngas is conditioned prior to synthesis. Details of the potential for energy recovery are not reported in most of the recent techno-economic studies published.

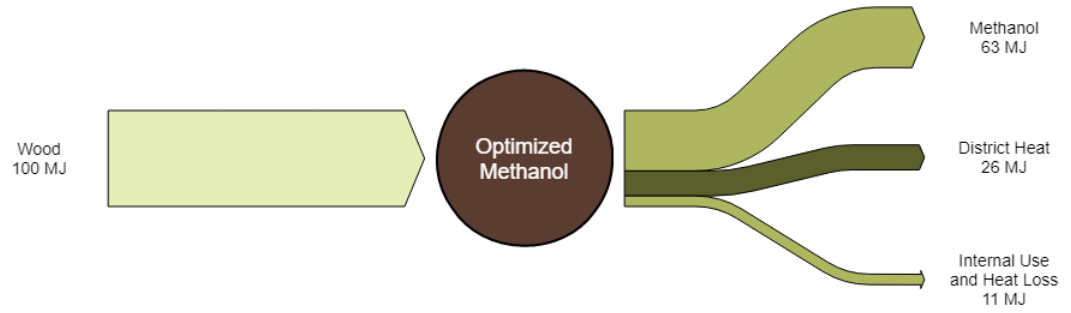
Other biomass to methanol systems have been proposed that offer higher efficiencies [2] [3]. The GreenSynFuels project provided the energy balance for both a traditional biomass to methanol plant and one integrated with a solid oxide electrolyzer to produce hydrogen to provide a better CO to H_2 ratio for the methanol synthesis stage. Clausen [3] provided information for a highly optimized biomass to methanol process. The energy balances for these systems are shown in the following figures.

Figure 3: GreenSynFuels Traditional Methanol Plant



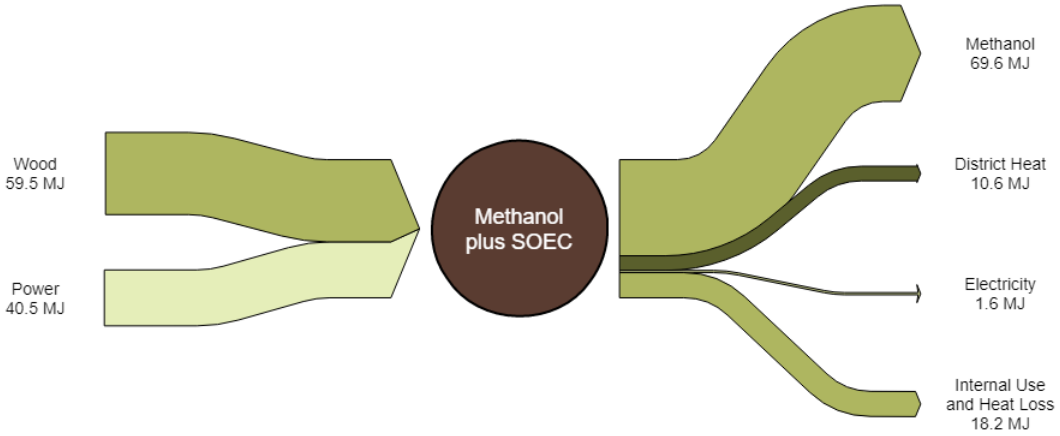
This plant produces electricity instead of consuming it and the methanol production rate is slightly higher per unit of wood consumed. The following figure shows the highly optimized system described by Clausen [3]. The methanol production rate is 8% higher per unit of feedstock.

Figure 4: Optimized Biomass to Methanol Plant



This final energy balance considers the supplementation of hydrogen to alter the carbon to hydrogen ration of the syngas to better match the methanol synthesis requirements. It produces more methanol per unit of energy input and has a much better carbon efficiency.

Figure 5: Hybrid Biomass to Methanol Plant



Typical Capacities

There are currently no commercial biomass to methanol plants in operation. In the past OCI operated a former natural gas to methanol plant on crude glycerine from biodiesel plants as the feedstock in the Netherlands but that operation is now processing natural gas again. There was also a bioDME pilot plant operated in Sweden for a number of years where methanol production was an intermediate product (Chemrec) [9]. It gasified black liquor from a pulp mill rather than biomass.

Commercial plants would likely be similar in size to the biomass to diesel and jet technology that was discussed in the previous section, with and early commercial plant consuming 500 to 1000 tpd of biomass and producing 125 to 250 million litres/year.

Eventually plants could be built larger with feedstock availability being the limiting factor.

Regulation Ability

While biomass gasifiers can operate down to about 35% of rated capacity, commercial methanol plants usually operate at steady state conditions close to the design capacity. Commercial methanol plants can take 2-3 days to reach full production so starting and stopping the plants is generally not an option for regulating capacity. Smaller scale systems with different catalysts may have better regulation capabilities than the large-scale plants.

Space Requirements

Space requirements will be similar to the space for the biomass to diesel and jet pathway, on an area per feedstock basis. The area per volume of fuel produced will be lower due to the lower energy density of methanol compared to diesel and jet fuel.

Based on the Velocys commercial FT liquids plant the area requirements for biomass to methanol are about 0.16 ha/million litres of methanol.

Advantages/Disadvantages

Methanol is not widely used as a transportation fuel today but there are several potential emerging applications that are generating some interest. One is the use of methanol as a hydrogen carrier for fuel cell vehicles such as those developed by Serenergy in Denmark. There is also some interest in methanol as a marine fuel to meet the new IMO sulphur limitations. In China there is some methanol gasoline blending with 10 and 15% methanol. Low level methanol blends

(3%) with a co-solvent have been used in the UK in recent years. Methanol has also been used in blends with ethanol and gasoline in performance vehicles.

Methanol from biomass can be used for the same applications as fossil methanol, while reducing GHG emissions.

Much of the world's methanol is produced from stranded natural gas and is very low cost. It will be difficult for biomass to methanol to complete against these projects on only an economic basis.

Environment

Biomass to methanol should have a very low GHG emission profile, especially when they are designed to be self-sufficient in electric power.

Methanol as a fuel is a biodegradable product.

Research and Development Perspective

Biomass gasification for methanol production from wood or straw is a category 2 technology, a pioneer phase technology with limited applications to date. The technology has been proven to work through demonstration facilities or semi-commercial plants. However, due to the limited application, the price and performance is still attached with high uncertainty, since development and customization is still needed. The technology still has a significant development potential.

This technology pathway is the combination of two commercial systems. There has been considerable development work on biomass gasification in Europe over the past several decades but there has not been a commercial break through yet.

The production of a synfuel from a biomass gasification system is a more demanding application than the use of the gas in an engine or in an external combustion system. It is reported that the Chemrec BioDME plant operated for more than 11,000 hours between 2011 and 2016 [4]. Production during that time was reported to be 1000 tonnes of DME. The capacity of the plant was 165 kg/hour which works out to 6,000 hours of operation.

More work is required on the integration of the two main systems.

Examples of Market Standard Technology

There are no market standard technologies. There is a biomass gasification to methanol proposal for a plant in Sweden, Värmland Methanol [5]. The plant is cost estimated at approximately 350 million Euro and will produce 375,000 liters of methanol per day (130 million litres/year). As a "byproduct" 15 MW of district heating is obtained.

An EPC contract with ThyssenKrupp Industrial Solutions of Germany has been signed. The project was proposed in 2009 but has been unable to raise financing for the project. ThyssenKrupp Industrial Solutions do have experience and expertise in gasification and methanol production technologies.

Enerkem, a Canadian company has operated a municipal solid waste (MSW) gasification to methanol production system in Edmonton Alberta for the past two years. The company is focussed on MSW as a feedstock due to the favourable economics. The Edmonton plant is in the process of being converted to produce ethanol rather than methanol from the syngas.

Predication of Performance and Cost

The prediction of performance and cost is based on published techno-economic papers rather than on actual plant performance.

Uncertainty
There is a high level of uncertainty for the technology given the state of development and the fact that there are no operating plants in the world at this time.

Additional Remarks
There is a Danish Methanol Association promoting bio methanol but the syngas is produced from biogas from anaerobic digestion and not from solid biomass. These systems will be smaller due to the resource availability but with less technology risk since there is no biomass gasification involved.

Quantitative Description
The available quantitative data that is available on the technology is mostly from third parties and not from the technology providers or plant operators. No actual plant data is available.

There are three basic reactions that occur in the process. The first reaction breaks the biomass down, at high temperature and low oxygen, to a combination of hydrogen, carbon monoxide and carbon dioxide. A simplified reaction is shown below. Actual biomass has highly variable composition and complexity with cellulose as one major component.

$$C_6H_{12}O_6 + O_2 + H_2O \rightarrow CO + CO_2 + H_2 + CH_4 + \text{other species}$$

Note: The above reaction uses glucose as a surrogate for cellulose.

Stoichiometry for methanol production of syngas requires the ratio of H_2/CO to equal 2. The product gases are then subjected to the water-gas shift reaction to increase the quantity of hydrogen. The equilibrium for this reaction is temperature dependent which controls the CO to CO_2 ratio.

$$CO + H_2O \leftrightarrow CO_2 + H_2$$

Carbon monoxide and hydrogen react over a catalyst to produce methanol. Today, the most widely used catalyst is a mixture of copper and zinc oxides, supported on alumina. At 50–100 bar and 250 °C, the reaction is characterized by high selectivity (>99.8%):

$$CO + 2\ H_2 \rightarrow CH_3OH$$

Compared to the production of diesel and jet fuel from the gasification of biomass, this pathway requires a lower H_2/CO ratio and operates at lower temperatures but higher pressures.

Typical Plant Size
As with the biomass to diesel and jet process, the plant size will be determined by the feedstock availability. The proposed plant in Sweden would produce 130 million litres of methanol per year (65 MW) from 1,100 tpd of wood [6]. It is not stated but this is likely on a wet basis (660 dry tpd).

NREL undertook a techno-economic analysis of a wood to methanol plant [7]. They based the plant on 2000 tpd of feedstock producing 380 million litres per year (200 MW).

Input and Output
The input and output of a typical system are shown in the following table [2]. These will be n^{th} plant values. Pioneering plants typically have a lower efficiency.

Table 1: Inputs and Outputs

Parameter	Input	Output
Wood, dry	100 MJ	
Power		1.8 MJ
Methanol		58.2 MJ

Forced and Planned Outage
The plants are expected to operate for 350 days per year. Wood gasifiers are capable of operating at these rates as shown earlier and fossil methanol plants are also capable of operating at these rates. Forced outages are expected to be minimal.

Technical Lifetime
Due to the maturity of the technology, plant lifetime is estimated to be 20 years for plants build before 2025. Hereafter, it is expected to increase to 25 years.

Construction Time
Construction time for the technology is expected to be about 24 -36 months.

Financial Data

Techno-economic analyses of standalone biomass gasification to methanol systems have been published [1] [7]. These are used as the basis for the financial analysis and where possible compared to the published data for the proposed Swedish plant.

Investment Costs
The estimated capital cost of the 270,000 tonne per year methanol plant is 369 million Euros [1] (0.91 €/litre). This is the same cost as the Värmland Methanol plant but it is more than twice the capacity. This is a reasonable ratio between the n^{th} plant and a pioneering plant.

The NREL plant [5] was $259 million in 2007 dollars, this would be €0.75/litre.

The EU Sub Group on Advanced Biofuels reported plant sizes in the range of 100 to 200 MW and capital costs of €1,850 to 3,450/kW depending on the scale (€1.00/litre to €1.85/litre).

The capital costs range from €0.75/litre (NREL) to €2.70/litre (Värmland).

Operating and Maintenance Costs
Anderson only presented the combined operating and maintenance costs (Costs for wages, insurance for employees and chemical, water and ash disposal) of €0.10/kg (€0.08/litre). NREL did not provide a breakdown of the operating costs.

Start-up Costs
The start-up costs are included in the costs reported.

Technology Specific Data
Some the properties of methanol are shown in the following table.

Table 2 Methanol Properties

Property	Value
Density, kg/m^3	791
LHV, MJ/kg	19.9
LHV, MJ/litre	15.7
Oxygen content	50 wt%
Blending Octane number	~115
Flash point, C	12

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96 Methanol from Hydrogen and Carbon Dioxide

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12/2018	-	Datasheet revised
03/2023	-	New chapter and datasheet
01/2024	-	Updated datasheet

Qualitative Description

The conversion of hydrogen to methanol is one of the key conversion pathways, which is often considered in Power-to-X concepts and projects. Methanol is of special interest, since it is an important chemical building block and can be used as a green fuel, when produced based on green feedstocks and green energy.

Brief Technology Description

The conventional method of producing methanol is based on the reaction of a syngas composed of H_2, CO and CO_2 in a methanol synthesis reactor with the following main reactions:

$$2H_2 + CO \leftrightarrow CH_3OH$$

$$3H_2 + CO_2 \leftrightarrow CH_3OH + H_2O$$

$$CO + H_2O \leftrightarrow CO_2 + H_2$$

Here, the two first reactions are producing methanol, while the third reaction is the water-gas-shift (WGS) reaction, which occurs in the reactor. In the conventional fossil methanol production pathway, the syngas is often generated based on coal gasification or from natural gas through steam methane reforming. In order to maximize methanol production, it can be beneficial to add hydrogen to the syngas. An optimal composition is achieved when the module *M* is around 2 [1]. The module is defined according to the following equation:

$$M = \frac{x_{H_2} - x_{CO_2}}{x_{CO} + x_{CO_2}}$$

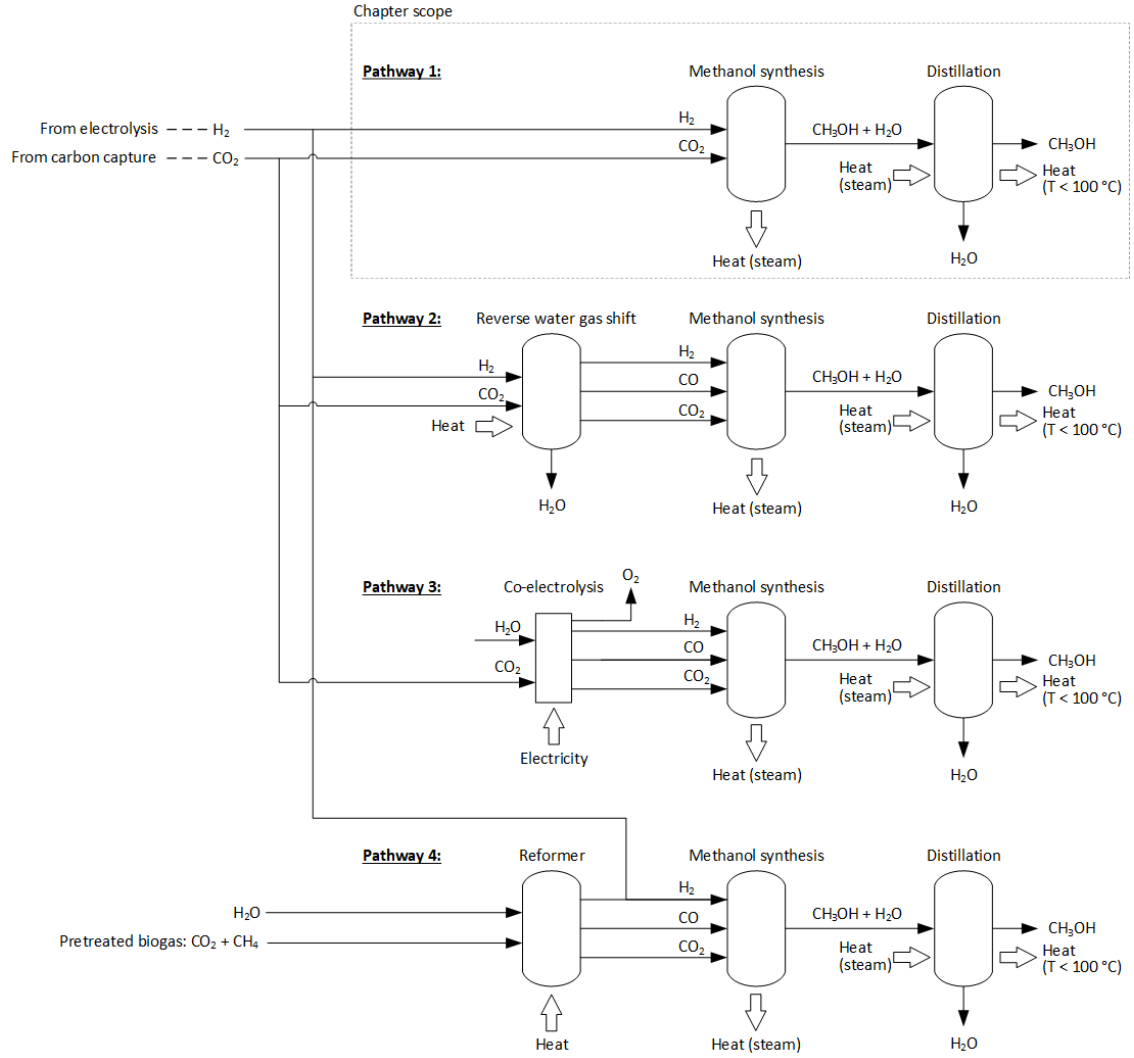
where *x* denotes mole fraction.

Syngas (or synthesis gas):

A syngas is a gas mixture, which can include H_2, CO, CO_2, CH_4 and H_2O. A syngas is a typical intermediate product stream involved in chemical conversion of fuels.

The production of methanol from hydrogen requires an additional feedstock delivering the required carbon atom. Within the scope of green methanol production, the feedstock to the methanol synthesis can be green hydrogen produced by electrolysis with green electricity, and a green CO_2 resource, for example captured from a biogenic point source or by direct air capture (DAC). Another possibility is the use of biogas (CH_4 and CO_2) where a full conversion of the carbon content to methanol can be achieved if H_2 is added as feedstock.

Figure 1: Overview of four pathways enabling the production of green methanol



Ann.: Pathway 1 is the focus of the present chapter, while the other three are examples of alternative pathways involving hydrogen for methanol production. The four pathways are not a full list of possible pathways to methanol. The methanol synthesis generates high temperature heat which can be used to generate steam and, for example, for use in the distillation. The distillation generates lower temperature heat at 50-100 °C, which can be utilized for district heating.

The following four conversion pathways (pathway 1 is the focus of the present chapter) are examples, and not an exhaustive list of possible options, of how green hydrogen can be involved in the production of green methanol or e-methanol:

Pathway 1, direct conversion of H_2 and CO_2 to methanol: The methanol production is based on H_2 and CO_2 as feedstocks. The feed stream does therefore not include CO, which is a difference compared with the other pathways. This pathway is used at George Olah Renewable Methanol Plant in Iceland operated by Carbon Recycling International (CRI) [2].

Main reaction in the methanol synthesis:

$$3\,H_2 + CO_2 \leftrightarrow CH_3OH + H_2O$$

Pathway 2, RWGS route: The H_2 and CO_2 are preconditioned in a RWGS reaction in order to reach a syngas composition including CO, similar to fossil-based methanol synthesis.

Reverse water gas shift reaction:

$$CO_2 + H_2 \leftrightarrow CO + H_2O$$

Pathway 3, co-electrolysis: In a solid oxide electrolysis cell (SOEC) it is possible to co-produce CO and H_2 based on steam and CO_2. This is currently a technology under development with TRL below 5, but could be part of an e-methanol pathway in the future.

Overall reaction in the co-electrolysis:

$$\mathrm{CO_2} + H_2O \leftrightarrow CO + H_2 + O_2$$

Pathway 4, bio e-methanol: The production of green methanol can be achieved by using biogas as a feedstock. Biogas is a mixture of CO_2 and methane, which is often upgraded to biomethane by separating and releasing the CO_2. Instead of separating and releasing the CO_2, it is possible to utilize the CO_2 and methane for methanol production. This can be achieved via steam reforming, which enables the generation of a syngas from biogas. Due to the stoichiometry of the reactions and the composition of biogas, it is necessary to add H_2 in order to achieve a full conversion of the CO_2-content. An example of this pathway has been demonstrated by Lemvig Biogas [3]. In this pathway, the addition of hydrogen can be considered as optional, and thereby the hydrogen production can be operated flexibly depending on the availability of renewable electricity. However, the gas composition in the inlet of the methanol synthesis reactor can then change significantly resulting in intermittent partial conversion. This would then affect the resulting product and require additional downstream operations.

Main reaction in the steam methane reformer:

$$CH_4 + H_2O \leftrightarrow CO + 3H_2$$

The scope of the current chapter and the following description is pathway 1 as indicated in Figure 1. The core of the hydrogen to methanol technology via pathway 1 is the catalytic conversion of H_2 and CO_2 to methanol, which follows the following overall reaction, as named previously:

$$3\,H_2 + CO_2 \leftrightarrow CH_3OH + H_2O$$

The reaction occurs at around 200-300 °C and 50-100 bar [4], and is exothermal with ΔH = -49.16 kJ/mol of methanol [2].

In addition to the reaction above, the reverse water-gas-shift reaction is also present in the methanol reactor, which results in the formation of CO [2]:

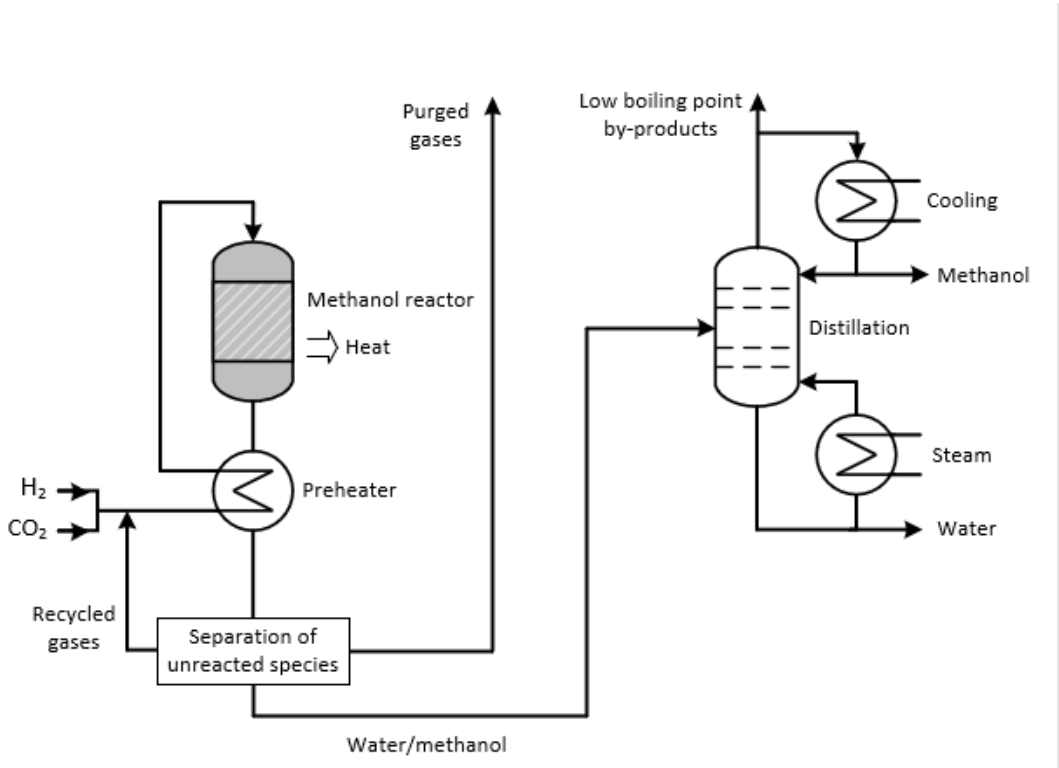
$$CO_2 + H_2 \leftrightarrow CO + H_2O$$

The reaction is endothermic with ΔH = 41.22 kJ/mol of CO. The heat for this reaction will be supplied via the exothermal reaction above, and the overall energy balance results in a net heat output from the reactor.

The methanol reactor can be constructed as a boiling water reactor or a tube-cooled reactor [2], where the heat released from the reaction is carried away as steam or heated water. Boiling water reactors are typically more expensive than tube-cooled reactors [2], however, in terms of heat recovery, steam is a more valuable output stream, since steam is available at higher temperature compared with heated water.

In a methanol plant, there is a range of process steps around the methanol reactor, which are depicted in a simplified sketch in Figure 2. The sketch does not include any pressurization or conditioning of the feed streams, since it is assumed that the CO_2 and H_2 streams are supplied from a central pipeline at the right conditions and purity. Only one distillation column is included, although typical plants include multiple distillation steps.

Figure 2: Simplified sketch of a hydrogen to methanol production plant.



After the methanol reactor, the effluents are used for preheating the incoming feed, and unreacted gases are separated and recycled or purged. The need for purging depends on purity of the feed streams. In case of hydrogen and CO_2 produced from electrolysis and amine-based carbon capture respectively, less pretreatment and/or purging can be expected when compared with fossil based syngases, due to higher purity of the feed streams.

The raw methanol produced in the methanol synthesis is mixed with water, and therefore a distillation is needed in order to separate water and methanol and other byproducts from the reaction, for example higher alcohols, esters, ethers and ketones [2]. The byproducts are small in volume compared with the methanol output and exits the plant as off-gases or as waste water. Low boiling point byproducts can be handled via oxidation, and the waste water can be treated using conventional waste water facilities.

Input

The inputs are feed streams of CO_2 and H_2. In the following it is assumed that the CO_2 and H_2 are supplied from a central CO_2 and H_2 network and therefore entering the considered methanol plant at 5 °C for both streams and at 100 bars for CO_2 and 70 bars for H_2. Based on these conditions the feed streams are already at appropriate pressure levels, and therefore further compression of H_2 and CO_2 is not considered.

Many alternative delivery methods for CO_2 and H_2 can be relevant depending on the specific project, which can result in different feed temperatures and pressures. The following are non-exhaustive examples of alternative situations:

Pipeline based transportation of CO_2 and H_2 at lower pressures for example at around 30 bar. This would require compression of feed streams prior to the methanol synthesis reactor. This will increase CAPEX and OPEX and needs to be added to the costs presented in this chapter.
Truck based transportation of CO_2 for example at 15 bar and -26 °C. This would require compression of the CO_2 prior to the methanol synthesis reactor, and enable the utilization of the cold CO_2 for cooling purposes. Additional CAPEX will be required for compression and system for utilization of cooling, which needs to be added to the costs presented in this chapter. The reduction in OPEX due utilization of cooling from cold CO_2 is not accounted for the in cost and energy data.

Relevant information in relation to feed stream consitions can be found in other Technology Catalogs, for example the following:

For information on compression costs and CO_2 transport modes and their corresponding pressure and temperature conditions see “Carbon Capture, Transport and Storage” specifically sections “CO_2 post treatment” and “Introduction to CO_2 transport”

For information on cost and technologies for producing H_2 see chapter 80 Electrolysis.

For information on cost of compression of H_2 as well as possible different H_2 transport modes and their correspondent pressure and temperatures see “Energy transport” specifically section “Introduction to transport of gases and liquids”

Additional inputs to the process include electricity, cooling and heating. Electricity is used for auxiliary equipment, cooling and heating (electrical or steam) is used in the distillation column(s). Electricity for auxiliary equipment is required at 400 V-AC level and a steam pressure level at around 10 bar(g) and 184°C is sufficient.

Output

The primary output from the process is methanol at a given grade for example US Federal specification grade AA or IMPCA reference specifications, both specifying a methanol content above 99.85%wt.

Additional output streams are purge gases, in case inert gases are present, and separated by-products. Treatment of purge gases and separated by-products are not included in CAPEX estimates. Energy balances and data sheet assumes that no hydrogen is purged.

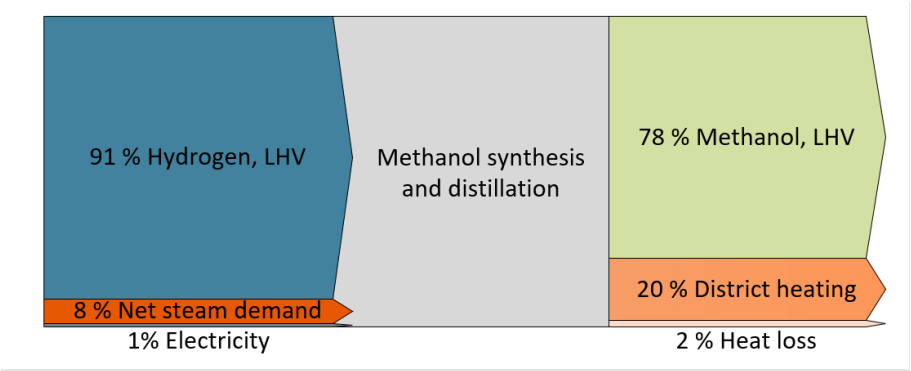
Energy balance

The energy balance of a methanol plant producing grade AA methanol is shown in Figure 3. The energy balance is based on the following information from Haldor Topsøe [5]:

Steam consumption: 1,600 kg-steam/ton-methanol for example at 10 bar(g)
H_2 flow rate: 2,130 Nm^3/ton-methanol
Steam production: 670 kg-steam/ton-methanol for example at 25 bar(g)

The medium pressure steam is generated in the methanol synthesis reactor based on the heat released during the reaction. This steam is at a higher pressure and temperature than the heating demand of the distillation process. In the energy balance it is therefore assumed (included in CAPEX estimate) that the heat (steam) generated in the methanol reactor can be used in the distillation section. The net steam demand in Figure 3 therefore represents the difference between steam consumption and production.

Figure 3: Energy balance of a methanol plant.



The electricity demand included in Figure 3 corresponds to 100 kWh/ton-methanol and includes pumping power and power for cooling systems, but excludes compression of H_2 and CO_2, since these feedstocks are assumed to enter the battery limits at sufficient pressure. Haldor Topsøe [5] also provides a figure for the electricity consumption of the methanol plant of 500 kWh/ton-methanol, which is based on partly pressurized H_2 from electrolysis and atmospheric CO_2. When a value of 500 kWh/ton-methanol is used, the inputs to the energy balance are 86 % hydrogen, 7% net steam demand and 7% electricity demand.

The energy content in the hydrogen is converted to energy content in methanol, but with losses due to the exothermal reaction. Additionally, heat is needed for separating methanol and water in the distillation section. Due to this, 22% of the input energy is lost as heat, where around 20 percentage points can be recovered as district heating. The potential for recovering excess heat for district heating is primarily relevant in the condensers in the distillation column, where heat is available at a temperature level of around 50-100°C. In order to allow heat utilization for district heating purposes, the cold side of the condenser must operate at higher temperatures compared with conventional cooling water temperatures. Alternatively, a heat pump can be installed which comes with additional CAPEX and power consumption, which is not included in the cost and energy data of this chapter.

Typical capacities

Typical capacity of e-methanol plant is around 100,000 ton/year equivalent to 300 ton/day based on the overview of completed and planned projects shown in Table 1.

Table 1: Overview of completed and planned methanol projects.

Project	Capacity	Status	Country
European Energy, Kassø [6]	32,000 ton/year	Expected to be ready for production in 2023	Denmark
Green Fuels for Denmark [7]	50,000 ton/year	Phase 2a planned start of commercial operation in 2025	Denmark
LiquidWind [8]	50,000 ton/year	Start of operation for first plant planned for 2024	Sweden
Project Star [10]	300,000 ton/year	Planned for production in 2025	United States
Shunli Project [11]	110,000 ton/year	Commissioned in Q3 2022	China
Sailboat Project [11]	100,000 ton/year	Commissioning planned in 2023	China
Finnfjord e-methanol [11]	100,000 ton/year	Investment decision expected in 2023	Norway
George Olah Renewable Methanol Plant [11]	4,000 ton/year	Operational since 2012	Iceland

Regulation ability

In conventional methanol production plants, dynamic operation has not been an important design criterion, since fossil feedstocks generally have constant availability. However, e-methanol production relies on renewable intermittent energy sources, and therefore the ability to regulate the production capacity is important for such plants unless sufficiently large storage facilities are implemented.

Nami et al. [1] assume a minimum capacity of a methanol plant to be 40% in 2020, while reducing to 10% by 2030. This assumption reflects that, although regulation ability has not been a concern in the past, there is potential to develop the technology in order to achieve lower minimum capacities, if required. Flows in distillation column and around compressors may have to be recirculated at low loads.

Cold start-up procedures are time consuming and pose a catalyst decomposition risk, and should be avoided as much as possible. The duration of a cold start-up is in the range of one to two days depending on how fast the system can be heated and pressurized and whether catalyst reduction is needed. Instead of shutting down the plant, measures such as hot standby mode, where the system is kept hot and pressurized, of operation should be implemented. Hot standby can be maintained for several days and enables start-up times below three hours.

Space requirement

Expected space requirement for a 300 ton/day methanol production plant is around 4,000 m^2, however, the space needed is subject to specific project requirements. The space requirement is according to equipment shown in Figure 5 and does, for example, not include electrolysers or carbon capture unit.

Advantages/disadvantages

The production of green e-methanol relies on the supply of a green CO_2 source and a green H_2 source. Compared with conventional methanol production, based on syngas from fossil feedstock (example natural gas or coal), the e-methanol production pathway provides the following advantages:

Enables a methanol production pathway which is not based on fossil feedstock
No steam reforming is needed
There is no CO in the incoming feed gas, which results in lower heat of reaction of the methanol synthesis, and it is therefore possible to select among multiple reactor types, when designing the plant. One example is tube-cooled reactors, which are not an option in conventional methanol plants due to the presence of CO in the syngas and the resulting high heat of reaction [2]
Purity of H_2 stream from electrolysis is high, and the same can be the case for CO_2 depending on CO_2 source and capture design. This results in less purge.

The following disadvantages are associated with production of e-methanol:

No CO in the syngas results in a less reactive gas and non-conventional syngas composition
H_2 availability relies on renewable energy sources, which can require dynamic operation design and result in fewer annual production hours compared with fossil feedstocks or need for significant storage of feedstocks or H_2 infrastructure
CO_2 availability can be variable depending on source
Increased water content in the raw methanol, resulting in increased energy consumption in the distillation section.

Environment

The main output streams are methanol, water, reaction by-products and inert gases. Depending on the degree of purification reached in the distillation, the water phase will contain small amounts of methanol and byproducts from the reaction. These organic compounds can be removed on site or in central wastewater facilities. Costs for cleaning waste water is not included in cost data.

Depending on the purity of the feed streams, gases are purged in order not to build up inert gases in the system. The purge stream contains inert gases and unreacted gases and can be combusted or recovered. In conventional methanol plants, purge gases are often burned as part of the combustion process providing heat to the steam reformer. In an e-methanol plant, such a combustion process would not by default be part of the plant, for example if process steam is generated in an electric boiler. However, the purity of the feed stream for an e-methanol plant can also be expected to be higher compared to conventional fossil-based plants, since the purity of hydrogen from electrolysis is higher compared with hydrogen from steam reforming. A higher purity of feed streams would reduce the need for purging.

Similar to purging, low boiling point byproducts, which are separated in the distillation section, would also need to be handled safely – for example combusted or recovered.

Research and development perspectives

Large-scale methanol production has existed for decades, and many of the unit operations required for converting H_2 and CO_2 to e-methanol would be based on existing technology. The research and development perspectives are therefore primarily aimed at adapting and optimizing plant designs for a syngas without CO, but with H_2 and CO_2 of intermittent availability.

Examples of relevant research and development areas are the following:

Optimizing reactor and catalyst design for feedstocks based on H_2 and CO_2
Plant design enabling dynamic operation or design of storages enabling constant feed streams
Realizing synergies with processes such as carbon capture, electrolysis and other PtX-processes
System designs enabling sector coupling (notably utilization of waste heat or oxygen)

Examples of market standard technology

The first e-methanol plant entering into commercial operation is the small-scale CRI plant in Iceland with a capacity of 4,000 ton/year completed in 2011. As indicated in the list of typical capacities section above and in IRENAs Innovation Outlook [4], many plants are planned to enter into commercial operation in the coming years. In 2022, the first commercial scale plant (see Shunli project in Table) of 110,000 ton/year has entered into operation.

Based on the list of planned projects and the recently commissioned Shunli project, the current market standard is assessed to be a production capacity of around 100,000 ton/year. However, plant designs must be tailored to the feedstock available at the relevant location – most often to the CO_2 capacity of an available point source.

Financial data

Financial data are all in Euro (€), fixed prices, at the 2020-level and exclude value added taxes (VAT) and other taxes.

Prediction of performance and costs

The performance of the hydrogen to methanol conversion process is governed by the electricity demand for compression, heat loss in the methanol synthesis and the heat demand for the distillation process, which are well known in the field.

The scope of the present chapter and the following datasheet consider battery limits as illustrated in Figure 5. The H_2 and CO_2 streams are assumed to be supplied at pressure levels above the reactor pressure. The following elements (including installation costs) are considered to be included in the CAPEX estimates:

- Methanol reactor incl. catalyst
- Methanol distillation for achieving grade AA methanol
- Piping between components inside the battery limits (ISBL)
- Electrical cabling on low voltage side
- SCADA (Supervisory Control and Data Acquisition)
- Methanol tank which is assumed to contain one day's production
- Distillation columns prepared for utilization of district heating (higher temperatures of cooling water)

The following elements are not included:

- H_2 and CO_2 compressors
- Electrical transformers
- High voltage electrical systems
- Fee to DSO for connecting to electrical grid
- Utility systems: cooling system, steam boiler
- Possible connection to district heating system
- Treatment of purge gases and separated by-products
- Contingencies
- Cost of land
- VAT and taxes
- Owner's costs

The cost estimates provided in the datasheet are based on cost data presented by Nami et al. [1], but adjusted +10 % based on discussions with suppliers of methanol plants. CAPEX is scaled to different plant capacities based on normal economy scale for chemical process plants. The effects of economy of scale is illustrated in Figure 4 where CAPEX is plotted as a function of capacity.

Figure 4: CAPEX as a function of capacity.

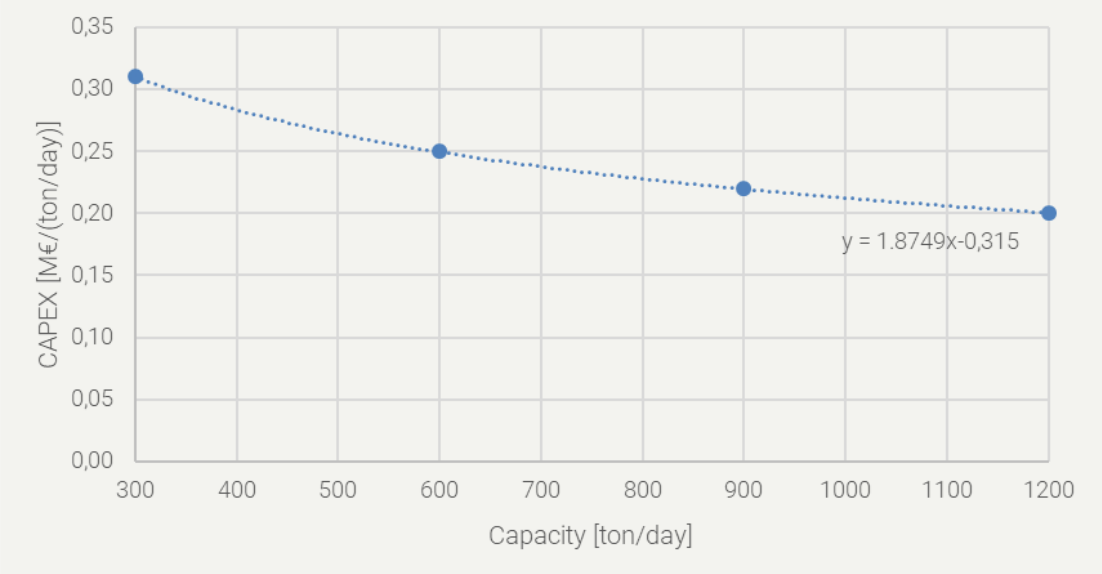
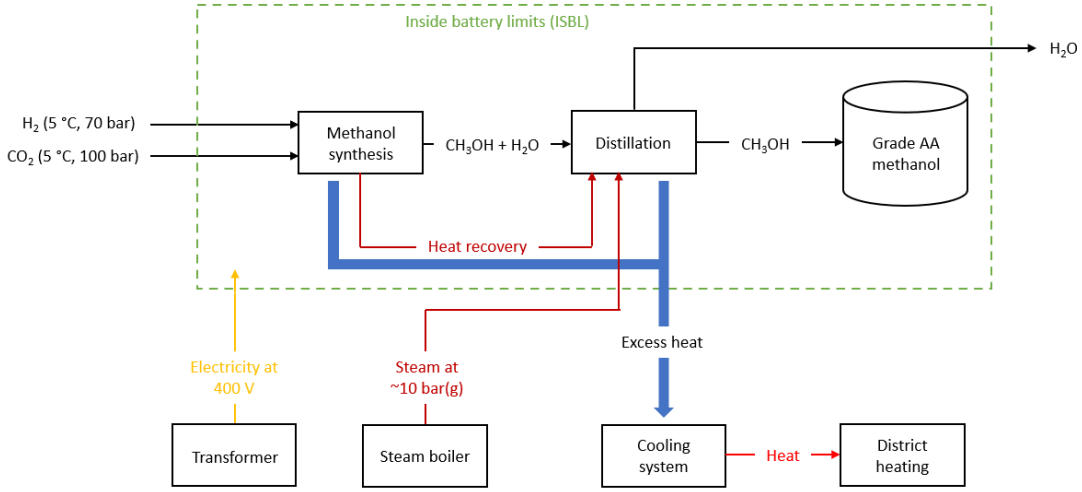


Figure 5: Illustration of battery limits.



A calculation example of a methanol plant with a capacity of 300 ton-methanol/day is shown in Table 2.

Table 2: A calculation example of a methanol plant with a capacity of 300 ton-methanol/day.

Specifications		
Plant capacity	300	ton-methanol/day
Equivalent full-load hours	5,000	hours/year
Assumptions		
Cost of CO_2	100	€/ton
Cost of hydrogen	5000	€/ton
Cost of electricity	100	€/MWh
Cost of steam	60	€/MWh
Price of district heating	30	€/MWh
Waste water tax	5	€/ton
LHV of methanol	19.9	MJ/kg
LHV of hydrogen	120.0	MJ/kg
Calculations		

Methanol output	300 ton-methanol/day/24 hours*5,000 hours*19.9 MJ/kg	345,486	MWh/year
CAPEX	300 ton-methanol/day*0.31 M€/(ton-methanol/day)	93,000,000	€
Fixed O&M	300 ton-methanol/day*9.00 k€/(ton-methanol/day)/year	2,700,000	€/year
Variable O&M	345,486 MWh/year*0 €/MWh-methanol	-	€/year
CO_2 input	345,486 MWh/year*1.4 ton/ton-methanol*100 €/ton	8,750,000	€/year
Hydrogen input	345,486 MWh/year*6.4 MWh/ton-methanol	400,000	MWh/year
Hydrogen input	400,000 MWh/year/120.0 MJ/kg*5,000 €/ton	60,000,000	€/year
Electricity input	345,486 MWh/year*0.1 MWh/ton-methanol	6,250	MWh/year
Electricity input	6,250 MWh/year*100 €/MWh	625,000	€/year
Steam input	345,486 MWh/year*0.58 MWh/ton-methanol	36,250	MWh/year
Steam input	36,250 MWh/year*60 €/MWh	2,175,000	€/year
Total energy input	400,000 MWh/year+6,250 MWh/year+36,250 MWh/year	442,500	MWh/year
District heating output	442,500 MWh/year*0.20 MWh/MWh total input*30 €/MWh	2,655,000	€/year
Water output	345,486 MWh/year*0.55 ton/ton-methanol*5 €/ton	171,875	€/year

Learning curves and technological maturity

The hydrogen to methanol technology has been demonstrated in full commercial scale and is therefore on TRL 9. The technological maturity is assessed to be in category 4 according to the four levels of technological maturity defined in this technology data catalog, however the eMethanol process with dynamic operation is still to be demonstrated at scale.

The methanol reactor is already a mature technology and therefore the development potential is limited. The performance and cost figures are therefore not expected to change in the future. However, there is significant potential in developing business cases for e-methanol plants considering the full supply chain and integrating methanol plants in national energy and carbon infrastructure.

Uncertainty

The uncertainty on the performance estimates is low, since methanol reactor and distillation technologies are well-established, but deviations from the stated values are possible, due to different system designs.

The accuracy of the cost figures is expected to be within ±50 %.

Acknowledgement
A special thank you to Topsoe A/S and European Energy for their support with input and data to this chapter.

Datasheet
The data for a green methanol plant can be found in the datasheets. It should be noted that the evolution of CAPEX with time is assumed to be due to increase of plant size and associated benefits of economy of scale, and not due to technological development. If, for example, a 1200 ton/day plant is expected in 2040, it is therefore possible to use cost data for 2050 corresponding to a 1200 ton/day plant.

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97 Methanol from Hydrogen and Biogas

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Qualitative Description

Methanol is currently a key chemical building block with potential to become a dominant sustainable transportation fuel. Conventionally, methanol is mainly produced from fossil-based feedstocks (coal or natural gas), which are also used to generate the required process heat. The feedstocks are converted into syngas, which again is converted into methanol. The conversion of syngas to methanol is explored in this chapter, but using more sustainable feedstocks including biogas and hydrogen which can be sourced from electrolysis using renewable electricity and water. The reaction pathway to more sustainable syngas has the potential to significantly reduce the GHG-emissions associated with methanol production. This can lower the carbon footprint of conventional methanol use in the chemical industry but also in the transport sector as a fuel for long distance and heavy weight freight such as shipping.

This chapter describes the process of converting biogas and hydrogen to methanol, described as Pathway 4 in chapter 96 Methanol from Hydrogen and Carbon Dioxide

The plant capacity and biogas composition used as basis for this chapter is in line with chapter 81 Biogas Plants having a capacity of 5,000 Nm3 per hour of raw biogas consisting of 57.5% CH4 and 42.5% CO2.

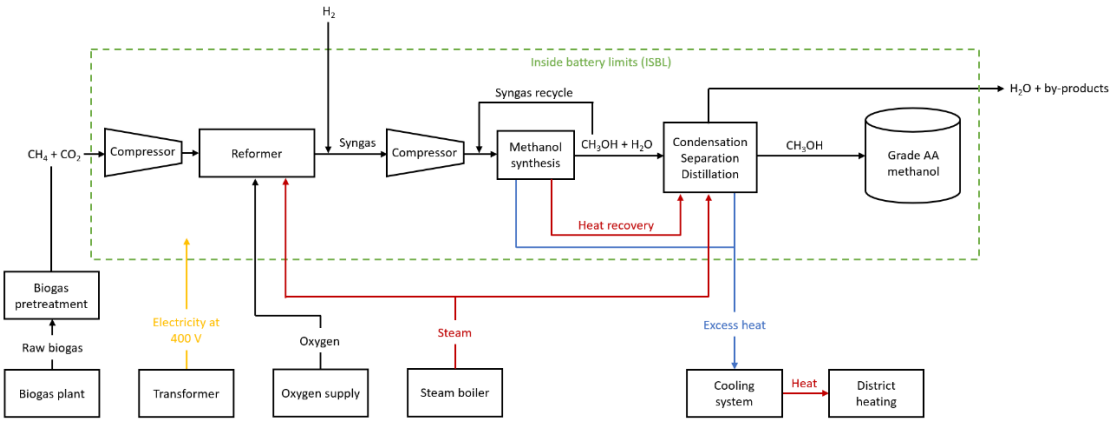
Data is mainly based on a simulated plant by FiW Aachen University [1] and [2] which is based on a pilot plant in Dinslaken Germany, also by FiW Aachen University. The data provided by FiW Aachen University is supplemented knowledge shared by the Technical University of Denmark and DBI Gruppe as well as NIRAS' knowhow of conventional methanol synthesis plants and emerging biogas-based methanol plants.

Brief technology description

Methanol consists of carbon, oxygen, and hydrogen, which can be sourced from various sources containing these atoms. This chapter focuses on methanol based on treated biogas and hydrogen as shown in

Figure 1.

Figure 1: Illustration of battery limits



Raw biogas is composed mainly of CH_4 (50-75 vol%) and CO_2 (25-50 vol%) plus minor concentrations of hydrogen, nitrogen, H_2S and NH_3 [3]. Sulphur compounds are harmful to the catalyst applied in the biogas-to-methanol process; thus, sulphur is removed prior to the process, which can involve biotrickling filters, active carbon, sulphur guard and zinc oxide adsorbers [4]. Details regarding biogas pretreatment are beyond the scope of this chapter as seen by the battery limits in

Figure 1.

In the following sections the reformer, methanol synthesis and distillation are described separately.

Reformer

Reforming is the process of heat treating a hydrocarbon. Reforming usually occurs using a catalyst to increase reaction rate, enhance selectivity towards a desired product and reduce energy applied to the reforming process. Steam reforming is an example of reforming and it occurs at high temperatures at around 1,000°C and at moderate pressures of 20 to 50 bar [5].

The investigated biogas contains CH_4, which together with CO can result in carbon deposition that profoundly lowers the activity of the catalyst in the reformer. Carbon deposition from CH_4 is favoured at high temperatures while carbon deposition from CO is suppressed at high temperatures. The chemical reactions resulting in carbon deposition are shown in the reaction scheme below [6]:

$$CH_4 \rightleftharpoons C + 2\,H_2 \qquad\qquad 2\,CO \rightleftharpoons C + CO_2$$

Another important parameter for methanol synthesis is the module defined as $M = (H_2 - CO_2)/(CO+CO_2)$, which for conventional methanol synthesis is around 2.05 (stoichiometric number for methanol synthesis corresponding to ideal conditions) due to to kinetic reasons [7]. Further details can be seen in chapter 96 Methanol from Hydrogen and Carbon Dioxide

It is desired to add oxidizing agents (e.g. steam) to prevent coke formation and increase CH_4 conversion. However, the metallic catalytic particles are mostly active as reduced (non-oxidized) metal particles, hence the addition of oxidizing agents must be controlled.

Addition of steam is known as steam methane reforming (SMR) and involves the reaction between CH_4 and H_2O to form CO_2 and H_2 [4] as shown in the reaction scheme below:

$$\mathrm{CH_4 + H_2O \rightleftharpoons CO + 3\ H_2}$$

This chapter does not consider SMR but rather two reforming technologies termed *bi-reforming,* being a combination of dry and steam reforming, and *tri-reforming* with an autothermal reactor and oxygen injection. The two considered types of reforming are described separately in the following sections.

Bi-reforming

The first step of bi-reforming is to compress the pretreated biogas before it enters the reformer, where a Ni-based catalyst accelerates the direct reaction between CH4 and CO2, which is known as dry reforming. The chemical reaction of dry reforming is shown in the reaction scheme below:

$$\mathrm{CH_4 + CO_2 \rightleftharpoons 2\ CO + 2\ H_2}$$

The chemical reaction of bi-reforming is shown in the reaction scheme below:

$$\mathrm{3\ CH_4 + CO_2 + 2\ H_2O \rightleftharpoons 4\ CO + 8\ H_2}$$

The above reaction assumes a CH_4/CO_2 ratio of 3/1, which is larger than most biogas compositions, including the investigated composition in this chapter. A smaller CH_4/CO_2 ratio results in a lower H/C ratio but addition of hydrogen can change the H/C ratio as desired for methanol synthesis.

The high steam reforming temperature of up to around 1,000°C has traditionally been achieved by burning fossil fuels. In order to lower GHG-emissions, fossil fuels can be replaced by biogas. However, recently a new synthesis gas production platform based on Joule heating (resistive heating using electricity to supply the high temperature) has been introduced, which allows for the utilisation of electricity based on renewable energy sources [4]. The pathway is termed *bi-reforming with eSMR*, where "e" symbolises the use of electricity [8].Using electricity rather than biogas to generate heat allows for more biogas to be used in the methanol synthesis. Moreover, the generated heat is not dependent on chemical reactions occurring in the reformer nor has it downstream impact on the syngas composition, as is potentially the case for tri-reforming using oxygen as described in the following section.

Tri-reforming

Tri-reforming using an auto thermal reformer (ATR) involves addition of oxygen to oxidize CH_4 and CO, generating heat as oxidation is exothermic, and conversion to syngas using a subsequent SMR catalyst [8]. The oxidation reactions are described by the reaction scheme below and takes place in the reformer.

$$\mathrm{2\ CH_4 + \ O_2 \rightleftharpoons 2\ CO + 4\ H_2}$$

$$\mathrm{2\ CO + \ O_2 \rightleftharpoons 2\ CO_2}$$

This name *tri-reforming* indicates three reactions occurring in the reformer [8]. The three reactions are as follows:

- Dry reforming
- Steam reforming
- Oxidation of CH_4 and CO

It is important to carefully control the amount of added oxygen as a surplus of unreacted oxygen can move to the methanol catalyst causing it to oxidize, which reduces its activity and thereby the methanol yield. The control of oxygen means that other heat sources such as external fuel burning may be needed to achieve the required temperatures. In addition, the reformer catalyst must be tolerant to the oxidation processes.

Both bi-reforming and tri-reforming related to methanol synthesis involve addition of H_2O and H_2, though potentially in different amounts. The main differences between the two pathways are as follows:

Tri-reforming involves oxygen as oxidant in combustion processes with fuels to generate heat, whereas other heat sources are used in bi-reforming, e.g. electricity

Heat can influence the syngas composition; thus, well controlled heat management is important to control the syngas composition. External heat supply (bi-reforming) e.g. from electricity is generally easier to control compared to the chemical oxidation (tri-reforming)

In this chapter it is chosen to focus on the tri-reforming technology as this is assessed to be a more promising towards high conversions of reactants compared to bi-reforming [9].

Methanol synthesis

Methanol synthesis from syngas involves three main reactions as seen below [10]:

$$CO + 2\,H_2 \rightleftharpoons CH_3OH\,, \qquad \Delta H^0_{298K} = -91\ \mathrm{kJ/mol}$$

$$CO_2 + 3\,H_2 \rightleftharpoons CH_3OH + H_2O\,, \qquad \Delta H^0_{298K} = -47\ \mathrm{kJ/mol}$$

$$CO + H_2O \rightleftharpoons CO_2 + H_2, \qquad \Delta H^0_{298K} = -41\ \mathrm{kJ/mol}$$

The first reaction is a combination of the other two. The first two reactions are exothermic, in relative terms, and result in a decrease in the number of moles. Thus, low temperatures and high pressures favour the reagent's conversion. Today's methanol reactor technology is designed to operate in the pressure range of 50-100 bar.

The CO and CO_2 pathways have different kinetics and thermodynamic characteristics, depending on syngas composition, one or multiple kinetics pathways can be favoured. Typical syngas composition ranges suitable for optimal methanol production are module = 2.02-2.1, where the module is $(H_2 - CO_2)/(CO + CO_2)$, and the carbon oxide ratio (COR) is 0.1-0.6, where COR = $CO_2/(CO + CO_2)$. Syngas parameters outside this ranges may be chosen in accordance to plant characteristics and systems integration.

Single pass conversion of carbon to methanol is typically low, between 5%-15% [11]. This limitation is mainly driven by the reaction heat extraction. Since the reactions are mainly exothermic, inadequate heat extraction from reaction sites comports a temperature increase which in turn will decrease the thermodynamic drive to further continuation of the reaction, lowering the global methanol conversion. For this reason, methanol producing

units employ relatively large recycling loop for recovering and cooling all the non-reacted chemical compounds looping them back in the reactor, as seen in

Figure 1 (syngas recycle). The inclusion of large recycling loop in the design of the plant allows for high carbon efficiency.

Distillation

The methanol synthesis process results in methanol combined with water. This necessitates distillation to separate methanol, water, and other minor reaction byproducts such as higher alcohols, esters, ethers, and ketones. These byproducts, although minor in quantity compared to the methanol yield, leave the plant as off-gases or waste water. Byproducts with a boiling point lower than methanol can be managed through oxidation, while conventional waste water treatment facilities can handle the waste water.

Input

Feedstocks are biogas (CH_4 and CO_2), hydrogen, oxygen, steam and electricity as shown in Figure 1. The biogas must be cleaned from sulphur and other components, which inhibit catalytic reactions from biogas to methanol.

The origin of biogas is described in chapter 81 Biogas Plants. However, biogas may also be generated through other processes such as biological methanation of syngas from gasification or pyrolysis gas.

Heat is supplied to initiate and drive the endothermic reactions, while heat from exothermic reactions including the methanol synthesis can be reused to provide energy for heat consuming reactions/processes.

Output

The output concentrations of the methanol reactor depend on the reaction conditions (temperature, pressure, separators/distillation efficiency and methanol reactor typology) and input feedstocks but is mainly methanol, hydrogen, and unreacted methane but also include CO, CO_2, water and nitrogen. The output mix is separated into a final methanol product with methanol as the dominant component by weight at >99 wt.%. Grade AA methanol corresponds to at least 99.85 wt.% methanol including other requirements such as <0.10 wt.% water, <10 ppm ethanol etc. [12].

Energy balance

Figure 2 shows the energy balance of methanol synthesis based on biogas and hydrogen. The energy balance is based on experimental and simulated technical inputs from FiW Aachen [1]:

Table 1: Energy balance parameters

Parameter	Value	Unit
Methanol production	0.35	Tons methanol per hour
Steam consumption	0.05	MWh steam per hour
Electricity consumption by auxiliary units	0.12	MWh electricity per hour
Effluent flow	0.07	MWh effluent per hour

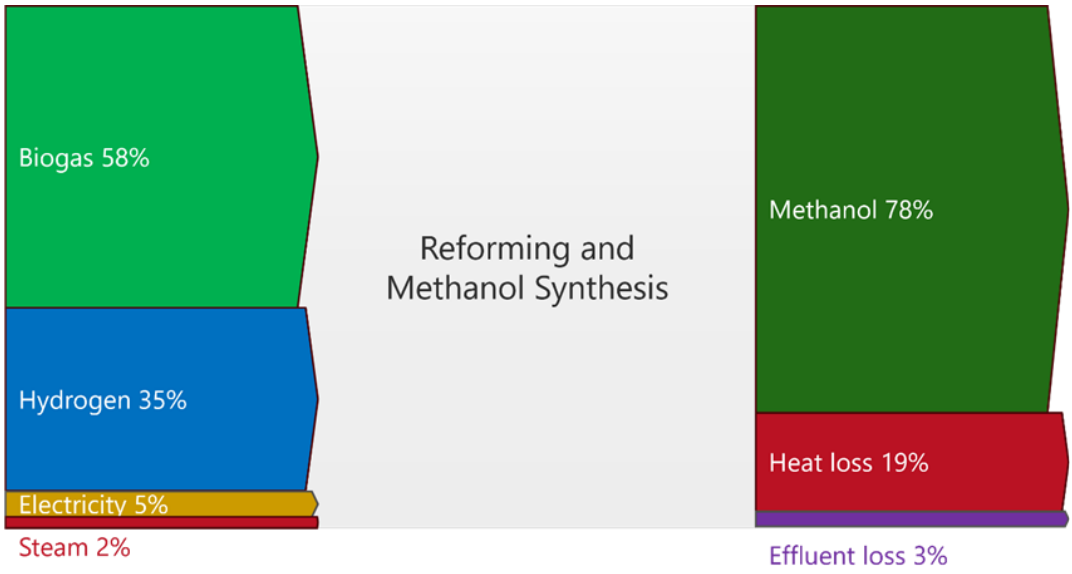
Hydrogen consumption	290	Nm^3 hydrogen per hour
Oxygen consumption	105	Nm^3 oxygen per hour
Water consumption for reformer	31	Kg water per hour
Heat loss	0.47	MWh heat per hour

Source: (Klimazukunft u.d.).

Values reported as per ton methanol are directly used for the case with 5,000 Nm^3 biogas/h, whereas other values including hydrogen consumption is changed to match the 5,000 Nm^3 biogas/h capacity.

Steam is used for separation of water and methanol (approximately 67% of steam consumption) and preheating syngas (approximately 33% of steam consumption).

Figure 2: Energy balance of a methanol plant using biogas and hydrogen as main feedstocks.



Typical capacities

The technology has not yet achieved commercialization, and existing plants are all at a pilot or demonstration scale corresponding to 10 to 20 Nm^3 biogas per hour [13] [14]. The assumed capacity is 90 Nm3 biogas per hour in 2025. With technology development capacity can increase to 900 Nm3 biogas per hour in 2030 and 5,000 Nm3 biogas per hour in 2040 and 2050. The largest capacities in 2040 and onwards match the capacities of biogas plants.

Regulation ability

The biogas to methanol plant is assumed to be capable of operating at minimum load capacity of 60%. Below 50-60% the heat generated in the exothermal methanol synthesis process is insufficient to provide the heat required for internal heat integration. If the plant starts in “cold” condition, it takes around three hours to get above 60% load capacity. If the plant is in “hot” condition it takes around one hour to reach about 60% load capacity, whereas it takes additionally around 1.5 hours to get from 60% to 100% load capacity. The 1.5 hours is used mainly to stabilize the temperature in the methanol synthesis part of the

plant. In general, it takes around one hour to stabilize the system upon small changes regarding gas flows, temperatures etc.

Space requirements

The expected space requirement for a 5,000 Nm^3 biogas/h plant is roughly 5,000-7,000 m^2 including safety distances between units but not including transformers, steam boiler, biogas pretreatment, production of hydrogen and oxygen and cooling system as shown by the battery limit in

Figure 1. The required space depends highly on the specific project including design, technology, and component choices etc. In addition, technology maturation may unveil new reactor designs with different space requirements.

Advantages/disadvantages

Advantages of biogas-based methanol compared to conventional fossil-based methanol synthesis include:

Upgrading of otherwise wasted CO_2 from biogas to valuable methanol, useful for decarbonizing hard-to-abate sectors, and reducing carbon emissions associated with methanol production

Biogas based methanol has the following disadvantages:

Limited to sites with sufficient and price competitive biogas supply, preferably close to renewable hydrogen sources
Adaption of reaction conditions due to variation of biogas composition as it varies with the feedstock and biological (uncontrolled) process
Requires potentially more thorough desulphurization compared to conventional upgrading of biogas today, which is distinctive for biogas
Requires sources of hydrogen and potentially oxygen nearby to facilitate the process e.g. through costly electrolysis
Costly due to significant demand for hydrogen and potentially oxygen, which are costly to produce through e.g. electrolysis

Environment

The final output is separated into a methanol product and effluents, most notably a stream of wastewater that can be treated by conventional methods, and off gases that can be oxidized.

Research and development perspectives

Industrial scale methanol synthesis is a well proven technology, where conventional unit operations are applicable, provided a syngas composition of mainly CO, H_2, and CO_2 is available

Research and development in biogas-based methanol currently focuses on:

- Optimizing the energy streams from and between the various reactions, including heat streams
- Testing of different biogas compositions at various conditions in the reformer
- Testing various catalysts [9]
- Upscaling from current pilot scale to industrial scale [9]
- Bringing down cost
- Verifying a complete plant at industrial scale

Examples of market standard technology

Methanol plants with full biogas utilization and hydrogen as feedstocks have been realised at pilot scale but not yet at a commercial scale, thus market standard technology is not yet available.

Financial data

Financial data are all in Euro (€), fixed prices, at the 2020-level and exclude value added taxes (VAT) and other taxes.

Prediction of performance and costs

The following elements (including installation costs) are considered to be included in the CAPEX estimations as they are inside the battery limits according to

Figure 1:

- Compressors prior to reformer and methanol reactor
- Reformer
- Methanol reactor
- Condensation unit
- Piping inside the battery limits
- Electrical cabling
- Instrumentation and controls
- SCADA
- Methanol tank

The following elements are not included:

- Production and treatment of biogas
- Electrical transformers
- High voltage electrical system
- Fee to DSO for connecting to the electrical grid
- Oxygen and hydrogen production
- Utility systems including steam boiler and cooling system
- District heat pipes and connection to the district heating system
- Treatment of purge gases and separated by-products
- Contingences
- Cost of land
- Owner's costs

A calculation example of a plant with a capacity of 0.70 MW-methanol is shown in

Table 2.

Table 2: Calculation example of a plant with a capacity of 0.70 MW-methanol.

Specifications		
Plant capacity	0.70	MW-methanol
Equivalent full-load hours	8,000.00	hours/year

Assumptions			
Cost of biogas		75.00	€/MWh
Cost of oxygen		200.00	€/ton
Cost of hydrogen		5,000.00	€/ton
Cost of eletrictrity		100.00	€/MWh
Cost of steam		60.00	€/MWh
LHV of methanol		19.90	MJ/kg
Calculations			
Methanol output	0.70 MW-methanol*8,000.00 hours/year	5,584.32	MWh/year
Methanol output	5,584.32 MWh/year/19.90 MJ/kg	1,010.23	ton/year
CAPEX	2.87 M€/MW*0.70 MW-methanol	2,000,000.00	€
Fixed O&M	114.61 k€/MW-methanol/year*0.70 MW-methanol	80,000.00	€/year
Variable O&M	4.92 €/MWh-methanol*5,584.32 MWh/year	27,457.84	€/year
Biogas input	4.10 MWh/ton-methanol*1,010.23 ton/year*75.00 €/MWh	310,336.28	€/year
Oxygen input	0.43 ton/ton-methanol*1,010.23 ton/year*200.00 €/ton	86,619.00	€/year
Hydrogen input	0.07 ton/ton-methanol*1,010.23 ton/year*5,000.00 €/ton	376,178.72	€/year
Electricity input	0.34 MWh/ton-methanol*1,010.23 ton/year*100.00 €/MWh	34,347.78	€/year
Steam input	0.15 MWh/ton-methanol*1,010.23 ton/year*60.00 €/MWh	9,273.90	€/year
Total energy input	(4.10 MWh/ton-methanol+2.48 MWh/ton-methanol +0.34 MWh/ton-methanol +0.15 MWh/ton-methanol)*1,010.23 ton/year	7,141.21	MWh/year

District heating output	0.00 MWh/MWh total input*7,141.21 MWh/year	0.00	MWh/year
Effluent gas output	0.03 MWh/MWh total input*7,141.21 MWh/year	202.05	MWh/year
Heat loss output	0.19 MWh/MWh total input*7,141.21 MWh/year	1,363.81	MWh/year
Water output	0.30 ton/ton-methanol*1,010.23 ton/year	305.86	ton/year

Data for the base year

Data for the base year is mainly based on a simulated plant by FiW Aachen University, which is based on a pilot plant in Dinslaken Germany, also by FiW Aachen University. The data provided by FiW Aachen University is supplemented knowledge shared by the Technical University of Denmark and DBI Gruppe as well as NIRAS' knowhow of conventional methanol synthesis plants and emerging biogas-based methanol plants.

Assumptions for projecting performance and costs into future years

Steam methane reforming and methanol synthesis technologies, particularly methanol reactor and distillation technologies, are well-established, thus the uncertainty on the performance estimates is low but deviations from stated values are possible, due to different system designs, improved integration and advancement of heat recovery methods as well as improved performances of machinery components in future. The accuracy of the cost figures is expected to be within ±50%.

Learning curves and technological maturity

The reformer feed with raw biogas and the complete integration of all necessary components displayed in Figure 1 are new territory, though the individual components (except the reformer) are mature technology of TRL 9. The integration of components also includes equipment tests based on raw biogas, which may yield compounds harmful for the equipment, that call for either gas treatment and/or development of equipment/technology.

The technological maturity of biogas and hydrogen to methanol is regarded to be at around TRL 5-6 with medium scale operation being the next step.

Uncertainty

The uncertainty of the cost data is ±50% to demonstrate the low TRL level of the technology concept especially concerning the reformer technology with raw biogas as main feedstock.

Acknowledgement

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98 Ammonia from Hydrogen and Air Capture

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Date	Ref.	Description
04/2023	-	Unit adjustment in datasheet regarding fixed O&M per TPD
04/2021	-	Adjustments in datasheet regarding specific energy content and minor clarifications in terms of naming

Abbreviations

1. REF	Primary reformer (=SMR)
2. REF	Secondary reformer
AEC	Alkaline electrolysis cell
ASU	Air separations unit
ATR	Autothermal reformer
BAT	Best Available Technology
BFW	Boiler Feed Water
BOP	Balance of plant (utilities)
CC	Carbon capture
CO2rem	CO_2 removal unit
DeOX	De-Oxygenation unit
DH	District heating
EIGA	European industrial gases association AISBL
EU	Electrolysis Unit
FG	Fuel gas
HC-feed	Hydrocarbon feed (normally fossil based but can also be bio-based)
HPS	High pressure steam

HSE	Health safety and environment
HTS	High temperature shift (=high temperature water gas shift)
LNH3	Liquified NH3
LTS	Low temperature shift (=low temperature water gas shift)
METH	MethanizationN2-EU Electrochemical synthesis NH3
MOF	Metal organic framework
MTPD	Metric ton per day
NH3syn	NH_3 synthesis
NH3rec	NH_3 recovery unit
NH3reg	NH3 refrigeration unit
PEM	Proton Exchange Membrane electrolysis cell
PUR	Feed purification unit
RE	Renewable power
SOEC	Solid oxide electrolysis cell
SMR	Steam Methane Reforming (typically = 1.REF)
TPD	Ton per day
TRL	Technology reediness level
WGS	Water gas shift

Qualitative Description

Today nearly all industrial production of ammonia is based on the Haber-Bosch process, where elemental nitrogen and hydrogen are combined under high pres-sure and temperature using a catalyst. Whereas nitrogen can be recovered from ambient air, the hydrogen is predominantly produced by steam reforming of natural gas (methane), a process that results in large emissions of fossil CO_2. Thus, reducing the CO_2 emissions from ammonia production is heavily linked to reducing emissions from hydrogen production. This can be achieved by capturing and storing CO_2 from conventional hydrogen production or alternatively substituting the conventional production of hydrogen with green hydrogen from electrolysis based on renewable energy.

In this chapter of the Technology Catalogue a brief description of the different NH_3 production paths is given. Thereafter, the catalogue focusses on the production of green ammonia.

Green ammonia has various applications and is primarily thought to be a carbon-neutral solution for shipping as a maritime transport fuel as well as to be used as feedstock for green fertilizers. It can potentially also be considered for applications in fuel cells, long-term energy storage, fuel for industry and peak power plants, or as an addition/mixture to conventional fuel among others.

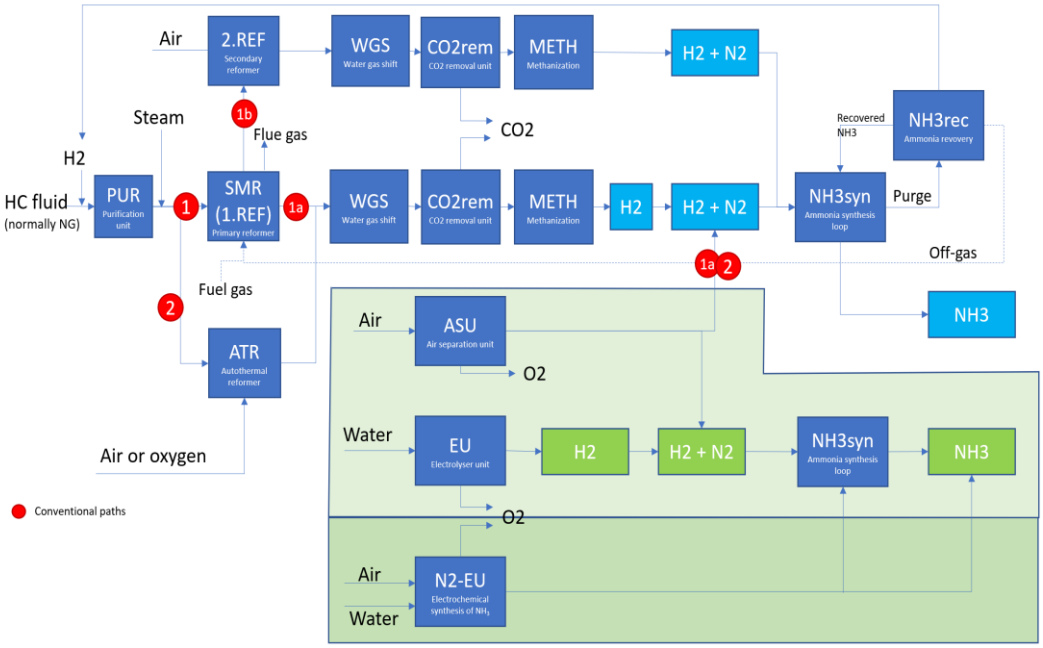
Green ammonia incorporates electrolysis for H_2 production, air separation unit (ASU) for nitrogen production and the ammonia synthesis (see light green box in Figure 1). Within this catalogue, performance and cost data are given for the ammonia synthesis. Cost and performance data for the electrolysis are given in a separate chapter within this Technology Catalogue. For ASU as a nitrogen source, cost and performance factor is given discretely within this chapter.

Brief technology description

Different production routes to ammonia, i.e. both conventional and green paths, are given in Figure 1. While the overall routes are described in subsection

Different configurations, each process step (i.e. dark blue boxes) is described in subsection *Description of each process step*.

Figure 1: Different pathways for production of NH_3



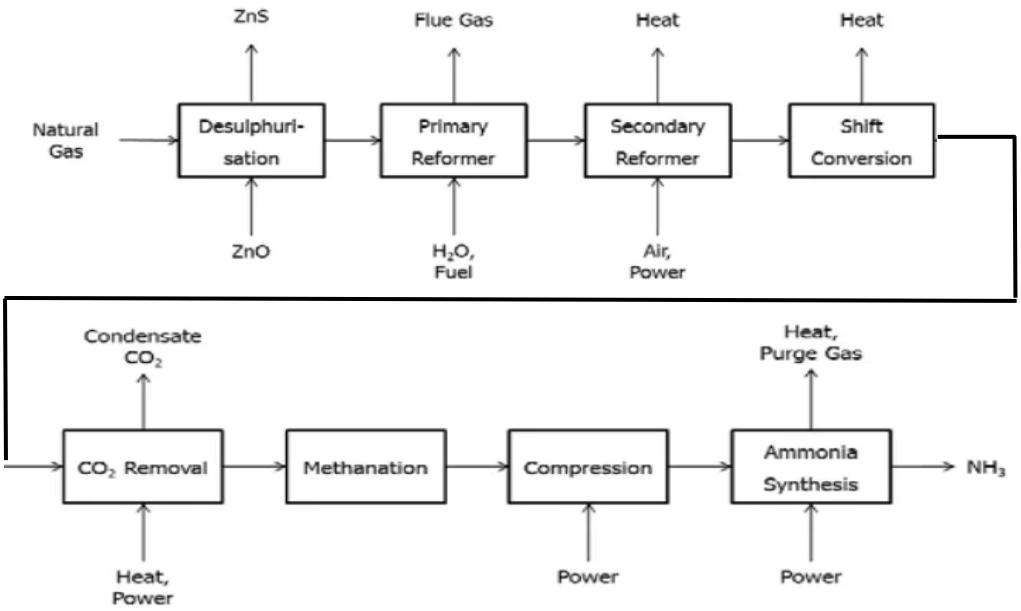
The light green area is the green NH_3 production part that is covered within this Technology Catalogue. The darker green area marks a potential future route (electrochemical). The white background shows the three conventional parts, i.e. 1a) SMR+ASU, 1b) SMR+2.REF and 2) ATR+ASU.

Different configurations

1 Conventional – grey NH_3

A conventional ammonia plant uses fossil fuels (in most cases natural gas) as its raw material. Figure 2 shows a conventional NH_3 plant based on primary and secondary reformer technology, where nitrogen is admitted via air to the secondary reformer. Alternative reformer configuration is autothermal reformer (ATR) or single steam methane reformer (SMR) combined with ASU unit to provide the nitrogen (see Figure 2).

Figure 2: Conventional ammonia plant



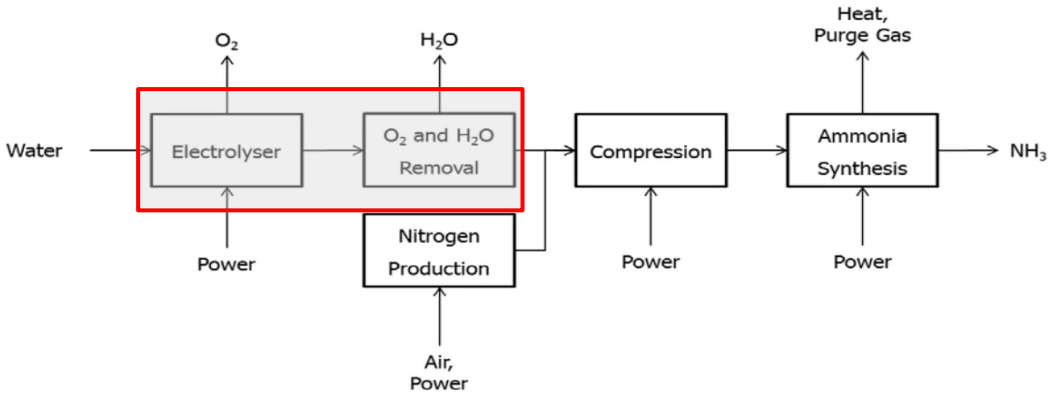
2 Conventional – blue NH_3

A blue ammonia plant is a conventional NH_3 plant with carbon capture (CC) to capture the CO_2 emissions from the reformer. This will significantly reduce the carbon footprint compared to that of grey ammonia. The raw material is however still natural gas, and the plant layout is similar to that of a conventional plant.

3 Electrolysis – green NH_3

A green ammonia plant uses green hydrogen produced via electrolysis to feed the ammonia synthesis loop (see Figure 3). The electrolysis shall be powered with renewable energy such as wind power.

Figure 3: Green ammonia plant



Any impurities of O2 in the H2 product is removed by reacting it with H2 over a DeOX (de-oxygenation unit).

4 Electrochemical synthesis of ammonia – green NH_3

Direct production of NH_3 by electrocatalytic reaction of water and air, i.e. eliminating the Haber-Bosch process, may become an alternative process for green NH_3 with use of renewable electricity. This technology is still only at research level (see further description below under *4) Autothermal reforming (ATR)*.

Description of each process step

[1] Feed purification (PUR)

The feed purification section removes impurities (sulphur, chlorine and heavy metals) that are poison to downstream catalyst.

The purification section typically consists of two reactors; The first one is a hydrogenator that converts organic sulphur (and chlorine) into H_2S (HCl) via the following reaction:

$$R\text{-}S + 1.5\ H_2 \xrightarrow[300 - 450\ °C]{\text{Catalyst}} R\text{-}H + H_2S$$

The second one is a downstream absorber that removes H_2S (and HCl) from the feed via absorption.

[2] Steam methane reforming (SMR)

Steam methane reforming is a method for producing hydrogen from hydrocarbons and steam via the following reactions:

$$C_nH_m + nH_2O \Rightarrow nCO + (n+1/2m)\ H_2 - Q$$

$$CH_4 + H_2O \Rightarrow CO + 3H_2 - Q$$

$$CO + H_2O \Rightarrow CO_2 + H_2 + Q$$

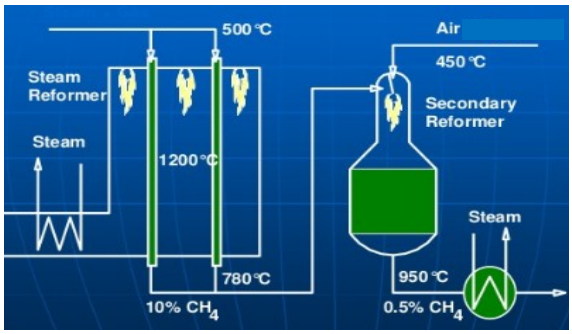
Typical outlet conditions						
Temp.	Pres.	CH_4	CO	CO_2	H_2	N_2, Ar
800-920 °C [10]	20-50 barg	5-10 dry %	15-20 dry %	5 dry %	70 dry %	< 1 dry %

The reforming reactions are highly endothermic, meaning that heat must be added. The SMR typically consists of several catalyst-filled tubes to which heat is added either via a fired radiation box or via convection with a >1000°C hot flue gas. In both cases, fuel must be added to provide the heat. The temperature of the flue gas leaving the reformer depends on the technology applied but is in the 900-1150°C range.

[3] Secondary reformer (2. REF)

Figure 4: Primary and secondary reformer arrangement

[10] If followed by secondary reformer, the outlet temperature (T_{out}) is ~800°C, while it is ~900°C if there is no downstream reformer



Typical outlet conditions							
Temp.	Pres.	CH_4	CO	CO_2	H_2	N_2	Ar
900-1000 °C	20-50 barg	<0.5 dry %	15-20 dry %	5-10 dry %	50-60 dry %	20-25 dry %	<0.5 dry %

Nitrogen for the NH_3 production can either be added via an ASU or via an air-fired secondary reformer. Within the secondary reformer, heat for the reforming process is provided by burning some of the syngas inside the reactor with admitted air. By adding air, N_2 is added, meaning that an ASU is not needed.

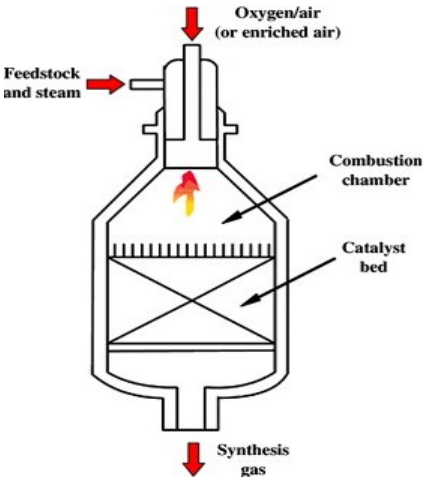
The reaction scheme is:

$CH_4 + ½\ O2 => CO + 2H_2$

The addition of air is controlled such that the hydrogen to nitrogen ratio at the inlet to the ammonia loop is approximately 3:1, which is the required stoichiometric ratio for the ammonia reaction.

[4] Autothermal reforming (ATR)

Figure 5: Autothermal reformer (ATR)



Typical outlet conditions							
Temp.	Pres.	CH_4	CO	CO_2	H_2	N_2	Ar
1000-1100 °C	20-50 barg	<0.5 dry %	15 dry %	5-10 dry %	50 dry %	25 dry %	<0.5 dry %

Like SMR, ATR can be used to convert hydrocarbon feed into a hydrogen rich syngas.

Within ATR, heat for the reforming reaction is provided by burning part of the syngas inside the reactor commonly with pure oxygen. This gives the following reaction scheme:

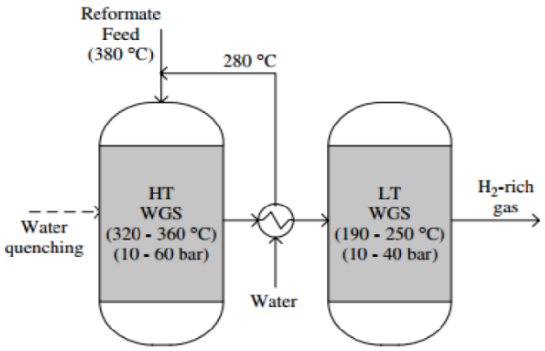
$$2\ CH_4 + O_2 + CO_2 \Rightarrow 3\ H_2 + 3\ CO + H_2O$$

$$4\ CH_4 + O_2 + 2\ H_2O \Rightarrow 10\ H_2 + 4CO$$

The advantage of ATR is that the product H:CO ratio can be varied, depending on the amount of steam and oxygen (O_2) added.

[5] Water gas shift (WGS)

Figure 6: Typically shift configuration in an ammonia plant



Typical outlet conditions							
Temp.	Pres.	CH_4	CO	CO_2	H_2	N_2	Ar
~ 160 °C	20-40 barg	<0.5 dry %	<0.5 dry %	20 dry %	60 dry %	20 dry %	<0.5 dry %

The purpose of the shift reactor(s) is to produce additionally hydrogen (H_2) by converting CO via the following reaction:

$$CO + H_2O \Leftrightarrow CO_2 + H_2 + Q$$

As the shift reaction is exothermic, low temperature favors a low equilibrium content of CO. However, a low temperature also decreases the reaction rate.

To ensure fast conversion and at the same time low CO slip, the shift section can be a series of shift reactors with interstage cooling (see Figure 6). A conventional ammonia plant typically includes a high temperature shift (HTS) and a low temperature shift (LTS).

[6] CO_2 removal unit (CO2rem)

The CO_2 in the syngas from the WGS must be removed before the syngas is admitted to the NH_3 synthesis.

The reason is that CO_2 is poisonous to the NH_3 catalyst. Alternatively, all CO_2 could be converted to methane in downstream methanization reactor (see methanization step in next subsection), but this will create a huge amount of inert (CH_4 and Ar are inert in the ammonia loop) in the NH_3 loop that must be compressed and purged out.

The CO_2 removal unit is typically based on amine absorption technology. Other applied technologies are Selexol, Benefield and Vetrocoke.

Typical outlet conditions							
Temp.	Pres.	CH_4	CO	CO_2	H_2	N_2	Ar
~30 °C	20-50 barg	<0.5 dry %	<0.5 dry %	0.05 dry %	75 dry %	25 dry %	<0.5 dry %

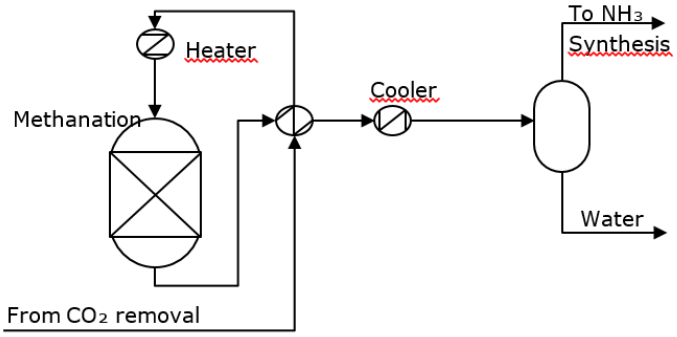
[7] Methanization (METH)

The methanization process aims to remove any residual CO and CO_2 (as they are poisonous to the ammonia catalyst) from the feed stream before it enters the ammonia synthesis reactor.

$$CO + 3H_2 \rightarrow CH_4 + H_2O + Q$$

$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O + Q$$

Figure 7: Methanization, cooling and water separation



Typical outlet conditions							
Temp.	Pres.	CH_4	CO	CO_2	H_2	N_2	Ar
~30 °C	20-50 barg	< 1 dry %	< 5 ppm	< 5 dry ppm	75 dry %	25 dry %	<0.5 dry %

[8] Ammonia Synthesis

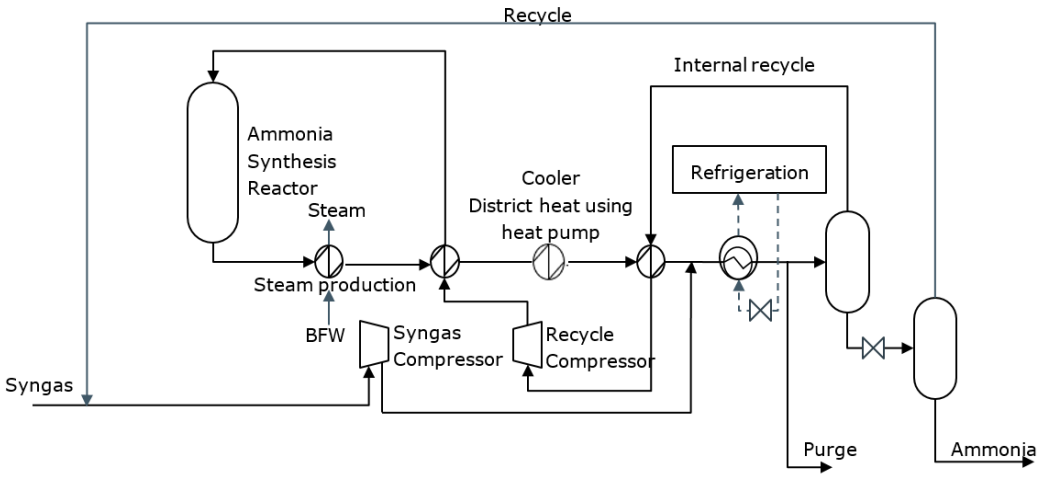
The hydrogen and nitrogen feed stream are compressed and admitted to the ammonia loop (referred to as the Haber-Bosch process).

$$3H_2 + N_2 \rightarrow 2NH_3 + Q$$

This ammonia reaction is highly exothermic, and the heat produced is used to generate steam. The steam generated is an export from the ammonia synthesis loop. In a conventional plant, some of the steam is used for hydrogen production in the steam methane reformer (SMR) and some for power generation in steam turbines.

The conversion rate is typically only ~25 % per single pass, so a large internal recycle is required to ensure high overall conversion.

Figure 8: Ammonia synthesis loop and downstream purification



Typical ammonia outlet conditions		
Temp.	Pres.	NH_3
-10 - 0	20-25 barg	> 99 %

The ammonia synthesis benefits from a high operating pressure. Depending on the technology provider the loop usually runs at anything between 150 to 250 barg. A common overall loop pressure drop is approximately 10 bar. Temperatures in the loop range from 350˚C to 550˚C.

The steam from the ammonia reactor is cooled, chilled and condensed. The condensed ammonia is separated from unreacted reactant first in a high-pressure vessel and then in a 20-25 barg vessel. The unreacted reactants are recycled back to the process.

[9] Electrolysis

For electrolysis reference is made to the existing catalogue chapter [4].

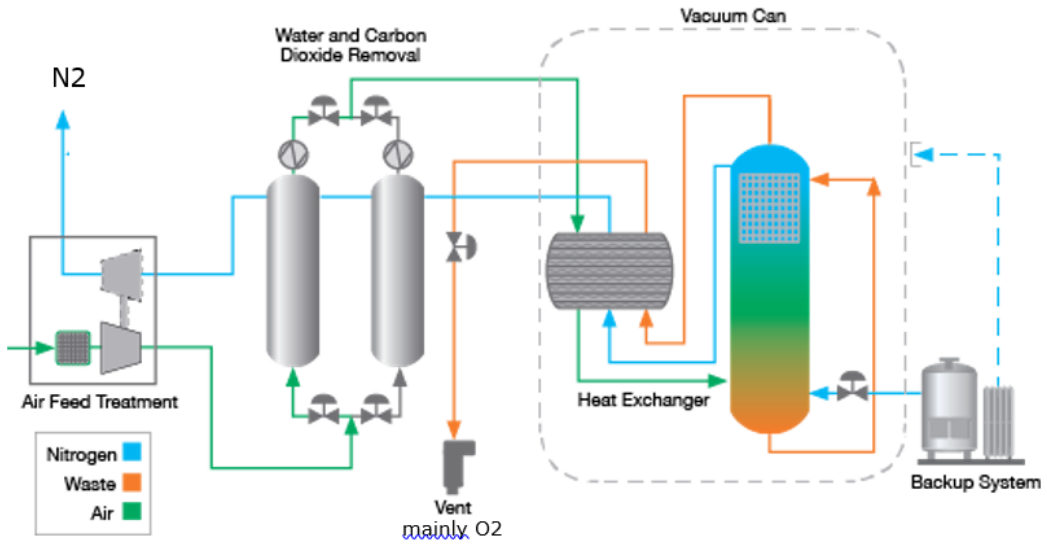
[10] Air separation unit

Pure nitrogen is required as feedstock for the Haber-Bosch synthesis of ammonia, as shown in Figure 1. Pure nitrogen is produced by an ASU, which uses a cryogenic distillation process to separate ambient air into nitrogen, oxygen, and argon.

Figure 9 shows a flow-diagram for a typical ASU configured for nitrogen production. Ambient air is compressed and dehydrated before it is chilled by heat exchange with the cold liquid $N2/O_2$

products from distillation. Final chilling is obtained by expansion of the air. The distillation column will separate liquid nitrogen from liquid oxygen and argon. The ASU does not produce any usable heat.

Figure 9: Schematic flow diagram of nitrogen plant.



Source: Source: AIChE

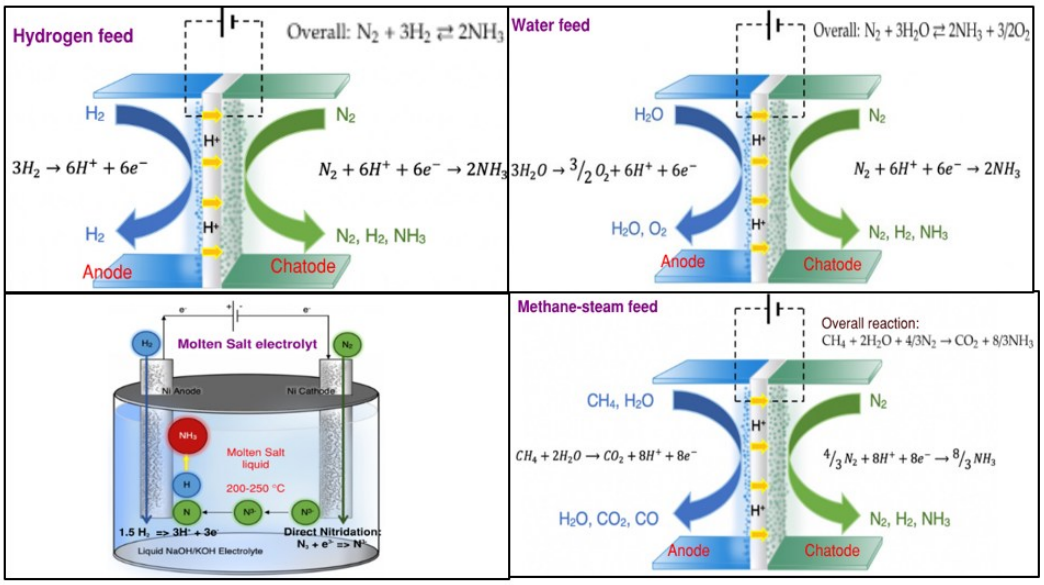
The ASU will deliver highly pure nitrogen (>99.9%), but can also be configured to coproduce pure oxygen, which may be used in the production of grey/blue hydrogen if the ammonia plant employs autothermal reforming (ATR).

[11] Electrochemical synthesis of ammonia

Direct electrochemical synthesis of ammonia from N_2/air and water by use of renewable electricity is an interesting alternative, as it avoids the Haber-Bosch process and potentially also the air separation. The electrochemical synthesis of ammonia is a process path that has been under development for the past 20 years and many different configurations are being examined.

Several studied paths exist. Some key paths are summarized in Figure 10.

Figure 10: Illustration of the anode and cathode reactions during electrochemical NH_3 production. [1]



The different approaches to electrochemical ammonia production that have been studied can be divided into a low and high temperature path:

Low temperature path (<100⁰C): This is typically conducted in an aqueous cell, where the aqueous solution is both the hydrogen source and acts as the electrolyte. Different aqueous electrolyte solvent and different catalytic materials have been investigated (Fe2O3, MOF (Au, Fe, Cu), Ni, etc.) to maximize efficiency and reaction rate. However, at low temperature only very low reaction rates have been achieved.

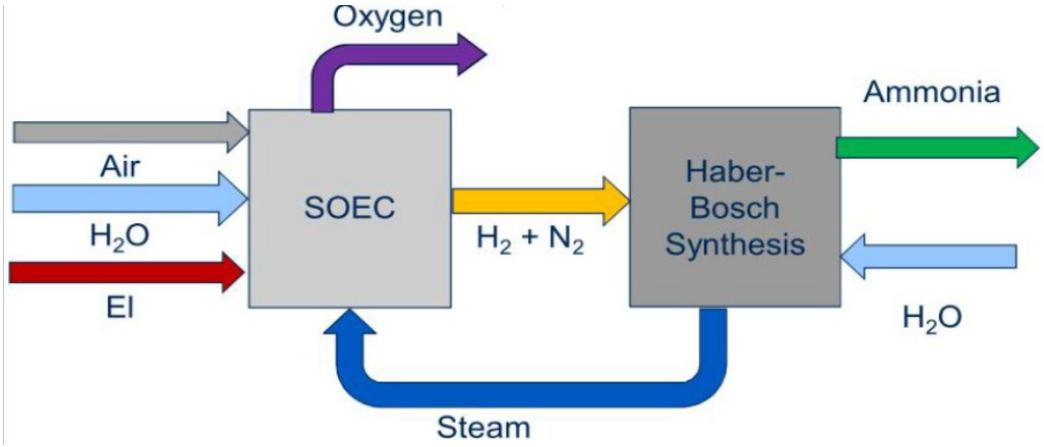
High temperature path (>100⁰C, typically 200-650⁰C): The high temperature path typically applies a solid-state electrolyte or a molten salt. The hydrogen source can be hydrogen itself, steam or methane-steam mixture. The main advantage of the high temperature path is that significantly higher reaction rates are achieved. However, the efficiency is lower. A major disadvantage with higher temperature is the competing hydrogen formation reaction and decomposition of the NH_3 product which start above 250°C and is dominating at 500°C [1].

Results indicate an inverse relationship between efficiency and reaction rate. Hence high temperature and the catalysts that provide the higher reaction rate tend to provide the lowest efficiency. The achieved reaction rates and efficiencies today are still far too low for practical application [1, 2], hence this process will be decades away from commercialization. The TRL is judged to be 1-2.

[12] Combining electrolysis and ASU in SOEC

Power2Ammonia is a EUDP-funded project for production of ammonia synthesis gas in an SOEC core.

Figure 11: SOC4NH3 project combining electrolysis unit and ASU (El = power) [15]



Within the electrolysis, the component that is transported from the anode to the cathode is:

[1] Alkaline electrolysis (AEC): OH^-
[2] Polymer exchange membrane electrolysis (PEM): H^+
[3] Solid oxide electrolysis (SOEC): O^{2-}

As the solid oxide transports O^{2-}, it might be used to separate O_2 from the air. Thus, the ASU can be eliminated. The EUDP project has demonstrated 50 kW operation, i.e. the TRL is low (4-6).

Operation range

The operation ranges of both conventional and green ammonia plants can be divided into:

1) Shut down – Cold standby
2) Hot standby – no production but plant is kept warm for fast startup
3) 0-20 % operation
4) 20-35 % operation
5) 30-100 % operation

1) Shut down – cold standby

Shut down/cold standby is when the plant is shut down and cooled to ambient temperature.

Cold shut down should generally only be used for maintenance. It should not be used for a short stop of the plant due to no product demand or missing feed availability. The reason is that frequent cooling and reheat will cause catalyst crunching (due to grinding among catalyst particles caused by expansion upon heating and shrinking upon cooling) and thereby reduced catalyst lifetime. Therefore, an ammonia plant must be kept in hot standby mode (see next section) in periods, in which it is not in operation.

2) Hot standby mode

Hot standby mode is an operation mode where there is no production but almost all units are kept at normal operation condition (i.e. at normal operation temperature and pressure) to enable a fast ramp up in capacity.

For an ammonia plant, the hot standby mode depends on the duration, i.e. it may be still-standing for a couple of days, while circulating hot gas may be used to keep the reactor warm for prolonged periods (weeks). Starting an ammonia loop from cold conditions can take up to one day, while ramping up from hot conditions is usually ~2 hours. Hot standby mode requires no feedstock. The energy that needs to be added during prolonged hot standby will be equal to heat loss to the surroundings, which will be very little if the plant is properly insulated. For start-up, a start-up heating system is needed anyhow, so the additional capital investment for facilitating a hot stand-by mode will be very minor.

Similar, for an electrolysis unit, a hot standby mode can enable fast ramp up (within seconds)11. Depending on weather forecast and knowledge about fluctuations in electricity generation and demand, the number of electrolysis cells that is kept in hot standby mode can be optimized.

3) Operation at 30 – 100% Capacity

A conventional ammonia plant usually has an operating capacity of 70-100%. However, as general turndown ratio of rotating equipment, many transmitters and control valves are 30%, these plants can normally handle loads down to 30% without mayor changes.

4) Operation at 20 – 35% Capacity

If there is a need to reduce operations to 20%, this can usually be achieved by additional CAPEX spending to buy equipment that can handle larger capacity ranges.

5) Operation at 0 – 20% Capacity

For operation at lower capacity than 20% a significant increase in CAPEX can be expected, as multiple valves, instruments and rotating equipment would have to be purchased to manage the wide range of operating loads.

Demand for operation flexibility

The requirements for operation flexibility depend strongly on the feed availability (power or hydrogen) and on requirements for product flow.

If the feed is hydrogen, i.e. the plant is connected to a hydrogen transmission net, the buffer within the hydrogen transmission net will ensure a stable feed flow, which cost is fairly stable. Thus, the demand for the operation flexibility will be low.

Alternatively, if the feed is power, i.e. hydrogen is produced by electrolysis of water, fluctuating power prices and the wish to maximize earning naturally imposes some desire for high flexibility in the capacity of the ammonia plant.

[11] Typical power connection requirement is: 50% ramp in power supply within <5s and 100% ramp in power supply within 30s (if a connection can guarantee this connection requirement, a higher price is given)

Regarding fluctuating power prices, the following scenarios must be considered:

1. Fast ramping: Grid connections that facilitate fast ramping cost less. The electrolysis unit is able to provide FCR[12] (West Denmark) and FCR-D (East Denmark) connection
2. Prolonged periods with high power prices

Point A: The ammonia synthesis cannot ramp as fast as the electrolysis unit. However, minor "hydrogen + nitrogen" storage can ensure that a green ammonia plant can fulfill point A.

Point B: It is crucial that the temperature within the ammonia reactor is kept constant as frequent cooling and reheat will cause crunching, whereby the catalyst lifetime is reduced. To maximize earnings under prolonged periods with high power prices and at the same time ensure a constant temperature in the ammonia reactor, the following design options (or a combination of them) can be applied:

2. Design NH_3 plant with large operation range + additional NH3 storage. Periods with high price of power can be optimized by ramping down the capacity of the NH_3 plant and even put it into a hot standby mode. This can be combined with additional NH3 storage13 to fulfill any contractual requirement on a minimum ammonia production rate.
3. Locate next to a hydrogen transmission net: As stated above this will minimize fluctuating feed cost.
4. Hybrid NH_3 plant. Combining the green NH_3 production with existing conventional NH_3 production will make it possible to ramp up the load of the reforming section when the power prices are high.

The most optimal option depends on the circumstances. The location next to a hydrogen transmission net or next to an existing conventional ammonia plant are likely to be the most cost-efficient solutions. The disadvantage of point 1 (and to some extent also to point 3) is that the huge capital cost of an ammonia plant normally requires >90% load to pay back the capital expenses.

Properties of ammonia

Thermal and physical properties of gaseous ammonia:

Property	Ammonia
Molecular weight	17 kg/kmol
Normal density	0.77 kg/m³
Lower heating value, LHV	19 MJ/kg

[12] FCR=Frequency containment reserves,

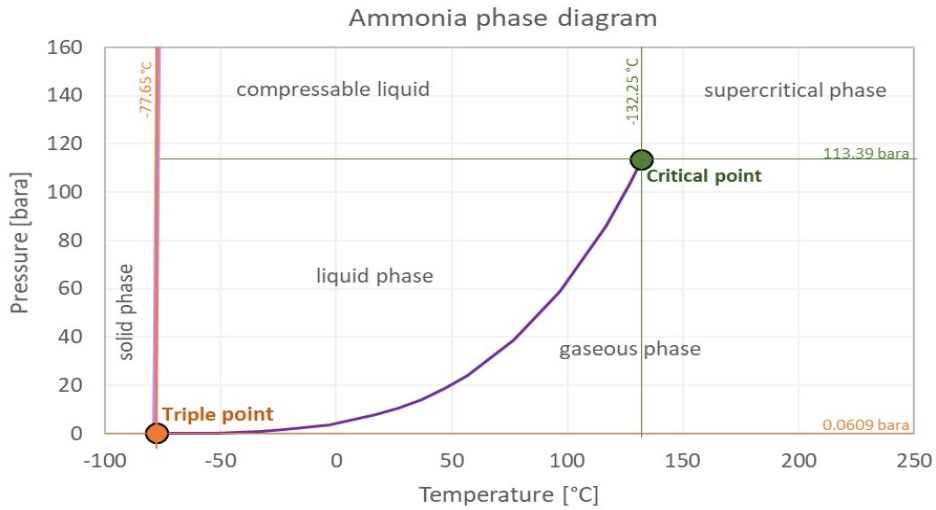
FRD =Ramp 100% within less than 30s (west DK).

FCR-D =Ramp 50% within 5s and 100% within 30s (east DK).

[13] Note: NH_3 storage is much cheaper than H_2 storage

Higher heating value, HHV	23 MJ/kg

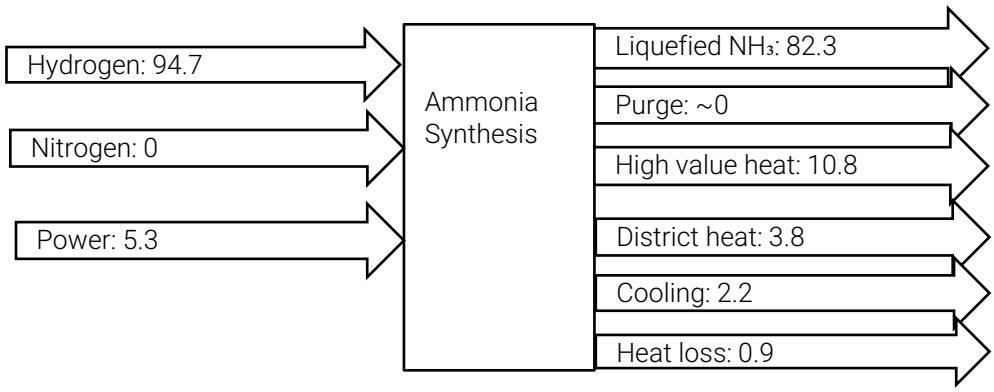
Figure 12: Phases of ammonia



Mass and energy balance

This section gives an overview of inputs and outputs of an ammonia synthesis in energy (e.g. MWh).

Figure 13: Overall energy balance of the ammonia synthesis. The electrolysis unit is covered in [4] and power required for the ASU is listed in Table 1.



Input

The input to the ammonia synthesis loop is hydrogen, nitrogen and power as per

Figure 13. Including the electrolysis unit and the air separation unit (ASU), the input streams are water, air and power.

Boiler feed water (BFW) is normally used to extract the high value residual heat from the ammonia synthesis reaction, while cooling water is used to remove the low-calorie residual heat.

Output

The output streams from the ammonia synthesis is ammonia, steam, hot water (district heating water) and a tiny purge stream.

The high value heat can be used to make steam at different levels. The steam can be converted to high pressure steam and used within the plant to drive the compressors. The steam that is not used to power the process can be exported.

The purge is needed to remove any accumulated impurities, but as the feed stream is almost 100% pure H_2 and N_2, the purge will be insignificant. As the purge contain impurities of NH_3, it must be burned off as a fuel or sent to a flare.

Including electrolysis unit and ASU, the overall output streams are, besides the above mentioned, oxygen from the electrolysis unit and oxygen from the ASU.

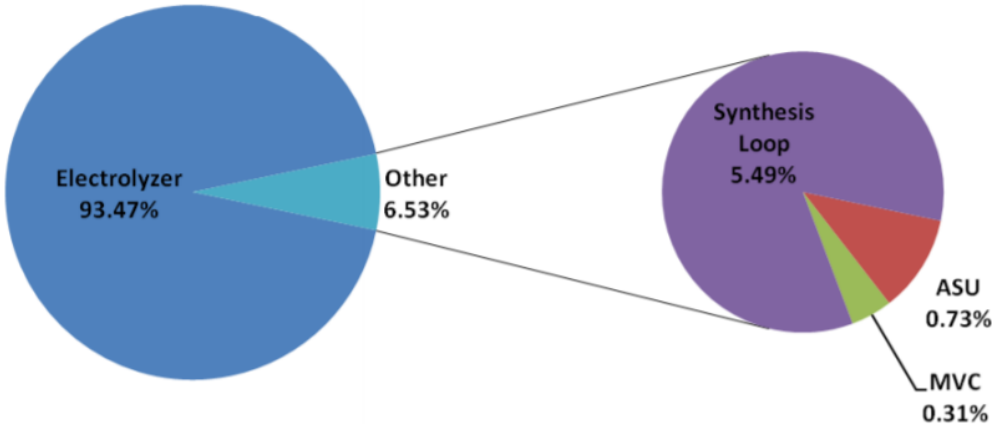
Energy balance

The energy balance of the ammonia synthesis is given in

Figure 13.

Energy balance of green ammonia plants including electrolysis unit and ASU: Many theoretical papers and studies have looked into the energy requirements of green ammonia plants. These vary greatly compared to a conventional ammonia plant, as the power consumption of the electrolysis unit makes up a large majority of the overall plant power requirements. Figure 14 shows an example of the energy balance breakdown of a green ammonia plant (operating at 150 bar), where the synthesis loop is the power required to drive compressors and pumps.

Figure 14: Power requirement breakdown for a green ammonia plant



Ann: (MVC=Mechanical Vapor Compression, includes pumping and desalination of feed water to electrolysis unit) [7].

Based on the authors' estimate the following energy consumption breakdown was found (the MVC above has not been included as it is minor).

Table 1: Major power consumption units in green ammonia plant

Plant area	Power consumption	%
ASU	250 kWh/t N_2	2.1%
Electrolysis unit (65% eff.)	9350 kWh/t NH_3	94.4%
Syngas & Make up compressor	290 kWh/t NH_3	2.9%
Ammonia refrigeration	50 kWh/t NH_3	0.5%
TOTAL	9900 kWh/t NH_3 (36 GJ/t NH_3)	

Ann: (within the sum the power in ASU is converted to power per NH3 instead of power per N_2 – see calculation in the subsection Datasheets)

The energy consumption of ASU's (200-400 kWh/ton N_2) depends on capacity, extent of integration, and whether a high share of liquid N_2 for back-up should be produced.

Integration possibilities

Integrations within conventional NH_3 plants are:

1. Steam produced in ammonia loop is normally used for:
   1.1. Steam addition to the steam methane reforming process
   1.2. Power production for the compressors and the pumps in the ammonia plant
   1.3. Export of steam
2. Recovered hydrogen in ammonia recovery unit (NH3rec):
   2.1. Is used as hydrogen required for hydrogenation within the feed purification section
3. Recovered fuel gas (i.e. off-gas) from ammonia recovery unit (NH3rec):
   3.1. Use as fuel for the steam methane reformer (SMR)
   3.2. If no steam methane reformer (SMR), normally exported

Integration possibilities within green NH_3 plants are:

A. Steam produced in ammonia loop can be used for:
   a. Electrolysis: In case of SOEC (future option), which operates at 700-900 °C, steam can be used as water for the electrolysis unit
   b. Power production for own consumption e.g. ASU, compressors and pumps in the ammonia plant
   c. Export of steam (for use in nearby industrial processes)
B. Oxygen from the electrolysis unit and the ASU:
   a. Export
C. Excess low-temperature heat
   a. Water/air coolers in electrolysis and ammonia loop can be used for district heating especially if combined with a heat pump

Hybrid NH_3 plant

If ammonia is becoming a transportation fuel (which is most likely the case), then the ammonia market will increase substantially. Thus, there will be an interest in increasing the capacity of existing ammonia plants and/or make them greener, which both can be accomplished by adding an electrolysis unit to the existing plant. Whether the secondary reformer can cope with the increased N_2 demand or an ASU needs to be added, will depend on the demand for increased capacity.

If the purpose of adding an electrolysis unit is to increase the capacity, it will normally be done by identifying the bottlenecks of the existing ammonia plant and replace the units (or add additional units) that inherit the bottlenecks. The capacity can usually be increased to 110% capacity with no or very minor changes. Increasing the capacity with 20-30% can often be done with acceptable investments (as only few equipment needs to be revamped/replaced), while larger capacity increase will require major investments as almost all items need to be replaced.

If the conventionally grey reforming section and the new green "ASU and electrolysis unit" section should be able to operate independently, i.e. without the other in operation, major integrations are not possible. Independent operation will be used if "ASU and electrolysis unit" is shut down when the power prices are high.

Integration possibilities within hybrid NH_3 plants:

1) Steam produced in ammonia loop can be used for:
    a. Steam addition to the steam methane reforming process
    b. Electrolysis: In case of SOEC (future option), which operates at 700-900 °C, steam can be used as water for electrolysis unit
    c. Power production for own consumption e.g. ASU, compressors and pumps in the ammonia plant
    d. Export of steam
2) Oxygen from the electrolysis unit and the ASU:
    a. Feeding the secondary reformer with enriched air: The capacity of the secondary reformer can be increased by feeding it with enriched air, as extra feeding duty (via partial combustion of feed gas with oxygen) can be obtained without having to add excess nitrogen [9]
    b. Export
3) Hydrogen and off-gas from the NH3rec units: Same as under conventional NH_3 plant
4) Excess low-temperature heat
    a. Water/air coolers in electrolysis and ammonia loop can be used for district heating especially if combined with a heat-pump
    b. Heat from electrolysis unit can be used for pre-heating of NH_3 recycle

A key feature of the electrolysis unit is that it can provide hydrogen for start-up. The feed purification section needs hydrogen, which is recycled from the downstream system, but as the downstream system is not in operation when starting the plant, imported H_2 is needed for conventional plants. This will not be the case for green or hybrid plants.

Quantitative Description

Typical capacities

The typical capacity of conventional ammonia plants erected today is in the range of 1000 to 3500 TPD of ammonia for a single line.

For green ammonia production, the size of the electrolysis unit or the available renewable electricity will set the limit for how large the units can be.

Regulation ability

For plants based on intermittent renewable energy, one (or a combination) of the following options must be selected

1. A turndown ratio of 0% (hot standby mode)
2. Possibility to use grid power
3. Possibility to take feeds (N_2 and H_2) from grid or storage

4. Possibility to increase capacity of a conventional front-end (hybrid solution)

Several technology providers have quoted the following figures for turndown.

Haldor Topsoe: 10-100% [6]

ThyssenKrupp: 30-100% (vendor info)

Casale: 20-110% (vendor info)

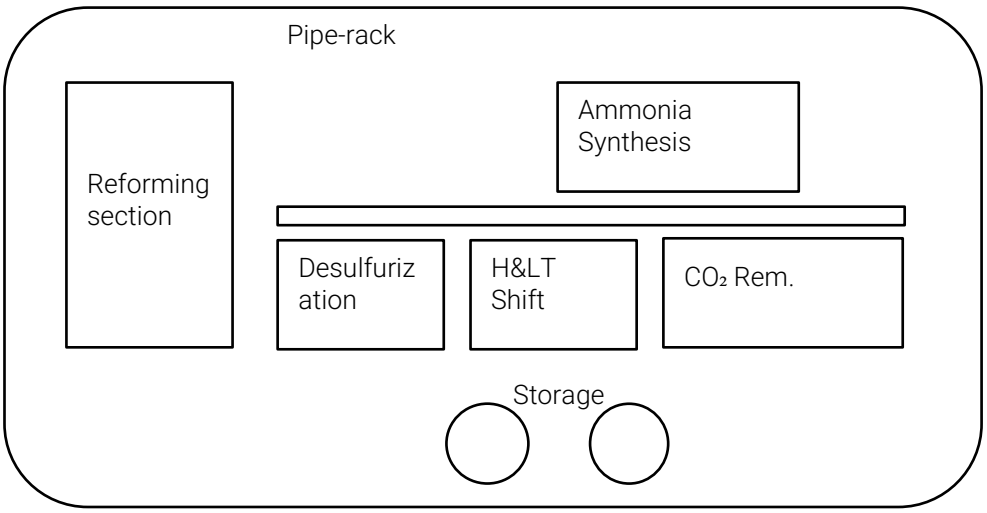
KBR: 30-100% (vendor info)

Space requirement

Looking at conventional ammonia plants the plot space required for a production capacity of 1390 TPD is around 150 x 100 m. This includes all operation buildings but not storage facilities. The actual placement of processing areas within the plant is not critical, as long as industry safety rules are followed.

A typical ammonia plant (with secondary reformer and no ASU) may have the following layout:

Figure 15: Typical ammonia plant layout



For a green ammonia plant, only the ammonia synthesis section of the plant is relevant, with the other areas removed to make space for electrolysis unit and ASU. The electrolysis unit and the ASU require less plot space than the reformer, desulphurization, shift reactors and CO_2 removal unit. Qualitatively speaking a green ammonia plant would require a smaller plot area than a conventional one for the same capacity.

Advantages/disadvantages

The main advantages of green ammonia production relative to conventional ammonia are:

1. No fossil fuel (natural gas) is required; hence production can be made CO_2 emission free
2. Location is not bound to areas/regions where inexpensive natural gas is available
3. N_2 and H_2 feedstocks are pure, which reduce purging requirement and need for NH_3 recovery section. This increases the overall efficiency of the NH_3-synthesis

4. Capacity variation can contribute to an increased flexibility in power consumption, i.e. if power production is high, power utilization can be increased. This will increase the average utilization factor (load factor) of the renewable power production units

The main disadvantages can be summarized as:

1. Fluctuations in renewable power generation leading to fluctuations in the operating profile will reduce the average utilization factor (load factor).
2. Today the cost of hydrogen produced via electric power is significantly higher than that of natural gas, which gives higher costs of green NH_3.

Safety and environment

Key HSE (Health, Safety and Environment) concerns to consider in an ammonia plant are:

1. Ammonia: Is a toxic component [3]
2. Hydrogen: Is a highly flammable and explosive component [3]
3. Leakage
4. High pressure equipment
5. Chilling unit
6. Hot surfaces

Research and development perspectives

The Haber-Bosch process for ammonia synthesis is a mature process that has been in use in the industry for 100 years. The process has undergone significant improvements over the years; hence it is believed that future improvements and cost reductions will be marginal.

It is expected that the electrolysis technology for H_2 production will improve substantially towards 2050. In a ten-year timeframe it is expected that H_2 can be delivered at high pressure directly from the electrolysis unit. This will lead to reduction of CAPEX and electricity consumption for feedstock compression to green ammonia synthesis.

Successful combining electrolysis unit and ASU in an SOEC or successful development of the electrochemical process for low temperature production of ammonia from air and water may be potential game changers. However, as mentioned in the section describing the process step of the Air separation unit, the processes are far away from commercial application today.

Examples of market standard technology

Only few NH_3 plants with electrolysis units are operational today.

One plant is the Pilot plant in Minnesota (operation since 2013) which output is 25 ton/year green ammonia. The electrolysis unit is powered by wind [11].

Yara is developing a hybrid solution at their Pilbara ammonia plant in Western Australia. The plan is to erect a 100 MW solar farm to drive a 50-60 MW electrolysis unit, which will increase the production of ammonia from the existing Haber-Bosch unit with ~80 TPD [12]. The engineering for the tie-in of green hydrogen was completed in 2018 and expected date to come online is 2021. It is planned that the plant will expand its green ammonia production in stages up until 2030 when an expected 90% of its production will come from green sources.

Yara are also partnering with Orsted to develop a 100 MW electrolysis plant to produce green hydrogen for ammonia production in Holland. This is expected to be operational in 2024/2025 and will produce approximately 200 TPD of green ammonia [13].

Air Products have recently announced investment in a new green ammonia facility to be operational by 2025 at the industrial hub of NEOM in Saudi Arabian. Using Haldor Topsoe technology, the 4 GW plant will produce 650 TPD of green hydrogen, an equivalent of 3250 TPD of green ammonia [14].

In Denmark, near Lemvig, a new green ammonia plant is planned to start operating in 2022 producing 5000 ton/year green ammonia. The project is a collaboration between Skovgaard Invest, Haldor Topsøe and Vestas [16].

Financial Data

Prediction of performance and costs

Investment cost (CAPEX)

Predictions of investment costs for green ammonia plants are based on data from the industry as a whole, as no large-scale plants or projects have been completed or are in operation yet.

Overall plant cost data has been broken down into major plant section to get a distribution of cost. Several different sources have been compared and values for each section of the plant have been determined from this.

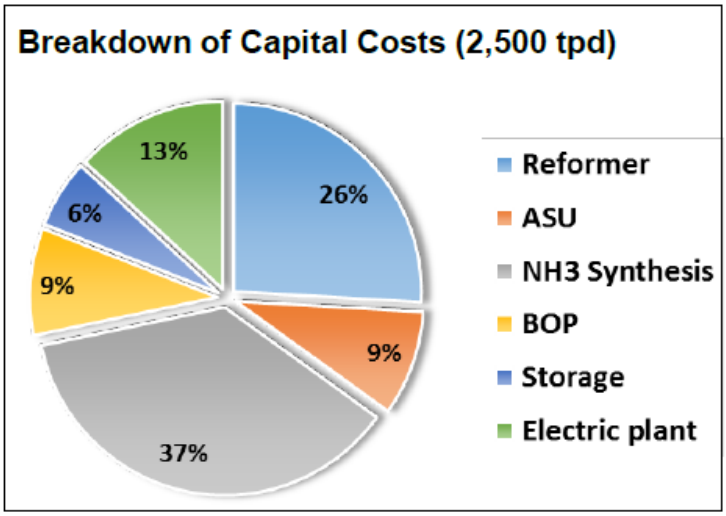
For the investment cost analysis, the following sections of the ammonia plant are included:

I. Ammonia synthesis
II. Balance of plant (BOP)[14]
III. Storage

Based on cost split data for various conventional plants and figures obtained from various vendors, average split factors have been estimated. This approach gives that the average cost of the ammonia synthesis including storage and BOP is around ~54% of that of the cost of conventional NH_3.

[14] Balance of plant (BOP): BOP typically is surrounding utility, storage, startup and shut down facility. There is often variation in what BOP includes. Here storage and electric plant is listed separately, meaning they are not included in the BOP.

Figure 16: Cost split on main systems in a conventional NH3 plant from Linde [0].

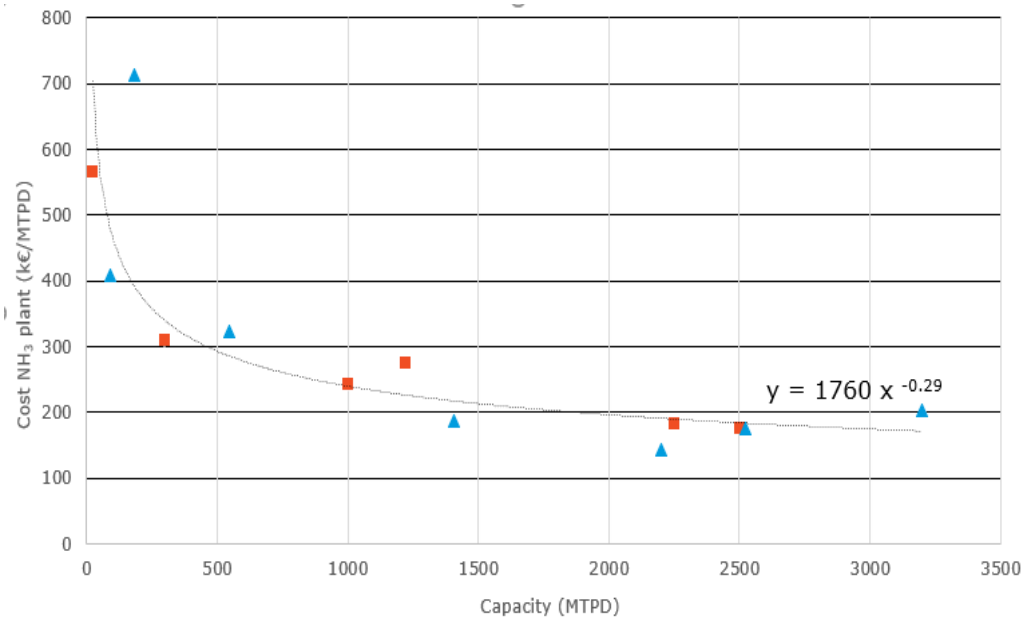


Ann: Reformer, electric plant (power plant) and some balance of plants (BOP) will disappear for a green NH_3 plant. Cost of green NH_3 plant (excl. electrolysis unit and ASU) is therefore taken as 54% of conventional plant, based on all obtained data.

In

Figure 17 a cost-capacity curve for specific CAPEX of green ammonia plant (k€/TPD) has been derived using cost data of conventional ammonia plants at different capacities. The 0.54 factor explained above has been used to remove ASU and scope not relevant for green ammonia. All costs in the figure have been scaled using the Chemical Engineering Plant Cost Index (CEPCI) to reflect 2019 costs.

Figure 17: Estimated cost of an ammonia plant



Ann.: Ammonia synthesis + BOP + torage (electrolysis unit and ASU are not included in the figure). Blue triangles represent data that is publicly available. All figures adjusted to reflect cost index for 2019.

It is observed that at low capacities (<300 TPD) there is a steep increase in the specific CAPEX. As the design hours, construction time and the amount of metal used per unit capacity is much larger for small plants than for large plants, customized small plants will always be much more expensive than large plants. However, skit-mounting and mass production can change this picture substantially, i.e. the steep increase for small capacities shown in

Figure 17 may decrease substantially if a market for small ammonia plant comes forward. But it is questionable whether a larger market for small ammonia plants will develop, as the advantages of having distributed ammonia plants is limited. In order to add the cost of the ASU, a multiplication factor of 1.06-1.09 must be added to the total cost of the ammonia plant sections as listed above.

Variable and Fixed O&M costs

Fixed operating and maintenance costs are taken as 3% of CAPEX.

Variable operating and maintenance costs are taken as costs of catalyst replacement and other minor consumables. Catalyst replacement is scaled based on a reference for a 1500 TPD NH_3 plant with 10 m3/year [5]. Iron catalyst price is assumed to be 3000 EUR/m3.

Uncertainty

With respect to small NH3 plants (< 500 TPD) there is high uncertainty as few plants are built in this size range today (the data from the small plants in

Figure 17 is vendor-estimated values and not values from actual constructed plants). For larger NH_3 plants, there are many references, hence the uncertainty is somewhat lower.

In addition, CAPEX will depend a lot on location and local conditions.

Datasheet

The datasheet has been produced for a 229 TPD green ammonia plant, which is equivalent of a plant using electrolysis units of a total of ~90 MW.

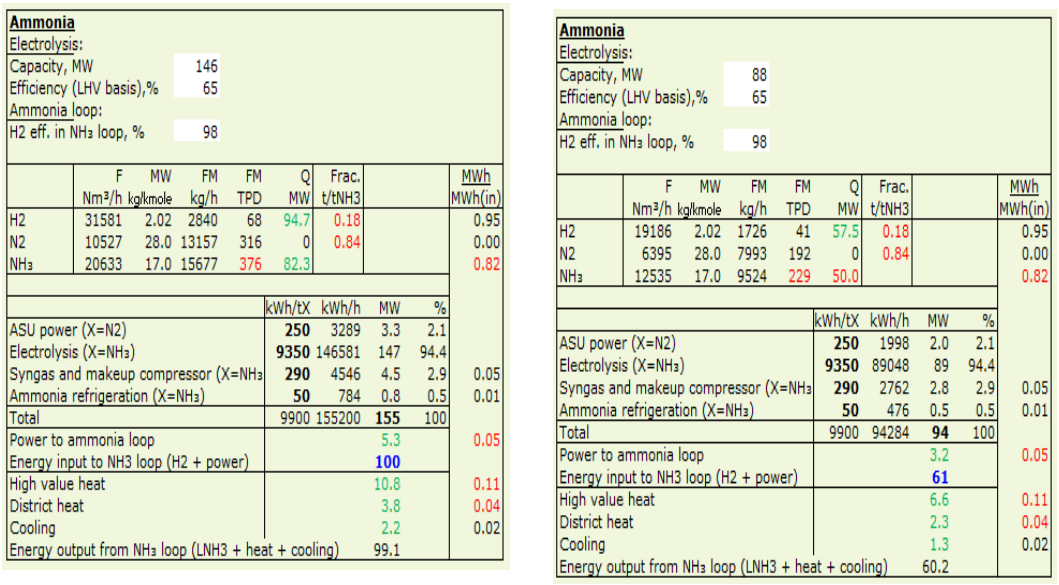
Figures for capacities increasing up to 2290 TPD (~900 MW electrolysis unit) are included to give a reference of potential future gigaplants.

In

Figure 18 below, a mass and energy balance example is given including electrolysis unit and ASU. The example has been adjusted so it matches the number in

Figure 13. Calculating back, 146 MW electrolysis is needed to give 94.7 MW H_2 that is fed to the ammonia loop producing 82.3 MW liquefied NH_3. The power needed for the ASU is 3.3 MW and the power needed for the compressors in the ammonia loop is 5.3 MW giving a total power consumption of 155 MW.

Figure 18: Calculation example including electrolysis unit and ASU.



Ammonia
Electrolysis:

Capacity, MW	146
Efficiency (LHV basis),%	65
Ammonia loop:	
H2 eff. in NH3 loop, %	98

	F Nm³/h	MW kg/kmole	FM kg/h	FM TPD	Q MW	Frac. t/tNH3		MWh MWh(in)
H2	31581	2.02	2840	68	94.7	0.18		0.95
N2	10527	28.0	13157	316	0	0.84		0.00
NH3	20633	17.0	15677	376	82.3			0.82

	kWh/tX	kWh/h	MW	%	
ASU power (X=N2)	**250**	3289	3.3	2.1	
Electrolysis (X=NH3)	**9350**	146581	147	94.4	
Syngas and makeup compressor (X=NH3	**290**	4546	4.5	2.9	0.05
Ammonia refrigeration (X=NH3)	**50**	784	0.8	0.5	0.01
Total	9900	155200	**155**	100	
Power to ammonia loop			5.3		0.05
Energy input to NH3 loop (H2 + power)			**100**		
High value heat			10.8		0.11
District heat			3.8		0.04
Cooling			2.2		0.02
Energy output from NH3 loop (LNH3 + heat + cooling)			99.1		

Ammonia
Electrolysis:

Capacity, MW	88
Efficiency (LHV basis),%	65
Ammonia loop:	
H2 eff. in NH3 loop, %	98

	F Nm³/h	MW kg/kmole	FM kg/h	FM TPD	Q MW	Frac. t/tNH3		MWh MWh(in)
H2	19186	2.02	1726	41	57.5	0.18		0.95
N2	6395	28.0	7993	192	0	0.84		0.00
NH3	12535	17.0	9524	229	50.0			0.82

	kWh/tX	kWh/h	MW	%	
ASU power (X=N2)	**250**	1998	2.0	2.1	
Electrolysis (X=NH3)	**9350**	89048	89	94.4	
Syngas and makeup compressor (X=NH3	**290**	2762	2.8	2.9	0.05
Ammonia refrigeration (X=NH3)	**50**	476	0.5	0.5	0.01
Total	9900	94284	**94**	100	
Power to ammonia loop			3.2		0.05
Energy input to NH3 loop (H2 + power)			**61**		
High value heat			6.6		0.11
District heat			2.3		0.04
Cooling			1.3		0.02
Energy output from NH3 loop (LNH3 + heat + cooling)			60.2		

Ann: Example to the left is adjusted such that the energy input to the ammonia synthesis sums up to 100 MW (blue value) producing the values given in Figure 13 (green values). The figure to the right is adjusted so it produces 50 MW NH_3, producing the values for 2020 in the datasheets (red values).

Note to use cost data in the datasheet

The cost development exclusively reflects effects from economy of scale and no further technological development is expected (see

Figure 17). Cost estimation of future gigaplants will also apply for earlier years. In case a gigaplant of e.g. 2290 TPD is expected for 2040 already instead of 2050, one should use the expected cost values for this plant size of 0.8 M€/MW, instead of the cost data for the given year.

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99 Jet Fuel from Biomass Gasification

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Publication date

August 2018

Amendments after publication date

Date	Ref.	Description
02/2019	-	The two chapters on bio fuels from gasification + Fischer Tropsch have been merged to the present chapter
12/2018		Datasheet revised

Qualitative Description

The production of diesel or jet fuel from biomass is a two-step process, in the first step the solid biomass is converted to the gas phase and in the second step the gas is converted to liquid fuels.

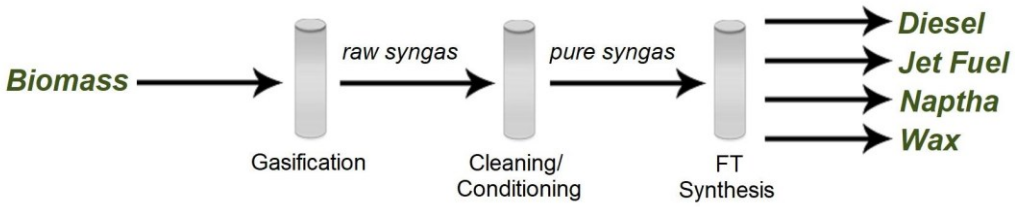
Gasification is a process that converts organic or fossil-based carbonaceous materials at high temperatures (>700°C), without combustion, with a controlled amount of oxygen and/or steam into carbon monoxide, hydrogen, and carbon dioxide (syngas). The carbon monoxide then reacts with water to form carbon dioxide and more hydrogen via a water-gas shift reaction.

The Fischer–Tropsch process (or Fischer–Tropsch Synthesis or F-T) is a set of chemical reactions that changes a mixture of carbon monoxide gas and hydrogen gas into liquid hydrocarbons. These reactions occur in the presence of certain metal catalysts, typically at temperatures of 150–300°C and pressures of one to several tens of atmospheres.

Brief Technology Description

The biomass could be agricultural or forestry residues. There is a wide range in the design of gasifiers used for biomass. Different technological solutions can be implemented in order to obtain different plant configurations; in particular, the mode of contact of the biomass with the gasification agent may be in counter-current, or co-current, or crossflow, and the heat can be transferred from the outside or directly in the reactor using a combustion agent; the residence time can be in the order of hours (static gasifiers, rotary kiln) or minutes (fluidized bed gasifiers). Different gasifier designs are better suited to different feedstocks and gas needs. Gasification is further described in chapter 83 Gasifier Plants (for Producer Gas) and 94 Syngas Methanation (from Thermal Gasification). The Fisher Tropsch reactions are practised commercially on syngas produced from coal (Sasol) and on natural gas (Shell, Chevron, Sasol, and others). The overall process is shown in the following simplified process flow diagram.

Figure 1: Biomass to Diesel and Jet Process



Input

The primary input for most process is just the biomass. The reactions are exothermic and generate enough heat for the process and to produce the power required for the system.

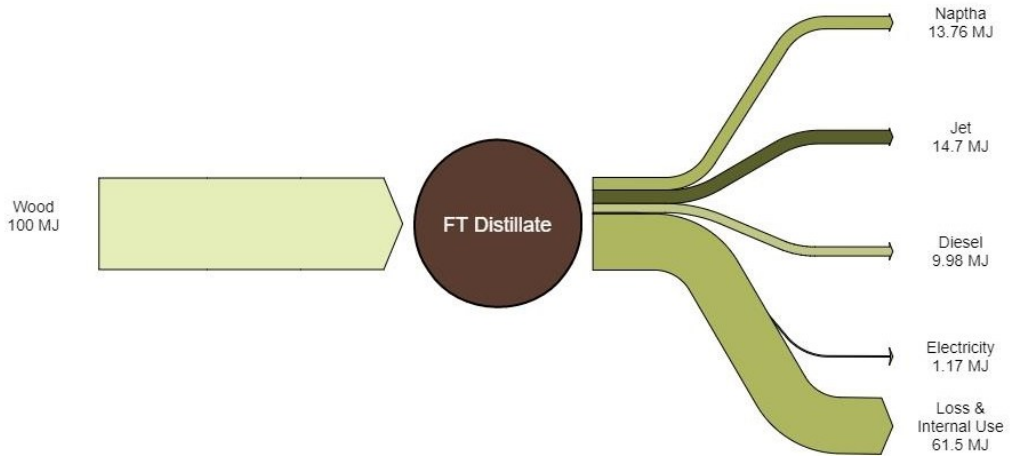
Output

The FT synthesis process produces a range of hydrocarbon from light ends to heavy waxes. It is difficult to control the selectivity of the process to produce just diesel fuel or jet fuel. In some commercial facilities the light ends and the heavy wax materials can be recycled through the process to improve the selectivity but usually at the expense of overall efficiency. Some systems will produce excess power for sale from the system.

Energy Balance

The energy balance for a system is shown in the following figure [1]. It is based on NREL's work on biomass to liquids. The carbon efficiency of the biomass gasifier to raw syngas is 72% and the carbon efficiency of the syngas to fuels is 46% for an overall carbon efficiency of biomass to fuels of 33.2%. The energy out per unit of energy in is 39%.

Figure 2: Biomass to Diesel and Jet Energy Balance



The overall energy efficiency of the process is relatively low. There are two potential means to recover some of the waste heat. The plants use some of the process heat to produce electricity for the plant use and potentially a small amount to be exported. Steam from the exit of the final steam turbine would be available for other uses. This could have a temperature between 150 °C and 185 °C depending on the design. There may also be some opportunity to recover some lower grade heat as the syngas is conditioned prior to synthesis. Details of the potential for energy recovery are not reported in most of the recent techno-economic studies published.

Other computer simulations of biomass to FT systems have reported higher efficiencies. Kreutz et al [2] reported 49 to 50.5% energy efficiency on a LHV basis. They had similar carbon efficiency of the feedstock to the fuels.

Baliban et al [3] modelled several optimized hardwood to FT liquid process configurations. The energy efficiency ranged from 56 to 61% and the carbon conversion efficiency ranges from 54 to 60%.

Sikarwar [4] identified feedstock characteristics that influence the performance of biomass gasification systems. These include moisture content, ash content, chlorine and sulphur, and the cellulose, hemicellulose and lignin fractions as the three components degrade at different temperatures during gasification. He reports that in general, the higher the cellulose and hemicellulose content, the greater the volume of gaseous products formed. Therefore, softwood, hardwood, wheat straw and bagasse with much higher cumulative percentages of cellulose and hemicellulose are preferred over sunflower seed hull, coconut shell, almond shell, larch plant or poultry litter, when attempting to obtain gas as the final product.

Typical Capacities

There are no commercial scale systems in operation. The NREL techno-economic analysis was based on processing 2,000 tonnes of dry wood per day. The plant would produce 175 million litres per year of fuel.

In a review of the state-of-the-art biomass gasification [5], Molino et al reports on the European biomass gasification plants. They identify 22 gasification facilities but only five that has more than 7,000 h/year of operation time. The capacities are in the 10 to 40 MW range with one larger facility that co-fires biomass and fossil fuels. The facilities are identified in the following table.

Table 1: European Biomass Gasification Plants

Site	Thermal Output, GJ/year	Running h/year
Harboøre (Denmark)	576,000	8,000
Gussing (Austria)	230,400	7,000–8,000
Skive (Denmark)	576,000	7,500
Lahti (Finland)	1,152,000	7,000
Buggenum (the Netherlands)	17,280,000	7,500

The commercial FT plants using fossil energy as the input are all much larger than the biomass gasification plants shown in the table above. The largest fossil plant is the 260,000 bbls/day (500 million GJ/year) Shell plant in Qatar. Shell's original GTL plant in Malaysia has a capacity of less than 15,000 bbls/day (30 million GJ/year).

There is work ongoing on small FT distillate reactors. Velocys claims that the commercially optimal size for their biomass to FT liquids system is 1,900 bbl/day (72 million litres/year) [6]. Their reference plant processes landfill gas and produced 200 bbls/day of finished products (375,000 GJ/year).

Regulation Ability

Biomass gasifiers can be operated down to about 35% of the rated capacity depending on the configuration, feedstock moisture contents and the acceptable efficiency loss [7]. However, the

gas composition will change over this range and when the gasifier is coupled to a synthesis reactor there is a need for relatively constant feed compositions. The high temperature and pressure of the FT reactor will limit the regulation ability of the overall system.

Space Requirements

The biomass gasification district heating plant in Harboøre is situated on a less than one hectare. The GoBiGas facility in Sweden, which is a biomass gasifier and an methanation facility is on a two-hectare site. The original design capacity was 100 MW (3 million GJ/year), although only the first phase of 20 MW was built.

The space requirements will be less than 1,000 m2/MW. This may be reduced if the size of the units is larger.

Advantages/Disadvantages

Biofuels that can be produced from non-feed or food feedstocks and can be used in heavy duty transport applications, which can't be easily electrified, are an attractive option for decarbonizing the transport sector. Drop-in biofuels, such as the FT fuels made by this technology can be used in the existing fuel infrastructure and are attractive to the existing fuel providers.

This technology combines gasification systems that have only been operated at small scale and FT synthesis systems that are commercialized at very large scale. Determining the combined size that will work, technically and economically, for both technologies is a challenge. The techno-economic analyses that have been done on this technology have considered plants in the range of 1,000 to 2,000 dry tonnes of wood per day (350,000 to 700,000 dry tonnes of wood per year. The Danish Energy Agency reports the following production of woody biomass in 2015 [8].

Table 2: Woody Biomass Production and Consumption for Energy

Type	TJ	Tonnes (dry)
Wood Chips	13,335	701,842
Firewood	21,943	1,154,895
Wood pellets	2,641	139,000
Wood waste	8,837	465,105
Total	46,756	

A single woody biomass to diesel and jet fuel plant would require a 15 to 30% increase in the current production and consumption of woody biomass in Denmark.

Environment

The sustainability of the feedstock production is a potential issue with all biomass systems. The overall energy out per unit of energy in the feedstock is relatively low for this technology. Biomass gasification systems will produce some ash that must be disposed of. The wood ash can be used to adjust the pH of soils but the availability of the nutrients in the ash may not always be fully bioavailable. Wood species and gasification type appear to have some influence on the properties [9].

The fuels produced have no sulphur, are low in aromatics and are considered clean burning. Their volumetric energy content is about 10% lower than diesel fuel due to the lower density.

Research and Development Perspective

Biomass gasification for diesel and jet fuel produced from wood or straw is a category 2 technology, a pioneer phase technology with limited applications to date. The technology has been proven to work through demonstration facilities or semi-commercial plants. However, due to the limited application, the price and performance is still attached with high uncertainty, since development and customization is still needed. The technology still has a significant development potential.

Both the gasification and the FT distillate technology have been known and practiced for almost 100 years. They are commercial technologies for other feedstocks.

There is work underway on integrating the two technologies, improving the gas clean-up system performance and addressing the issue of scale for the fuel synthesis stage.

Examples of Market Standard Technology

The technology has not yet been commercialized. In Europe, Repotec, an Austrian company, have been involved with the Gussing gasifier, the GoBiGas project in Sweden, and the Senden wood gasifier to power facility in Germany.

In Denmark, B&W Vølund built the wood gasifier at Harboøre but no other references for the technology were identified.

The UK-American company, Velocys is working on small scale FT plants. They are developing smaller scale microchannel FT technology that was originally developed by the Pacific Northwest National Laboratory in Washington State, USA. Their first project is using landfill gas but they are working with ThermoChem Recovery International of gasification systems for woody biomass that would be coupled with the Velocys FT technology [10]. The system would produce 1,400 bbl/day of FT products. This would require 1,000 tonnes of wood per day.

Velocys is working on a USDA Phase 2 application for a loan guarantee for a wood to FT liquid project [11]. Phase 2 applications include the environmental report, technical report, financial model, and the lender's credit evaluation. The plant is to be built in Natchez, Mississippi on a 40-ha site. The plan is to start construction in late 2018 or early 2019. Velocys are also working on a waste to jet fuels project in the UK with British Airways and other partners but this project is not as well defined as the wood project in the USA.

Velocys plc

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Gasifier design: Entrained flow reactor

The entrained flow gasifier technology is well suited for large scale gasification. This specific design is described in this section.

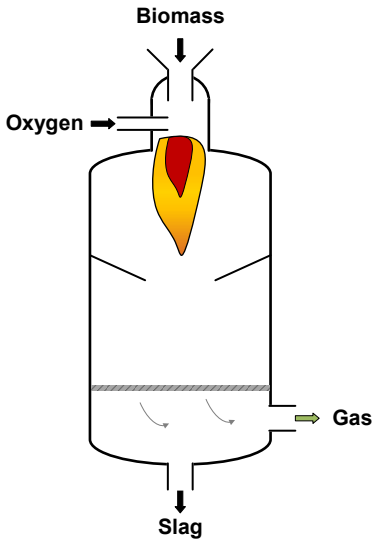
In an entrained flow reactor, the high temperatures and pressure produces a clean syngas with very little methane and tar [18]. This makes the gas well suited for further chemical processing and production of bio-fuels.

The entrained flow reactor has been used for large scale coal gasification for decades but is much less developed and demonstrated for biomass. The fine ground feedstock material which, when coal is used, may be mixed with water to a slurry, is fed from the reactor top together with steam and oxygen. When biomass is used, the input may be pyrolysis oil. Partial combustion and gasification take place in the pressurised reactor at high temperatures (>1,000°C, up to 1,600°C

for coal). Slag and ash are removed from the bottom. A high capacity is possible due to high reactivity at high temperature and pressure. The high combustion temperature results in formation of slag instead of ash as the main residue.

The high outlet gas temperature usually makes the thermal efficiency low, unless the process is integrated with other energy consuming processes. possibility is to pre-treat solid fuel by torrefaction at 300°C, whereby the fuel is easier to pulverise, and the overall efficiency is improved [19]. However, an additional energy loss in the torrefaction process be expected. As for the CFB gasifier processes, the oxygen necessary may be produced in a process-integrated air separation unit powered by electricity internally produced from excess process heat.

Figure 3: Principle of the entrained flow gasifier.



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Challenges when moving from coal to solid biomass feedstock comprise obtaining a uniform particle size distribution and feeding biomass into a highly pressurised vessel. Instead of pulverising the fuel, it been suggested as a pre-treatment to transform it oil/char slurry through a fast pyrolysis.

Also, ignition and flame stability as well as the alkali content and ash melting behaviour in biomass are challenging issues.

Advantages/disadvantages

Compared to CFB gasifiers, the entrained flow gasifiers can have considerably higher throughput, which together with the high temperatures favours upscaling. Further, the high temperatures produce a clean syngas with no tar and very little methane. The high outlet gas temperatures usually make the thermal efficiency lower, unless integrated with other energy consuming processes.

Entrained flow gasification has an advantage if the available fuel is a liquid that can be spray atomised, like for instance the residues of paper pulp manufacturing. However, for solid fuels like fresh wood, the grinding will use considerable amounts of energy. Pre-treatment by torrefaction or pyrolysis may reduce these costs and reduce feeding challenges into a pressurized system.

The output gas has a lower content of methane than will be possible with CFB gasifiers. This will reduce the fuel-to-methane efficiency. Thus, the entrained flow gasifiers appear to be more suitable for processes where the end-product is not methane, e.g. other synthetic fuels such as Fischer Tropsch diesel, or for direct combustion in gas turbines.

A disadvantage compared with CFB gasifiers is, that the combustion at high temperatures will result in a slag residue which cannot be recycled to the environment. This is particularly relevant for fuels with a high content of ash and nutrients such as straw.

The large plant sizes would require an efficient feed stock sourcing strategy and possibly increase the need for pre-treatment by torrefaction and/or pelletisation or pyrolysis - possibly decentralized - to optimise sourcing.

Predication of Performance and Cost

The prediction of performance and cost is based on published techno-economic papers rather than on actual plant performance. The NREL paper is based on a plant twice the size of the proposed Velocys project.

Uncertainty

There is a high level of uncertainty for the technology given the state of development and the fact that there are no operating plants in the world at this time.

Additional Remarks

One of the challenges for small scale FT plants has been that a range of products is produced, from gasoline boiling range products to waxes. Markets for all products are required for commercial success and finding markets for small volumes of gasoline blending components and the wax can be an issue. In some projects the revenue from the wax has been a significant portion of the total revenue.

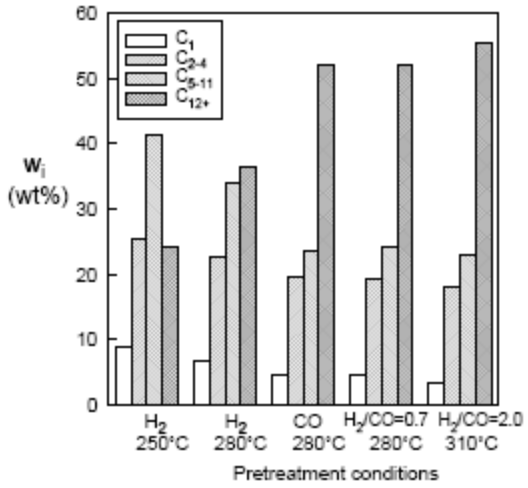
FT synthesis produces a range of products between C1 and waxes. The actual ranges will vary with process type, catalysts, and syngas quality but there is always a range of products. This implies that separation of the relevant fractions will be needed downstream the FT process. Tijm [12] reports on the product distribution for two different process severities as shown in the following table. Unfortunately, the paper does not provide the accompanying yield data for the two operating conditions but there is more gasoline produced in the kerosene mode than the diesel mode.

Figure 4: Product Distributions – Shell SMDS

	Gas Oil Mode	Kerosene Mode
	% wt	
Tops/naphtha	15	25
Kerosene	25	50
Gas Oil	50	25

In his 1999 thesis, van der Lann [13] showed that the quantity of each group of products did vary with operation conditions. This is shown in the following figure where the two right hand bars represent the liquid products and the two left hand bars represent the gaseous products. The sum of the two liquid products (and thus the yield) as well as the ratio of heavy to light liquid products does vary with the pretreatment conditions.

Figure 5: Selectivity vs. Yield



Quantitative Description

The available quantitative data that is available on the technology is mostly from third parties and not from the technology providers or plant operators. Actual plant data is considered confidential by the process developers.

There are three basic reactions that occur in the process. The first reaction breaks the biomass down to a combination of hydrogen, carbon monoxide and carbon dioxide. A simplified reaction is shown below. Actual biomass has highly variable composition and complexity with cellulose as one major component.

$$C_6H_{12}O_6 + O_2 + H_2O \rightarrow CO + CO_2 + H_2 + CH_4 + \text{other species}$$

Note: The above reaction uses glucose as a surrogate for cellulose.

Stoichiometry for methanol production of syngas requires the ratio of H_2/CO to equal 2. The product gases are then subjected to the water-gas shift reaction to increase the quantity of hydrogen. The equilibrium for this reaction is temperature dependent which controls the CO to CO_2 ratio.

$$CO + H_2O \leftrightarrow CO_2 + H_2$$

This is then followed by the synthesis reaction as shown below. In this reaction the carbons are added sequentially making it difficult to control the chain lengths of the final products.

$$(2n + 1)\ H_2 + n\ CO \rightarrow C_nH_{2n}{+}2 + n\ H_2O$$

Generally, the Fischer–Tropsch process is operated in the temperature range of 150–300°C. Higher temperatures lead to faster reactions and higher conversion rates but also tend to favor methane production. For this reason, the temperature is usually maintained at the low to middle part of the range. Increasing the pressure leads to higher conversion rates and also favors formation of long-chained alkanes, both of which are desirable. Typical pressures are up to 30 bar.

Typical Plant Size

The proposed Velocys plant will process 1,000 tpd of wood and produce 79 million litres of product. The NREL techno-economic work assumed a plant size of 2,000 tpd. This is about the wood required for an average pulp mill. New pulp mills are being built larger and can consume up to 10,000 tpd. Plant size will likely be determined by the feedstock availability.

Input and Output

The primary input and output for a wood to FT plant is summarized in the following table [1]. There are some chemicals and catalysts required but the quantities are very small.

Table 3: Inputs and Outputs

Parameter	Input	Output
Wood	4.06 kg	
Wood	76 MJ	
Naphtha		0.36 litre
Jet		0.38 litre

Diesel		0.26 litre
Power		0.26 kWh

Forced and Planned Outage
The plants are expected to operate for 350 days per year. Wood gasifiers are capable of operating at these rates as shown earlier and fossil FT plants are also capable of operating at these rates. Forced outages are expected to be minimal.

Technical Lifetime
Due to the maturity of the technology, plant lifetime is estimated to be 20 years for plants build before 2025. Hereafter, it is expected to grow to 25 years.

Construction Time
Construction time for the technology is expected to be about 24 months.

Financial Data

The financial data is only available from the literature. There is a preliminary cost estimate of $300 million for the 72 million litre Velocys plant in Mississippi (80 MW) [14]. This would be €4.25/litre for a European plant.

Investment Costs
Tan et al project that the total capital investment for the plant is $650 million for the 180 million litre plant. The cost basis is 2011. Converting this to 2015 Euros the cost would be 716 million Euro, or €4.0/litre of product.

Irena [15] report current capital costs for this pathway as $3,000 to $5,000/kW (€2.8 to €4.6/litre). The EU Sub Group on Advanced Biofuels [16] report the capital costs as €3,000/kW (€3.35/litre).

Considering the capital growth factor information from de Jong [17] the capital cost information for the n^{th} plant from Tan and the Velocys pioneering plant is too close together. It is likely that the pioneering cost estimate is too low, we have increased it to €5.00/litre, which may still be too optimistic.

Operating and Maintenance Costs
The Tan et al estimates of fixed and variable operating costs, excluding feedstock are shown in the following table. These costs are much less than the feedstock costs and the capital related costs in the analysis.

Table 4: Operating and Maintenance Costs

Parameter	Euro/litre	Parameter
Variable operating costs, ex feedstock	0.01	Variable operating costs, excluding feedstock
Fixed costs	0.12	Fixed costs

Start-up Costs
The start-up costs are included in the costs shown above.

Technology Specific Data

The typical properties of FT diesel are compared to petroleum diesel in the following table. The fuel has a higher cetane than petroleum diesel but a lower volumetric energy content. Due to the low content of aromatics the gasoline produced from FT naphtha will have a low octane number.

Table 5: Typical Fuel Properties

	Petroleum Diesel	FT Diesel
Density, kg/litre	0.84	0.77
Energy content, MJ/litre	36.0	33.9
Energy content, MJ/kg	42.8	44.0
Cetane	48	~70

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100 Jet Fuel from Hydrogen and Carbon Dioxide

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Publication date

May 2020

Amendments after publication date

Date	Ref.	Description
07 2020		Minor adjustment in DH output for hydrogen to jet, note letters fixed and note added

Qualitative Description

The pathway described in this chapter utilizes electricity to produce hydrogen and the hydrogen is reacted with carbon dioxide (CO_2) to produce syngas (hydrogen (H_2) and carbon monoxide (CO)), which is then used in a Fischer-Tropsch (FT) process to produce syncrude or Fischer-Tropsch wax, which can subsequently be hydrocracked and upgraded through refinery processes to diesel, kerosene, jet fuel, other hydrocarbons, and heat.

The system can take several forms. There are several different electrolysis technologies, the carbon dioxide could come from many different sources, and there are several different technologies being developed for the conversion of carbon dioxide to carbon monoxide, which along with hydrogen is the reactant for the Fischer-Tropsch synthesis. There is also some research underway on the direct utilization of carbon dioxide rather than first producing carbon monoxide. There are other production methods for the production of emission free hydrogen, for example methane pyrolysis [□]; these are not described in this chapter.

This technology for direct conversion of power to jet fuel is at the early stage of development with only a few pilot plants that are operable, while conversion of syngas to Fischer-Tropsch wax and the subsequent upgrade has been in large-scale industrial operation for decades in South Africa, Malaysia and the Middle East. The plants have been developed by technology aggregators, who might have developed one of the subsystems but rely on other technology providers for the balance of the plant.

There are other power to jet routes that could be considered as well. The power and CO_2 could first be used to produce methanol, this can be accomplished without the conversion of the CO_2 to CO and was described in the chapter

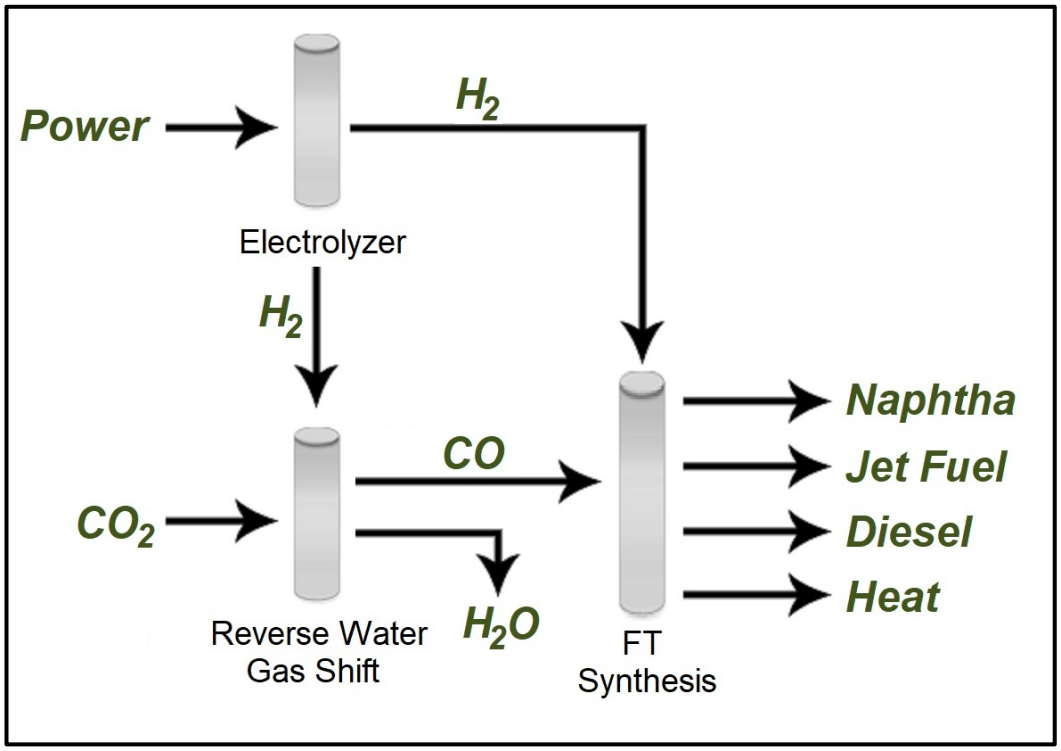
. The methanol could be converted to olefins and then to light ends, gasoline, and jet fuel fractions [□]. The power to methanol production portion of the system has a higher TRL than the FT pathway but the subsequent conversion of methanol to hydrocarbons involves more process steps and has a lower TRL. This route is being explored by a Finish group of companies [□].

Brief Technology Description

Electricity is used to make hydrogen via electrolysis and carbon dioxide is reduced to carbon monoxide and water. The two streams are combined to produce a syngas, which is

then synthesized through the Fischer-Tropsch reactions to produce liquid hydrocarbons and heat. The basic process flow is shown in Figure 1.

Figure 1: Power to Jet Process Flow



The water from the reverse water gas shift (RWGS) can be recycled back to the electrolyzer for hydrogen production.

There are a number of catalysts that can be used for the Fischer-Tropsch synthesis but iron and cobalt based catalyst are the most common. The iron catalysts typically operate in a temperature range of 300 to 350°C and the cobalt catalysts operate at lower temperatures (200 to 240°C) and both operate at pressures of 20 to 25 bar [].

Input

The process inputs are electricity (for hydrogen production) and carbon dioxide. Data sheets at the end of the chapter are provided for both electricity as the input and for hydrogen as an input. Renewable sources of electricity are preferred in order to reduce the greenhouse gas (GHG) emissions associated with the produced products. Some analyses include about 3% external fuel.

Cobalt catalysts required in situ regeneration every 9 to 12 months and replacement every five years []. Cobalt catalyst consumption rate of 0.0009 kg per kg of FT liquids produced was modelled in a recent lifecycle analysis of an FT system []. Iron catalysts have limited lifetimes of 40 to 100 days but are 1/1000th the cost of Cobalt catalysts. Large scale natural gas to FT plants employs multiple parallel reactors that can facilitate catalyst changes.

The carbon dioxide can be from concentrated sources such as ethanol fermentation facilities and ammonia plants, through medium concentration sources such as thermal power plants, and even to dilute sources such as direct air capture facilities. The energy requirements for the concept will increase as the concentration of the CO_2 sources decrease.

The FT synthesis actually needs carbon monoxide, not carbon dioxide, as one of the reactants. The traditional process to convert carbon dioxide to carbon monoxide is through the use of the reverse water gas (RWGS) shift reaction shown below. The process is described in more detail by Daza et al [ ]. The reaction is endothermic (requires heat).

$$CO_2 + H_2 \rightleftharpoons CO + H2O$$

The reaction is undertaken at temperatures between 350 to 600°C, depending on the catalysts used and at relatively low pressures. The reaction is reversible so that there will always be some CO_2 in the gaseous stream leaving the reactor. Konig et al [ ] designed for 80% conversion to CO in the RWGS and recycled the excess CO_2 from the FT synthesis back to the RWGS reactor.

It is also possible to use an electrochemical process to convert CO_2 to CO [ ]. There is development work ongoing in this area, and Haldor Topsoe offers a small-scale commercial product (eCOs). The reaction is shown below.

$$CO_2 + 2\ H+ + 2\ e- \rightarrow CO + H_2O$$

There is also work going on with thermochemical and photochemical CO_2 reduction processes but these are not yet commercially available [ ] [ ].

The FT synthesis process needs a H_2:CO ratio of about 2, for example C13H28 (tridecane, a typical component of jet fuel) the ratio is 28:13 = 2.15 [ ]. Considering the overall ratio, including the hydrogen required for the reverse water gas shift, the ratio for H:CO_2 required for the FT synthesis becomes about 3:1. There is a range of assumed carbon efficiencies in the literature. It is likely that early plants will have a higher ratio of hydrogen to CO and then improve over time.

Output

The Fischer-Tropsch synthesis is essentially a polymerization reaction in which carbon bonds are formed from carbon atoms derived from carbon monoxide, under the influence of hydrogen in the presence of a metal catalyst. The reaction leads to a range of products which depend on the reaction conditions and catalysts employed. The most abundant compound classes are paraffins, olefins, and alcohols (oxygenates) as shown below [ ]. The alcohols can be removed in the post reaction processing or used for energy to drive the process.

$$nCO + (2n+1)H_2 \rightarrow C_nH_2n+2 + nH_2O$$

$$nCO + 2nH_2 \rightarrow C_nH_{2n} + nH2O$$

$$nCO + 2nH2 \rightarrow C_nH_{2n}+1OH + (n-)H_2O$$

The FT reactions are not particularly selective and they typically make a range of alcohols, olefins and paraffinic hydrocarbons that range from light naphtha that could be used for gasoline production, through to jet fuel, diesel fuel and traditionally heavy waxes, which can be further processed into high quality lubricants. There can be trade-offs between liquid product yield and product selectivity. DeKlerk [ ] reported the typical product range for different catalysts and operating conditions. The results are shown in the following table; jet fuel is composed of the heavy end of the naphtha and the light end of the diesel fuel (C8 to C16).

Table 1: FT Synthesis Product Distribution

	Low temp Iron	Low temp Cobalt	High temp Iron
	Wt%	Wt%	Wt%
C_1 to C_2 gas	6	7	23
$C_2 - C_4$	8	5	24
Oxygenates	4	2	10
Naphtha (C_5 to C_{11})	12	20	33
Diesel (C_{12}-C_{20})	20	22	7
Wax (C18 –C100)	50	44	-
Total	100	100	97

DeKlerk reports that the production of on-specification jet fuel is the easiest of the transport fuel types to accomplish. It is necessary that the paraffins in the kerosene range (C8 to C16) are sufficiently branched to meet the cold flow specification, which requires a freezing point below minus 47°C. It is also necessary to have (8-25) % aromatics in the kerosene range. On specification, full synthetic jet fuel as well as semi-synthetic (50% blend with crude oil derived kerosene) are produced on industrial scale from Fe-HTFT derived synthetic oil using fossil feedstocks.

A US Patent was issued to De Klerk [] in 2014 for a process, which has FT jet fuel yield of 60%. The patent is assigned to Sasol in South Africa. The process includes at least four of the following five processes.

1. Hydrocracking the kerosene or heavier fraction and a C9 or heavier FT Syncrude fraction.
2. Olefin oliogmerizing produces kerosene range material from lighter olefinic materials.
3. Hydrotreating one or more of an FT syncrude fraction, a product from process b, and an alkylated FT Syncrude fraction.
4. Aromatizing one or more of an FT syncrude fraction including hydrocarbons in the range C2 to C8, a product from process a, a product from process b, a product from process c, and a product from an aromatic alkylation process.
5. Alkylating one or more of an FT Syncrude fractions including hydrocarbons in the C2 to C6 range, a product from process b, and a product from process d.

Such a process would add significantly to the capital cost of a facility. Many of the processes are typically found at a petroleum refinery so co-locating the Power to FT facility at a petroleum refinery may be an option to take advantage of the economies of scale typically found in petroleum refineries.

The FT reactions are also exothermic so some excess heat can be captured by the process. The total amount of excess heat is about 25% of the heat of combustion of the synthesis gas, which means that both significant energy is available for export, (up to 25%

of the syngas input) and that there is a theoretical limit for the efficiency of the process (75%) [?]. Konig [?] reports that low (125°C), medium (175°C) and high (>200°C) pressure steam is available from the FT process. Seventy percent of the steam is high pressure, seventeen percent is low pressure, with thirteen percent being medium pressure.

Energy Balance

The overall system energy balance will depend on the choices of technology that are employed.

For the hydrogen production there are three choices; Alkaline electrolyzers, which are the most common systems used today, polymer electrolyte membrane (PEM) electrolyzers, and solid oxide electrolyzer cells (SOEC), which use a solid ceramic material as the electrolyte that selectively conducts negatively charged oxygen ions (O2-) at elevated temperatures and generate hydrogen in a slightly different way.

Alkaline electrolyzers have the largest market share, they have no expensive noble metals (platinum, palladium, ruthenium, etc.) like PEM systems, but they have a lower efficiency than PEM or SOEC systems.

PEM systems typically have slightly higher efficiencies than alkaline systems, operate at a slightly lower temperature, but they use noble metals as the catalyst which can result in higher capital costs.

Solid oxide electrolyzers must operate at temperatures high enough for the solid oxide membranes to function properly (about 700°–800 °C, compared to PEM electrolyzers, which operate at 70°–90°C, and commercial alkaline electrolyzers, which operate at 60°–80°C). The solid oxide electrolyzers can effectively use heat available at these elevated temperatures (from various sources, including from the FT process) to decrease the amount of electrical energy needed to produce hydrogen from water. They are not yet used commercially. Further information is available in chapter 80 Electrolysis.

Lehner et al [?] compared the characteristics of the systems and provided some advantages and disadvantages. This 2014 book has been cited over 200 times in the literature. It is possible that further technological advances have occurred with the technologies since publication, particularly with the developing technologies. Power requirements for electrolysis systems are not always quoted on a consistent basis. Not all reports will include ancillary loads for pumps and fans and some only quote the stack power requirements without including the losses from the conversion of AC to DC power. The size in the table refers to a module size; multiple modules could be used at a single site for increased capacity. These are summarized in

Table 2.

Table 2: Electrolyzer System Comparison [14]

	Alkaline	PEM	SOEC
Maturity	Mature	Commercial	Demonstration
Efficiency, kWh/m3H2	4.5-6.6	4.2-6.6	>3.7
Mid-point Efficiency, % LHV	50%	52%	75%
Size, m3H2/hr	<760	<40	<40

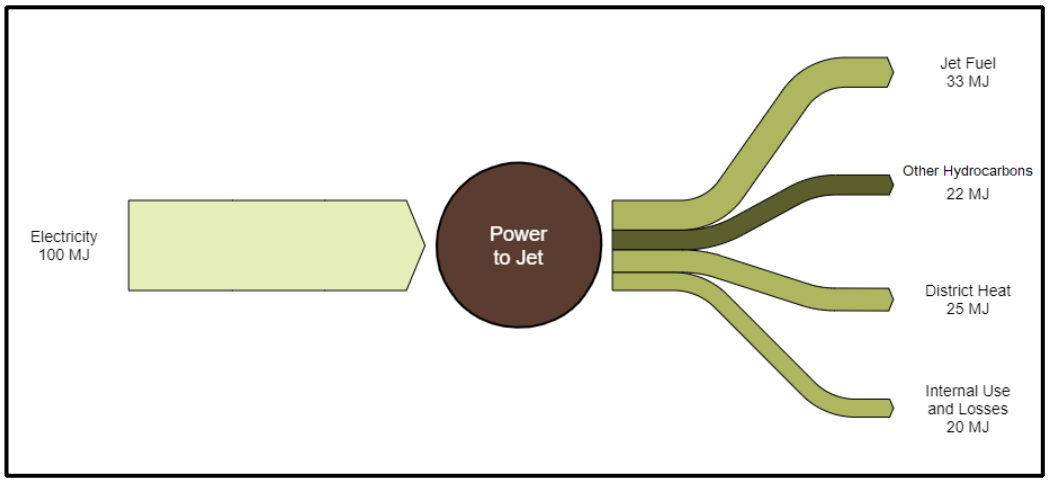
Gas purity, %	>99.5	99.99	99.9
Capital costs, €/kw	1000-1200	1860-2320	>2000

There are only two operating power to FT synthesis pilot plants [] [] and neither have publicly released any performance data and production rates are on the order of 100's of litres per day. Sunfire first produced FT distillates at their research facilities in Dresden Germany in 2015. They used CO_2 from direct air capture and a solid oxide electrolyzer to produce the hydrogen. They claimed up to 70% efficiency for the power to liquids technology but no detail of that calculation is publicly available. Carbon Engineering [] also use direct air capture for the CO_2 but use an alkaline electrolyzer for hydrogen. They have also not provided any technical performance data. The energy balance must therefore be estimated from information in the literature.

There have been a number of papers and publications that have performed prospective techno-economic assessments of power to jet systems. The most detail was in the 2019 report "Nordic GTL" by Mortensen et al []. That study also reported the highest liquid fuel efficiency at 65% as a SOEC system was assumed. A 2016 report for the German Government by Schmidt et al [] reported fuel efficiencies from 53 to 64% depending on the electrolysis system used. The World Energy Council [] reported fuel efficiency of 45%, which they considered a realistic process efficiency considering an economic perspective. Konig et al [] undertook process modelling and reported a 43.3% power to liquid efficiency and a 73.7% carbon efficiency, which is lower than reported in some other studies. Light fuel products that are generated by the FT synthesis are separated and used for fuel to drive the process, thus there are carbon losses from the system. See the mass balance detail figure in the Quantitative sections. This study included the product separation and upgrading to gasoline, diesel, and jet fuel.

The technology that is closest to being commercially available at a reasonable scale is the use of an alkaline electrolysis system and the reverse water gas shift reaction. The estimated energy balance for this system is shown below []. Carbon dioxide does not contain any energy but depending on the source of the CO_2, some energy may have been expended to capture and concentrate the CO_2. The distribution between losses and steam for district heat is our estimate. This reference did not include any external fuel.

Figure 2: Power to Jet Fuel Energy Balance 2050 Projection



This energy balance has a higher jet fuel fraction than DeKlerk [] and others have suggested. The other hydrocarbons will include gaseous hydrocarbons and liquids that are both lighter and heavier than jet fuel. Different FT systems could have different distributions of energy available for district heat versus own use.

Using the mid-point of the efficiency ranges reported by Lehner [] which is shown in Table 3 which are representative of commercial systems, and the high end of the FT synthesis reported by [] the following Table 3 can be developed. While the reverse water gas shift reaction is endothermic it has been assumed that this energy can be obtained from the exothermic FT synthesis.

Table 3: Potential System Efficiencies

	Alkaline	PEM	SOEC
kWh/m^3H_2 [271]	5.5	5.3	4
MJ electricity/m^3 H_2	19.8	19.08	14.4
MJ H_2 (LHV)/m^3	9.84	9.84	9.84
H_2 Efficiency	50%	52%	68%
FT Synthesis Efficiency []	75%	75%	75%
Overall Efficiency	37%	39%	51%

From the table it would appear that some of the expected energy efficiencies in the publicly available feasibility studies [] [] [] may be difficult to achieve in the first demonstration and commercial facilities.

Typical Capacities

The existing pilot plants have the capacity to produce 160 litres of fuel per day (~1 barrel). Commercial plants will be much larger.

The Schmidt study [] looked at a plant that produced 123,000 tonnes per year of fuel (~150 million litres/year). The Nordic GTL study [] was based on a plant that produced 190,000 tonnes of liquid fuels. Plants of this size are similar to commercial biodiesel or grain ethanol plants.

A plant that produces 150,000 tonnes of fuel will consume about 580,000 tonnes of CO2 at a 95% carbon efficiency. It will need 400 MW of power.

Carbon Engineering has recently reported [] that the first plants are likely to produce about 2,000 bbls/day of fuel and will utilize 1,000 tpd of CO_2. This is about 85,000 tonnes per year of fuel from 350,000 tonnes of CO_2 (this is about 75% carbon efficiency).

The first plants will likely be smaller than the plants studied in the literature.

Regulation Ability

There is little published on the performance of continuously operated plants. Given the high pressure and temperatures required in the reactors and the required reactor residence time it is likely that the performance will be altered when the process is operated at rates below the design capacity. Goldmann et el [] reported that the FT process (including the RWGS)

has a low tolerance for variations in the supply of reactants and that hydrogen would need to be stored in excess to be available at a constant rate. The regulation ability will therefore have a linkage to the capital cost of the system. Overtoom [] reported that the Shell FT plant in Malaysia requires two to three days to start the complex and to bring it to full production. During start-up the process is consuming energy without producing products and frequent start-up and shut down can have a significant negative impact on overall system efficiency and economic performance.

Space Requirements

The Schmidt report [] calculated that the land requirements were 0.001 to 0.002 ha/GJ Jet fuel/year, however this included the land requirements for the wind power which is outside the scope of this chapter. A 4,000 bbl/day plant would require 7,000 to 14,000 ha according to this metric. The authors report that some of the land could still be used for agriculture.

Without any commercial scale plants in operation, it is difficult to estimate the space requirements. The Pearl GTL plant in Qatar has a 140,000 bbl/day capacity and occupies 550 ha. It has 29 FT reactors, so each reactor would produce 4,800 bbl/day, which is 20% larger than the scale of the Nordic plant that was analyzed []. This suggests that the Nordic plant might require 20 ha. It is possible that the space requirements might be larger than this as the Qatar plant is part of a large petrochemical process and it might share some facilities with other operations and the power to jet fuel plant will have electrolyzers and probably some hydrogen storage.

The space requirements do not include any required space for CO_2 capture if that is part of the processing system.

Advantages/Disadvantages

The primary attractiveness of the technology is that the liquid fuel can have a very low GHG emission profile if the power is from a low carbon source such as wind, solar, or hydro.

The challenge with this technology will be that the availability of the low carbon electricity will likely be intermittent and since the fuel synthesis portion of the process likes to operate at a constant rate there will be a requirement for hydrogen storage to allow for operation when the power isn't available to produce the hydrogen. Hydrogen storage will increase the capital costs. Carbon dioxide storage might also be required depending on the stability of the supply source.

The FT process also does not have a high selectivity for jet fuel (or any of the liquid product components). In addition to producing jet fuel there will be significant quantities of gaseous fuels, gasoline, diesel fuel, and perhaps heavy waxes produced, depending on the catalysts used and the operating conditions. The heavy wax can be used as a feedstock to produce high quality lubricating oil base stocks. These products will also have low carbon intensity.

Environment

The GHG emission performance of the product will depend on the carbon intensity of the power use to produce the fuel. The carbon dioxide used to produce the fuel will be released when the fuel is combusted. One cannot generate carbon credits for the CO_2 capture and consider the combustion of the fuel from the process to be carbon neutral. In a lifecycle assessment only one of these options can be considered. That is, if CO_2 credits are issued for the CO_2 capture then the fuel combustion emissions are the same as a fossil fuel. It is important that if the CO_2 from the combustion is not counted in the system GHG emissions that there is no credit given for the use of the CO_2 going into the process.

The production of hydrogen by electrolysis also requires water. Schmidt [] calculated a theoretical net process water demand of 0.8 litres per litre of jet fuel but estimated that in reality the demand would be 1.3 to 1.4 litres per litre of jet fuel produced. Water is produced in the reverse water gas shift process that can be recycled back to the electrolysis step.

There is some evidence that FT jet fuels may reduce soot formation compared to crude oil-based jet fuels []. With fewer soot particles, it is likely but not yet proven that jet planes would form thinner, less dense contrails and cirrus clouds, which would reduce their contribution to global warming.

Research and Development Perspective

Power to jet fuel is a category 1 technology, a technology in the research, development and deployment stage. The individual components (electrolysis, RWGS, FT synthesis) of the process are mature technologies but they have not been combined nor commercialized at the scale envisioned for this technology. There is significant uncertainty with respect to the performance and costs of the technology.

There is potential to improve yields and reduce costs as more experience with the technology is gained from a demonstration facility and then the technology is scaled to commercial plants.

Examples of Market Standard Technology

There are two companies that have built power to FT fuel pilot systems. Both systems are capable of producing about 160 litres of liquid fuels per day.

Sunfire Gmbh develops and manufactures high-temperature electrolysers (SOEC) and high-temperature fuel cells (SOFC). They work with Ineratec Gmbh who convert the hydrogen and CO2 to liquid fuels. They have also worked with Climeworks to include direct air capture to produce the CO2 for the system. KIT have also been involved with the German Government funded P2X Kopernikus project [].

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https://ineratec.de/en/home

The other company that has developed a pilot plant is Carbon Engineering in Canada. Their core area of expertise is the direct air capture system and they have partnered with Hydrogenics for the electrolysis and Greyrock for the FT synthesis step.

Carbon Engineering Ltd.

PO Box 187,

37322 Galbraith Road,

Squamish, B.C.,

Canada

V8B 0A2

https://carbonengineering.com

Greyrock

2020 L Street, Suite 120

Sacramento, California 95811

http://www.greyrock.com/

Sasol and Haldor Topsoe, have entered into a collaboration agreement to offer G2L™-single-point licensing of GTL solutions to produce diesel, kerosene and naphtha from natural gas. The two companies have worked together on numerous GTL projects and technologies, and Topsoe's SynCOR™ technologies and Sasol's Fischer-Tropsch technologies have been licensed into several world-scale GTL ventures. Under the collaboration agreement, the companies will continue to offer these core technologies and will now also provide Topsoe's hydroprocessing and hydrogen technologies.

Haldor Topsoe A/S (HQ)

Haldor Topsøes Allé 1

DK-2800

Kgs. Lyngby

Denmark

https://www.topsoe.com/

Prediction of Performance and Cost

There have been several techno-economic studies done on the concept of power to jet fuel. In addition to the technical analyses already identified in this chapter the report by Frontier Economics for Agora Energiewende contains some economic analysis for the years 2020, 2030, and 2050 [].

The capital cost estimates that have been reported in the literature could be categorized as Class 5 or Class 4 estimates []. The Cost Estimate Classification System maps the phases and stages of project cost estimating together with a generic maturity and quality matrix,

which can be applied across a wide variety of industries. The classes range from 1 (Check Estimate or Bid/Tender with Detailed Unit Cost and Detailed Take- Off) to class 5 (Concept screening using factored parametric models or judgement).

Class 5 estimates have uncertainty on the low end of -20 to -50% and on the high end of +30 to +100%. Class 4 capital cost estimates are feasibility type estimates with slightly narrower ranges of -15 to -30% on the low end and +20 to +50% on the high end of the range.

Uncertainty

There is significant uncertainty with respect to both economics and the performance due to the stage of the development of the technology. There is a significant range for performance and costs found in the literature.

The capital cost estimates have a wide range due to the early stage of development of the concepts. It is more likely that capital costs will be above the central estimates rather than below the estimates.

Both of the main technologies (electrolysis and FT synthesis) have uncertainty. In the case of the electrolysis there are three potential technologies. The mature alkaline electrolysis which can be characterized by lower cost and efficiency, the emerging PEM systems which have higher cost and efficiency, and the solid oxide electrolytic cells which have the highest efficiency, highest cost and the lowest TRL.

FT synthesis stage which is employed commercially in large scale fossil production systems but not yet at smaller scale electrofuel systems where liquid fuel selectivity will be important to keep the system cost and complexity at reasonable levels.

Additional Remarks

There are a number of different FT catalysts that are used commercially or have been developed for commercial use. Commercial FT catalysts suppliers include Johnson Mathey, UOP, CRI/Criterion (Shell), and BASF. The catalyst can have different optimum operating conditions and produce fuels of different quality. One of the challenges of FT jet fuel is that it has a very low aromatics content and some aromatics in jet fuel are required to prevent elastomeric seals from shrinking. Currently FT jet fuel is only approved as a blend of up to 50% with the remained being crude oil derived jet fuel. The comparison of the properties of the two fuels is summarized in the following table [□]. Fossil JP-8 is the military equivalent to Jet A-1 but has some corrosion and anti-icing additives, the additives don't impact the properties in the table.

Table 4: Comparison of Fuel Properties Between FT Jet and Fossil Jet

	FT Jet Fuel	Fossil JP-8 Fuel (Jet A-1)
Flash Pt., °C	54	48
Density, kg/L	0.754	0.794
Aromatics, vol. %	<1	19.7
Sulfur, mass %	<0.002	0.08
Freeze Pt., °C	-52	-49
Visc., 40°C	1.4	1.3

Simulated Distillation., °C		
Initial Boiling Point	131	109
10%	156	159
90%	272	251
95%	282	259
Final Boiling Point	332	282
Cetane Index	70	47

The FT jet fuel has lower aromatics (which is one of the reasons that it is blended with crude oil derived jet fuel to avoid seal shrinkage), lower sulphur, and much higher cetane value.

Quantitative Description

Since there are no plants operating at a large scale, the quantitative information on the process is synthesized from the literature and presented below.

Energy/Technical Data

Reports and have the most complete set of technical and economic data and are used as the basis for the quantitative information presented. There is a fundamental difference between the assumptions made in the two reports, The Schmidt report assumes less than 4,000 hours per year of operation whereas the Mortensen report assumes almost 8,000 hours per year of operation for the FT synthesis portion of the plant. Operation of 8,000 hr/year would likely require large scale hydrogen storage, or as Mortensen assumes a hybrid concept with the alternative supply of synthesis gas from methane. The Schmidt report explicitly includes capital for hydrogen storage (in one scenario) whereas the Mortensen report is silent on the issue.

Typical Plant Size

The technology has not yet moved beyond the pilot plant stage. These pilot plants could produce 50,000 litres per year of liquid fuels if they were operated continuously. Not all of this would be jet fuel. Commercial plants will have to be larger to be economical. A possible progression of plant sizes is shown in the following table. It is assumed that the plants operate for 8,000 hours per year, that 5.5 kWh of power produces a cubic meter of hydrogen, one mole of CO2 requires 3 moles of hydrogen and the FT synthesis process has a 95% carbon efficiency.

Table 5: Typical Plant Size

Input Power MW for Hydrogen	H2 Required, tonnes/year	CO2 required tonnes/year	Liquid Fuel Litres /year
50 (Alkaline)	6,000	46,000	18,000,000
150 (PEM)	18,000	138,000	55,000,000
600 (SOEC)	72,000	552,000	222,000,000

The FT synthesis and the reverse water gas shift processes are chemical processes and should benefit from larger plants through economies of scale, although the Shell Pearl GTL plant has 24 reactors [X] which will limit the economies of scale achievable from larger reactors.

The production of hydrogen through electrolysis has limited potential for economies of scale. The more efficient PEM and SOEC electrolyzers are currently smaller than the alkaline systems and will require more modules for the same hydrogen production rate.

The economies of scale will also be impacted by the quantity of CO_2 available at a single site unless CO_2 pipelines are developed to collect CO_2 from multiple sources.

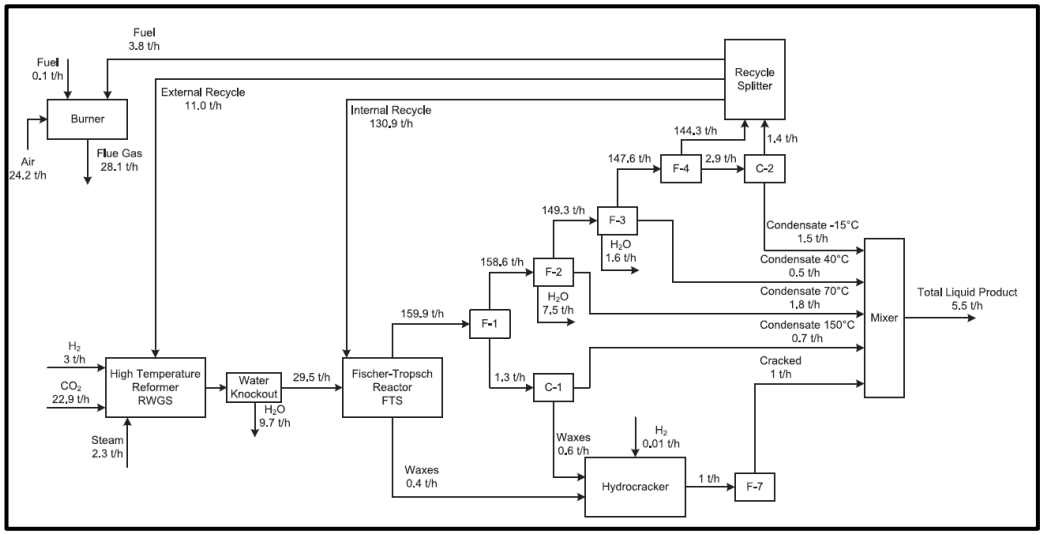
Input and Output

The high-level input and output are shown in Table 5. The detailed output will depend on the design of the FT synthesis process as shown in Table 1.

District heat in the form of steam can be recovered from the FT synthesis step (~15% of energy in the FT products). It can also be recovered from the electrolysis stage. The quantity and quality will be a function of the type of electrolyzer used.

Konig [X] developed a detailed mass balance based on the process simulations undertaken in his study. That figure is shown in Figure 3. There is a small amount of external fuel for the system and some steam, although it should be possible to generate the steam internally.

Figure 3: Mass Balance



Forced and Planned Outage

The Mortensen report assumed that the plant has 93% online availability. This is 340 days per year. The Schmidt report [272] appears to be based on only 3,750 hours per year of operation. They include some capital for hydrogen storage when the hydrogen is produced by low temperature electrolysis but not when it is produced by high temperature electrolysis (SOEC). Hydrogen production by SOEC would utilize the excess heat from the FT synthesis to produce the hydrogen, thus both systems must operate at the same time.

Technical Lifetime

Plants of this type would normally be designed for at least a 20-year lifetime. The Mortensen techno-economic study assumed a 25-year life.

Construction Time

Construction periods of 2-3 years are typical for large complex thermochemical production facilities. Smaller plants will take less time and there may be some technology learning with multiple plants.

Financial Data

There is limited financial data available for the process. The information from the two studies [] [] are mostly used below.

Investment Costs

There are two other estimates of capital costs, Schmidt also published a paper [] in addition to the report for the German government, most of the information is the same except that the paper also had a capital cost for a 60 MW plant. The Agora [] report also has some capital cost estimates but with very little detail. The various estimates are shown in the following table. The plant size is the MW of power input. Schmidt assumes very little technology learning between 2020 and 2050.

Table 6: Capital Costs

	Plant Size, MW	Year	Capital, million €	Capital per MW
Schmidt []	588	2020	308	0.52
Schmidt []	613	2040	322	0.52
Schmidt []	60	2020	84	1.4
Schmidt []	588	2050	308	0.52
Mortensen []	449	2030	407	0.91
Agora []	Not stated	2020	Not stated	1.3
Agora []	Not stated	2030	Not stated	1.1
Agora []	Not stated	2050	Not stated	0.83

The breakdown of the capital cost in the Schmidt paper is shown in the following table. These are for the utilization of a concentrated source of CO_2 and exclude any capital costs for CO_2 capture and concentration.

Table 7: Capital Cost Breakdown

	Low temperature Electrolysis	High temperature Electrolysis
Power Input, MW	588	613
Fuel Output, kt/year	97	123
Efficiency	53%	64%

Electrolysis, M€	140	159
H_2 Storage, M€	30	0
CO_2 Supply, M€	45	53
Synthesis and Conditioning, M€	94	111
Total, M€	308	322

Operating and Maintenance Costs

There is even less information on the operating and maintenance costs. Mortensen assumed that the Fixed O&M costs were 5% of the capital costs and that the variable costs were 1.5% of the capital costs. Agora assumed that operating costs were 3% of the capital costs but did not differentiate between fixed and variable costs.

Start-up Costs

None of the reports identified start-up costs or time. We have assumed that 3 months are required, similar to the assumptions made in other chapters.

Technology Specific Data

There is very little technology specific data available for this process. Wu et al [] reported some of the physical properties of gas to liquids diesel fuels. These values are shown in Table 8.

Table 8: Properties of FT Fuels

Property	Units	Value
Density	Kg/m3	779
Lower Heating Value	MJ/kg	43.6
	MJ/litre	34.0
	kWh/litre	9.44
Carbon	Wt %	84.9
Hydrogen	Wt %	15.1

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101 FAME Biodiesel (Vegetable Oil)

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Amendments after publication date

Date	Ref.	Description
12 2018		Datasheet revised

Qualitative Description

Fatty Acid Methyl Esters (FAME) otherwise known as biodiesel can be produced from vegetable oils such as rapeseed and soy oil as well as from used cooking oil and recycled animal fats. This section discusses the production from vegetable oils and the next section discusses the production from used cooking oil and animal fats as the feedstock free fatty acid (FFA) content has some impact on capital and operating costs.

Biodiesel production and use began in Europe in the early 1990's and that region is still the leading biodiesel marketer and producer. Biodiesel production has since expanded to all regions of the world and there is some international trade in both biodiesel feedstocks and biodiesel.

The European Biodiesel Board reported that the FAME production capacity in Denmark in 2016 was 250,000 tonnes.

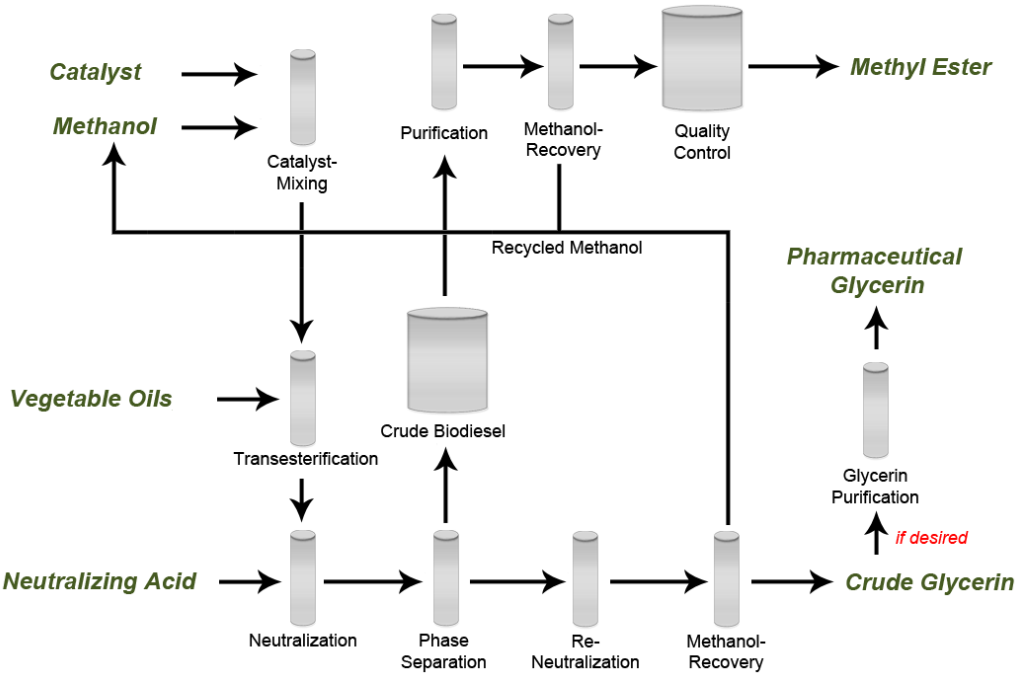
Brief Technology Description

The production of biodiesel, or methyl esters, is a well-known process. Vegetable oils are mixed with methanol in the presence of a catalyst at moderate pressure and temperatures to produce biodiesel and glycerine. Since the methanol is not soluble in the oil, this reaction will proceed either exceedingly slowly or not at all, so heat, as well as catalysts (acid and/or base) are used to speed the reaction. Almost all biodiesel produced from virgin vegetable oils uses the base-catalyzed technique as it is the most economical process for treating virgin vegetable oils, requiring only low temperatures and pressures and producing over 98% conversion yield (provided the starting oil is low in moisture and free fatty acids).

After the vegetable oil and methanol react in the presence of a catalyst to produce the biodiesel, the product goes through a number of separation and purification steps to recover the excess methanol for re-use, isolate the glycerine for potential upgrading and ensue that the methyl ester meets the required specifications.

The general biodiesel production process is shown in the following figure.

Figure 1: Biodiesel Production Process



There are other processing schemes that can be used to produce FAME from vegetable oils including the use of heterogeneous catalysts, enzymes instead of chemical catalysts and a supercritical process. This catalyst-free method for transesterification uses supercritical methanol at high temperatures and pressures in a continuous process. In the supercritical state, the oil and methanol are in a single phase, and reaction occurs spontaneously and rapidly. The process can tolerate water in the feedstock; free fatty acids are converted to methyl esters instead of soap, so a wide variety of feedstocks can be used. Also, the catalyst removal step is eliminated. High temperatures and pressures are required, but energy costs of production are similar or less than catalytic production routes. These alternative processes are not yet widely practiced.

Input

The primary inputs to the production process are the vegetable oil, methanol, electricity, some thermal energy, the catalyst, and some acids and bases to treat the feedstocks and finished products.

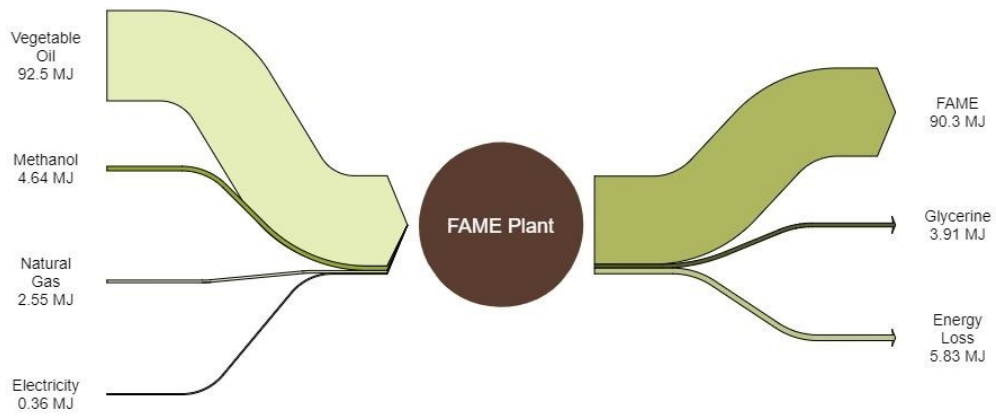
Output

The plants produce FAME, glycerine (of various qualities), and in some cases potassium salts that can be sold as fertilizer.

Energy Balance

The energy balance is shown graphically in the following figure. The external energy inputs, in the form of heat and power, are relatively small and the process is quite efficient. The sum of the inputs totals 100 MJ.

Figure 2: Vegetable Oil FAME Energy Balance



A small portion, about 5%, of the energy losses has the potential to be recovered as low-level district heat [1].

Typical Capacities

Production capacities for individual plants can range from less than 10,000 tonnes per year to almost 600,000 tonnes per year (10 to 750 MW).

It was reported that in 2010, 46 FAME plants in Germany had a production capacity of almost 5 million tonnes per year [2]. The average plant size was 100,000 tonnes but more than 50% of the plant production capacity was found in 9 plants with more than 200,000 tonnes of capacity each.

Regulation Ability

FAME plants can be either batch or continuous processes. The time between batches regulates the total production from batch plants. Larger plants are generally continuous processes. The throughput can be regulated to a degree but process start-up is generally fast so regulating the days of operation is also a viable means to regulate production.

Space Requirements

A 100,000 tonne per year facility can be located on a site of less than 5 ha. This is equivalent to 0.04 ha/million litres of fuel, or 400 m2/MW. Factors that impact the required area include storage capacity of inputs and outputs and whether or not rail access is included at the facility.

Advantages/Disadvantages

FAME biodiesel is a relatively simple process that is well proven commercially. The capital costs are relatively low and the non-feedstock operating costs are also reasonable.

The feedstock costs are high compared to crude oil. The finished fuel has less desirable cold weather properties than fossil diesel fuel limiting its potential inclusion rates in diesel fuel in cold weather.

Environment

With low energy use the GHG emissions from the biodiesel plants are relatively minor. The GHG emissions of the methanol must also be included in the carbon footprint. Methanol and the process chemicals used have safety hazards associated with their use but these are well known and there are establishes procedure to accommodate their safe use.

Research and Development Perspective

Vegetable oil FAME is a Category 4 technology, a commercial technology with large scale deployment. The potential for significant improvements in capital and operating costs is limited. The price and performance of the technology today is well known and normally only incremental improvements would be expected. Therefore, the future price and performance may also be projected with a relatively high level of certainty.

Examples of Market Standard Technology

A number of companies provide biodiesel technology packages. Desmet Ballestra [3] is one such company.

Desmet Ballestra Group N.V.
Fountain Plaza Office Park, Building 503
Belgicastraat 3
B - 1930 ZAVENTEM
Belgium

They have been in involved in over 85 plants around the world with capacity in excess of 12 million tonnes/year. Their plants have capacities from 50,000 to 300,000 tonnes per year.

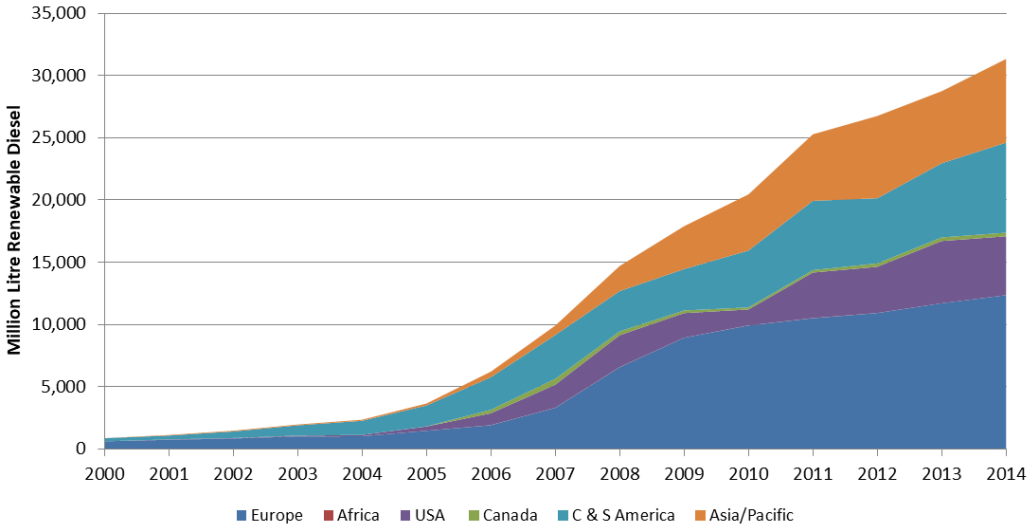
Predication of Performance and Cost

Biodiesel production from vegetable oils is mature technology. Conversion performance in most plants approaches the theoretical maximum conversion rates. There is therefore limited development potential for the technology.

Worldwide production rates are about 30 billion litres per year with the cumulative production between 2000 and 2014 being 195 billion litres as shown in the following figure [4].

At current production rates the cumulative production since 2000 will double in six years. Berghout investigated the Progress ratio for the German Biodiesel industry between 1991 and 2004 and found that the limited data suggested a value of 0.97. This is a very low value and may be due to the fact that while the industry is new, the processing steps are based on mature chemical processes with limited potential for improvement. The cost reduction would also not apply to feedstock so the impact on production economics will be marginal.

Figure 3: World Biodiesel Production



Uncertainty

Given the mature status of the technology there are low levels of uncertainty associated with the data that is collected, but there will always be some uncertainty regarding future projections.

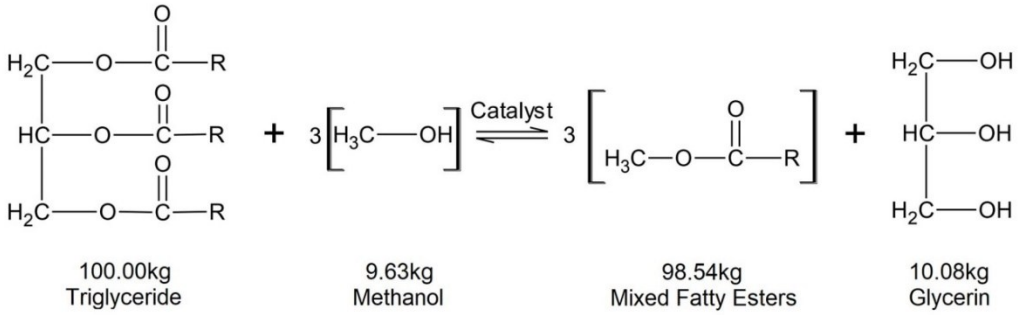
Additional Remarks

This technology is commercially practiced in Denmark today by Emmelev A/S. Denmark blends FAME and HVO at about 7% by energy content in diesel fuel [5]. The FAME could be from vegetable oil or from UCO.

Quantitative Description

The basic chemical reaction is depicted below. One hundred kilograms of a fat or oil is reacted with 10 kilograms of methanol in the presence of a catalyst to produce 10 kilograms of glycerine and 100 kilograms of methyl esters or biodiesel. The methanol is charged in excess to assist in quick conversion and recovered for reuse. The catalyst is usually sodium or potassium methoxide which has already been mixed with the methanol. R1, R2, and R3 indicate the fatty acid chains associated with an individual fat.

Figure 4: Basic Biodiesel Reactions



Typical Plant Size

While there has been a wide range of plants built in the past, large scale plants can enjoy some economies of scale. The one vegetable oil biodiesel plant in Denmark (Emmelev) has

a production capacity of 100,000 tonnes per year [6]. This is the average plant size reported in Germany in 2010 and will be used for the data sheet.

Input and Output

The typical mass and energy balances for the primary materials and energy sources are shown in the following table. The data is from a 2016 Survey of vegetable oil biodiesel plants operating in the United States [7]. The data is from 12 plants with production capacities ranging from 40,000 to 330,000 tonnes per year. The plants operated at an average of 74% of capacity.

Table 1: Mass and Energy Flows

	Inputs	Outputs
Vegetable oil	0.893 kg	
Methanol	0.086 kg	
FAME (one litre)		0.88 kg
Glycerine		0.09 kg
Electricity	0.036 kWh	
Natural gas	0.93 MJ (LHV)	

The input and output data for future years assumes that the industry average moves towards the current best in class values.

Forced and Planned Outage

Biodiesel plants can operate continuously with limited downtime for planned maintenance. The NBB energy survey found that plants operated from about 40 to over 95% of production capacity in 2015.

The most likely reasons for forced outage are either a lack of feedstock or a lack of markets for the product.

Technical Lifetime

The first European plants were built in the 1990s and many of those are still operating more than 20 years later. Processing conditions are relatively mild with moderate pressures and temperatures for most processes. The corrosive nature of some of the reagents (methanol, acids and bases) are well understood and manageable.

Construction Time

Large scale plants should have a typical construction time of 12 to 18 months from the start of site preparation to the beginning of production.

Financial Data

Vegetable oil FAME plants are commercial technologies with capital and operating costs that are relatively well understood. There have been relatively few FAME plants constructed in recent years as the global industry is operating at less than full capacity.

Investment Costs

Most European vegetable oil FAME plants were built a decade ago. The most recent vegetable oil biodiesel plant that has been announced is a 200,000 tonne soybean plant to be built by Cargill in Kansas, USA. The plant will be built on the site of an existing soybean crushing facility and will replace an existing soy oil refinery. The capital cost is reported to be $90 million US$ [8].

Converting this cost to Europe, the location factor, exchange rate, and greenfield factor is applied as shown below:

$$US\ capital\ cost \cdot greenfield\ factor \cdot currency\ factor \cdot location\ factor$$

$$\$90{,}000{,}000 \cdot 1.2 \cdot 0.85 \cdot 1.21 = €111\ million$$

It is estimated that a 100,000 tonne per year plant would have a capital cost of €64 million. This is $0.66/litre, which is in the middle of the range reported by Irena [9].

Future investment costs are not expected to decline significantly as few new plants are being built. Berghout, with limited data, concluded that learning-by-upscaling to be the primary and learning-by doing the secondary driver behind reductions in specific investment costs over time for German biodiesel plants [10]. Since the same plant size has been used for the study period of 2015 to 2050 only a small reduction (~5%) in investment costs is projected by 2050.

Operating and Maintenance Costs

The production cost profile is dominated by the feedstock costs. Feedstock costs are highly variable and can increase or decrease in any given year due to the local supply and demand situation.

Iowa State University [11] maintains a biodiesel profitability model for a 100,000 tonne/year soy biodiesel plant. The 2016 data has been used and the power and natural gas costs have been adjusted to Danish values. The capital cost has been adjusted to 64 million Euro to match the value above. The results for 2016 are shown in the following table 2.

The other variable costs include chemicals, power, water, repairs and maintenance, water, transportation and other costs. The fixed costs include depreciation, interest costs, labour and management, property taxes, insurance, and marketing and procurement costs.

Berghout found a very low Progress Ratio (97.4%) and a low R2 value (0.65) for the FAME processing costs, suggesting little potential for reductions in O&M costs over time.

The impact on production costs as the future industry moves towards the best practices of the existing plants is limited as gains in yield will impact the feedstock costs which are not included here; the natural gas, power and methanol costs which are a small portion of the operating costs.

Table 2: Production Costs Vegetable oil FAME

Parameter	Costs (Eur/litre)
Natural gas	0.01
Methanol	0.02
Other Variable Costs	0.06
Fixed Costs	0.02
Depreciation and Interest	0.06
Total Costs ex feedstock	0.17

Start-up Costs

The O&M costs stated in this catalogue includes start-up costs and considers a typical number of start-ups and shut-downs. Therefore, the start-up costs should not be specifically included in more general analyses.

Technology Specific Data

Some of the key fuel properties of biodiesel are shown in the following table.

Table 3: Vegetable Oil Biodiesel Properties

Property	Value
Density, kg/m^3	885
LHV, MJ/kg	37.2
LHV, MJ/litre	32.7
Oxygen content	11%
Cetane number	~55
Cloud point, C	~-3

Data sheet

The quantitative data reported above is summarized in the data sheets. This information is determined from the NBB operating data for vegetable oil feedstock plants and the operating costs estimated from the Iowa State model. The best biodiesel plants operate a close to theoretical yields and methanol consumption rates, and with low energy consumption values. The 2015 data is based on the 2015 NBB data, it has been assumed that by 2050 the industry moves to have an average performance the same as the best plant in 2015. The 2020 and 2030 values are estimated from the transition from the 2015 to 2050 values.

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102 FAME Biodiesel (Used Cooking Oil and Animal Fat)

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Date	Ref.	Description
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Brief Technology Description

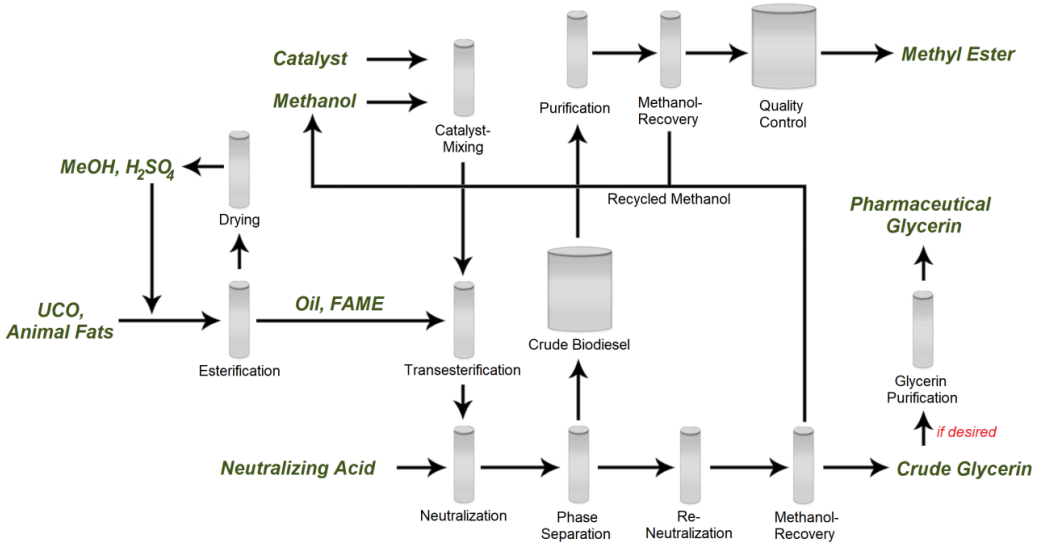
Used Cooking Oil (UCO) and animal fats can also be used to produce FAME. This section of the report discusses FAME production using these two feedstocks. The primary technical differentiation factor for these two feedstocks is the free fatty acid (FFA) content of the feedstocks. Free fatty acids can be produced when the lipids are heated, as they are when the cooking oils are used or when the animal carcasses are rendered.

UCO and animal fats are generally less expensive feedstocks than vegetable oils. UCO is also treated as a waste material in most regulatory systems so it generates a better carbon intensity score and the volume is double counted in some systems. These attributes make it an attractive feedstock, more than outweighing some processing challenges described below.

The base catalyzed transesterification process that produces FAME or biodiesel from vegetable oils is only effective with triglycerides. FFAs require an acid catalyzed esterification reaction for commercial production. This reaction is slower and involves an extra processing step. Some UCO and animal fat biodiesel producers accept the lower yield that results from not converting the FFAs to biodiesel and some use the two-step process that converts both triglycerides and FFAs to biodiesel.

The two-step process is shown below, the basic difference between this and the process used for vegetable oils is the initial esterification steps.

Figure 1: Two Step Biodiesel Production Process



Input

The primary inputs to the production process are the UCO or animal fat, methanol, electricity, some thermal energy, the catalyst(s), and some acids and bases to treat the feedstocks and finished products.

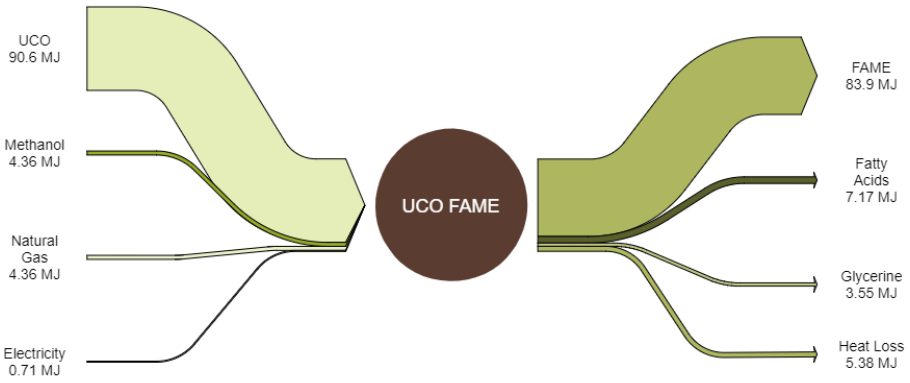
Output

The plants produce FAME, glycerine (of various qualities), in some cases potassium salts that can be sold as fertilizer, and unreacted FFAs.

Energy Balance

The average energy balance for the year 2015 from 15 US biodiesel plants processing mixed feedstocks is shown in the following figure [1]. Most of these plants do not do the two-step process and accept the lower yield due to the higher FFA content of the feedstock.

Figure 2: UCO and Animal Fat FAME Energy Balance



The potential for recovering some heat for district heating is similar to that of the vegetable oil FAME process, about 5% of the heat loss.

Typical Capacities

The size of UCO and animal fat plants is slightly smaller than the size of vegetable oil plants. Production capacities of up to 150,000 tonnes per year (185 MW) have been built in Europe.

The Daka ecoMotion plant in Denmark has a capacity of 50,000 tonnes per year (60 MW). This plant uses the two-step process of esterification followed by transesterification. The plant was designed by BDI in 2008.

Regulation Ability

FAME plants can be either batch or continuous processes. The time between batches regulates the total production from batch plants. Larger plants are generally continuous processes. The throughput can be regulated to a degree but process start-up is generally fast so regulating the days of operation is also a viable means to regulate production.

Space Requirements

The Daka plant in Denmark sits on 2.1 ha of land. This is 0.038 ha/million litres or 340 m2/MW of biodiesel.

All of the materials in and out of the plant come by truck. The plant is shown in the following figure. The actual storage and processing equipment cover less than 50% of the site.

Figure 3: Daka FAME Plant



Advantages/Disadvantages

UCO and animal fats have many of the same advantages and disadvantages as vegetable oil FAME when compared to petroleum diesel fuel.

When UCO and animal fat FAME is compared to vegetable oil FAME the feedstock costs are usually lower but capital and operating costs are slightly higher. The cold weather properties of UCO and animal fat FAME are usually slightly less attractive than the vegetable oil FAME.

Environment

With low energy use the emissions from the biodiesel plants are relatively minor. Methanol and the process chemicals used have safety hazards associated with their use but these are well known and there are establishes procedure to accommodate their safe use.

The energy content of the UCO and animal fat biodiesel is about 20 times the power and natural gas energy consumed by the plant.

Research and Development Perspective

Used Cooking Oil (UCO) and animal fat FAME is a Category 4 technology, a commercial technology with large scale deployment. The price and performance of the technology today is well known and normally only incremental improvements would be expected. Therefore, the future price and performance may also be projected with a relatively high level of certainty.

The potential for significant improvements in capital and operating costs is limited. There is work being undertaken on making the use of lower quality feedstocks commercially attractive. These materials (e.g. brown grease) can have very high FFA levels and contain other contaminants such as sulphur and chlorine.

Examples of Market Standard Technology

One of the leaders with the multi-feedstock biodiesel process is the Austrian company, BDI Bioenergy International.

BDI - BioEnergy International AG
Parkring 18, 8074
Raaba-Grambach, Austria
They have built more than 30 biodiesel plants, mostly in Europe, since 1991.

Predication of Performance and Cost

Biodiesel production from UCO and animal fats is relatively mature technology. Conversion performance in the best plants that use the two-step process approaches the theoretical maximum conversion rates. There is therefore limited development potential for the technology.

Berghout [2] investigated the Progress Ratio for the German Biodiesel industry between 1991 and 2004 and found that the limited data suggested a value of 0.97. This is a very low value and may be due to the fact that while the industry is new, the processing steps are based on mature chemical processes with limited potential for improvement. The cost reduction would also not apply to feedstock so the impact on production economics will be marginal.

Uncertainty

Given the mature status of the technology there are relatively low levels of uncertainty associated with the data that is collected. One area of potential uncertainty is the proportion of plants that practice esterification in the United States (where the plant data comes from) vs the EU.

Additional Remarks

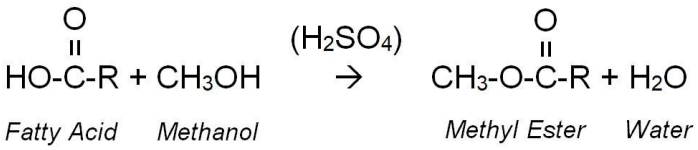
This technology is commercially practiced at the Daka plant in Denmark. Europe is an importer of used cooking oil as demand for the product exceeds the quantity supplied in Europe [2]. It is also reported that that there are limited opportunities to increase the collection rate of UCO in Europe [3], resulting in about 60% of the UCO consumed in Europe is imported into the EU.

Quantitative Description

Additional quantitative information on the production of UCO or animal fat FAME is presented in the following sections.

The esterification reaction that is practiced with some technologies prior to the transesterification process is shown below. In this reaction there is no glycerine produced but there is some water produced.

Figure 4: Esterification Reaction



Typical Plant Size

The plants can range in size from 5,000 to over 100,000 tonnes per year of capacity. Older plants generally have lower production rates as market access for the production would have been an issue when they were built.

A plant of 50,000 tonne/year is used for the data sheet. This is in the middle of the range of plants built and is the same size as the existing Danish plant.

Input and Output

The typical mass and energy balances for the primary materials and energy sources are shown in the following table. The data is from a 2016 Survey of vegetable oil biodiesel plants operating in the United States [1]. The data is from 15 plants with production capacities ranging from 3,000 to 200,000 tonnes per year. The plants operated at an average of 73% of capacity. The primary feedstocks for these plants were animal fats, distillers' corn oil (extracted from the stillage of corn ethanol plants), and used cooking oil. About 10% of the feed for these mixed feedstock plants was vegetable oils.

Table 1: Mass and Energy Flows

	Inputs	Outputs
Vegetable oil	0.977kg	
Methanol	0.087 kg	
FAME (one litre)		0.88 kg
Glycerine		0.09 kg
Electricity	0.08 kWh	
Natural gas	1.71 MJ (LHV)	

The FAME yield is lower, indicating that many of the plants do not practice the esterification step and the energy use is about double that of the vegetable oil plants. The energy use would be higher if more of the plants were batch processes rather than continuous operations.

Forced and Planned Outage

Biodiesel plants can operate continuously with limited downtime for planned maintenance. The NBB energy survey found that plants operated from about 40 to over 97% of production capacity in 2015.

The most likely reasons for forced outage are either a lack of feedstock or a lack of markets for the product.

Technical Lifetime

The first European plants were built in the 1990s and many of those are still operating more than 20 years later. Processing conditions are relatively mild with moderate pressures and temperatures for most processes. The corrosive nature of some of the reagents (methanol, acids and bases) are well understood and manageable.

Construction Time

Large scale plants should have a typical construction time of 12 to 18 months from the start of site preparation to the beginning of production.

Financial Data

Information on capital and operating costs are presented below.

Investment Costs

Argent Energy built a 75,000 tonne per year multi-feedstock facility near Chester, UK in 2015-2016. The reported cost was £75 million (€100 million) [4]. This is considerably more expensive than the recent cost estimate for a soybean biodiesel plant in the United States. The plant has significant pretreatment facilities and can reportedly process 100% FFA feedstocks. Given the extra processing required for the plants processing higher FFA feedstocks some increase in price is expected but this €1.17/litre, almost double the soy oil cost.

The future capital costs have been reduced by about 25% by 2050. The rationale for the larger percent cost reduction than a vegetable oil FAME plant is that a greater yield improvement is expected as the industry moves towards the best-in-class performance, due to the higher capital costs there should be more room for improvement, and there is greater emphasis being place on the use of “waste” feedstocks by many governments.

Operating and Maintenance Costs

There is no public source of economic data for UCO or animal fat biodiesel plants like the Iowa State information on soybean biodiesel plants. Given the higher energy requirements, the higher capital costs, the potentially lower yields the operating costs should be higher than they are for a vegetable oil plant.

A German biodiesel Petrotec, a UCO biodiesel producer, used to supply some financial information to their shareholders. The last information available is for 2014 [5], before they were purchased by REG Inc. from the United States. The plant has a capacity of 185,000 tonnes per year and operated at 75% capacity in 2014. The information that can be developed from the annual report is shown in the following table.

Table 2: UCO Biodiesel Operating Costs

Category	Euro /litre
Services	0.05
Labour	0.05
Operating Costs	0.04

Marketing Costs	0.03
Admin Costs	0.01
Legal Costs	0.01
Other Expenses	0.01
Interest	0.01
Total	21

The fixed costs (labour, marketing, admin, legal, other, interest) total €0.13/litre. There is no breakdown of feedstock costs from other supplies in the published information. The vegetable oil plant has variable operating costs of €0.05/litre, the chemical costs will be higher for the UCO facility, it is estimated that the variable operating costs of €0.07/litre.

As with the vegetable oil FAME plants only limited improvements in the fixed and variable production costs are forecast due to the low Progress ratio found by Berghout.

Start-up Costs

The O&M costs stated in this catalogue includes start-up costs and considers a typical number of start-ups and shut-downs. Therefore, the start-up costs should not be specifically included in more general analyses.

Technology Specific Data

Table 3: UCO and Animal Fat Biodiesel Properties

Property	Value
Density, kg/m^3	885
LHV, MJ/kg	37.2
LHV, MJ/litre	32.7
Oxygen content	11%
Cetane number	50-60
Cloud point, C	0-10 C

Data sheet

The quantitative information collected for a UCO biodiesel facility is shown in the datasheets. This information is determined from the NBB operating data for mixed feedstock plants and the operating costs estimated from the Petrotec annual report. The best biodiesel plants operate a close to theoretical yields and methanol consumption rates, and with low energy consumption values. The 2015 data is based on the 2015 NBB data; it has been assumed that by 2050 the industry moves to have an average performance the same as the best plant in 2015. The 2020 and 2030 values are estimated from the transition from the 2015 to 2050 values.

References

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103 HVO Jet Fuel

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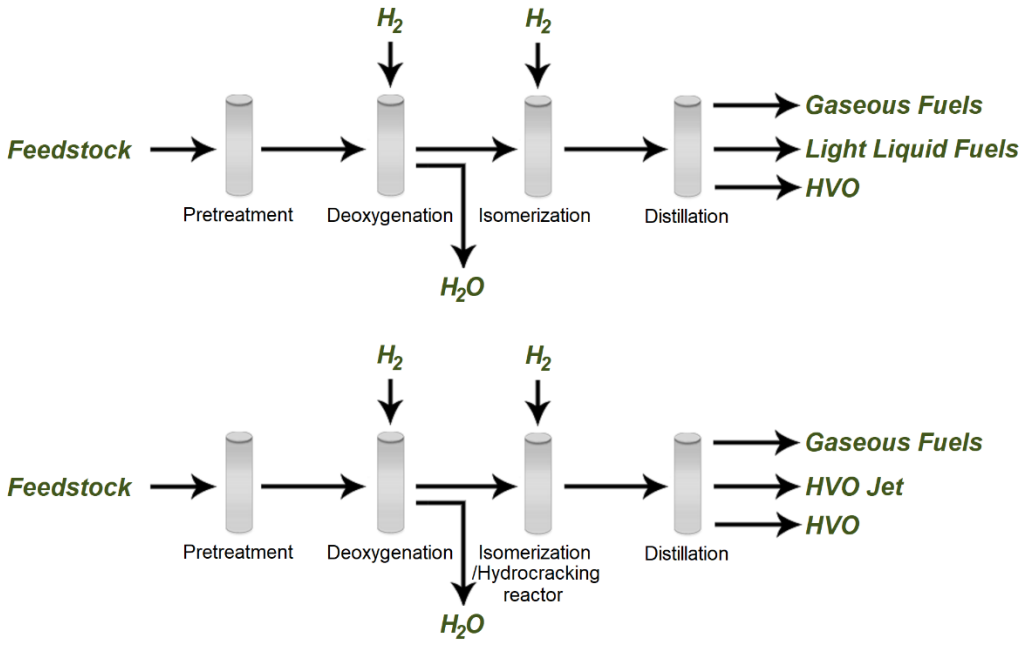
Amendments after publication date

Date	Ref.	Description
12 2018		Datasheet revised

Qualitative Description

This technology is very similar to the HVO renewable diesel process with one exception, in addition to the isomerization step there is a hydro-cracking step to produce products in the jet fuel boiling range rather than the diesel fuel range. This can be seen in the following figure which is a comparison of the UOP process diagrams for both technologies.

Figure 1: HVO Diesel vs Jet Process



There is one dedicated HVO jet facility in the world. It is the 130,000 tonne per year AltAir plant in Los Angeles, California. It employs the UOP process shown above. Detailed information on the AltAir operations is not publicly available and thus there is more reliance on secondary information in the description than there was for the previous section that considered the production of renewable diesel fuel.

This technology would be classified as being between the Pioneering and the Early Commercial phase.

Brief Technology Description

To meet the jet fuel specification, the produced bio-jet fuel has to have not only a high flash point, but also good cold flow properties. Therefore, it is required to hydro-isomerize and hydrocrack the normal paraffins produced from deoxygenation to a product with carbon chains ranging from C9 to C15. The hydrocracking stage results in some yield loss of jet and diesel fuel with increased production of naphtha and fuel gas.

The AltAir facility produces Renewable F-76 (a marine distillate fuel) and JP-5 (a military jet fuel) for the US military and Renewable Jet, Diesel, and Gasoline [1]. The company has not provided information on what percentage of each fuel is produced. The facility is using animal fats and used cooking oil as feedstocks. The technology should work for any of the same oils and fats used for renewable diesel fuels.

Compared to the HVO for diesel fuel described in the previous section, the yield of jet and diesel fuel is expected to be lower and the hydrogen demand higher when jet fuel is produced as a result of the additional hydrocracking step.

Like most of the operating HVO plants, the AltAir facility takes advantage of existing equipment and services to reduce costs and risks.

Input

The primary inputs to the production process are the fats and oils and hydrogen, along with some electricity, thermal energy, and chemicals.

Output

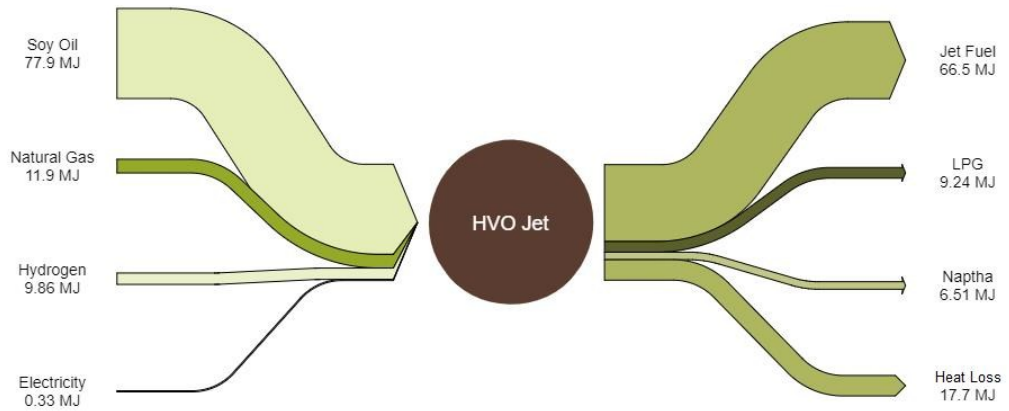
The outputs from the process are the jet fuel, some diesel fuel, some fuel gas (a mixture of hydrogen and light hydrocarbons), propane, and a light liquid stream that can be used as fuel or as a renewable gasoline blending component.

Energy Balance

Several of the analysis of HVO Jet fuel have used the GREET model data for the energy balance [2] [3]. Other studies have investigated the optimization of the process and reported significant process variability [4] [5].

The energy balance using the GREET information is shown in the following figure.

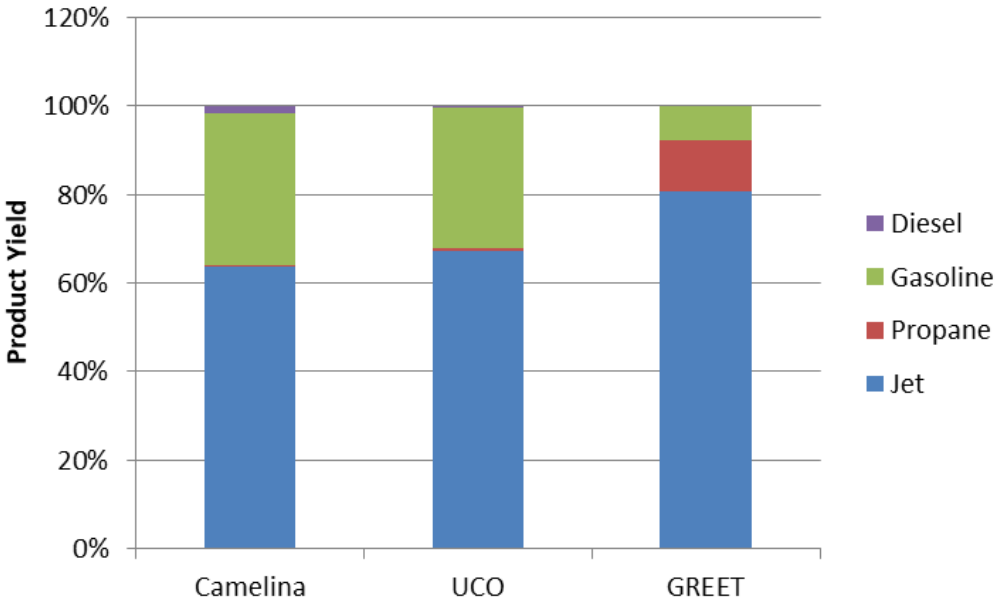
Figure 2: HVO Jet Fuel Energy Balance



As with the HVO technology in the previous chapter it is likely that 15% of the heat loss can be recovered as a high temperature stream that could be used for district heating.

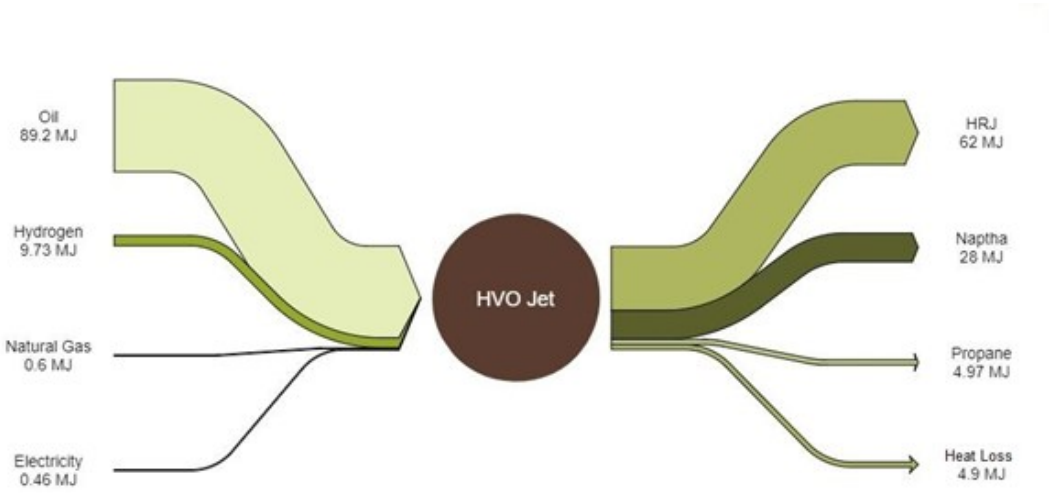
The product yields for two feedstocks in the Tao paper [4], which are derived from Aspen modelling, are very different than the GREET product yields as shown in the following figure. Different product yields will impact the energy balance for the process.

Figure 3: Comparison of Product Yields



An earlier analysis of the process published in 2010 also showed much lower yields than is used in GREET [6]. This report was based on a theoretical analysis of the process. The energy balance from this work is shown in the following figure.

Figure 4: Alternative HVO Jet Energy Balance



Typical Capacities
Demand for biojet fuel is potentially much larger than the capacity of the single plant supplying the product. The one existing plant is also small compared to the HVO plants producing diesel fuel. There is no reason why the typical plant capacities won't be similar to the HVO for diesel plants, 150 to 1,000 million litres per year.

Regulation Ability
Very little information is available on the turn down ratios of the process. It is a process with a high operating temperature pressure and thus will likely have limited capacity to regulate throughputs.

Space Requirements
The AltAir plant is a re-purposed oil refinery so the space requirements for this facility are not likely to be representative of greenfield plants.

The Neste HVO plants in Singapore and Rotterdam and the Diamond Green Diesel and REG plants in Louisiana are greenfield plants that have some integration with nearby utility suppliers. The space required for the plants are in the 0.02 to 0.025 ha/million litres of production range (220 m2/MW). This is likely to be the same space requirements of a HVO jet plant.

Advantages/Disadvantages
HVO Jet Fuel is the only commercial process for the production of renewable jet fuel. It is fully approved up to 50% blend level with fossil jet fuel. Most of the commercial biojet flights operating in the world today use fuel from this process.

Due to lower yields and higher energy and hydrogen consumption the production costs of HVO Jet will be higher than HVO diesel fuel, which are already higher than the fossil fuel prices.

Environment
Based on the information in the peer reviewed literature the jet fuel production is more energy intensive than the production of HVO diesel fuel.

The hydrotreating catalysts have a long life and can be re-generated. There is limited waste generated as part of the process, although this can depend on the feedstock and the need to pre-treat the feedstock to remove minor contaminants.

Research and Development Perspective
Hydrogenated vegetable oil (HVO) Jet Fuel has been classed as a Category 3 technology, a commercial technology with development potential. There is one production facility in operation that is dedicated to maximize jet fuel. These technologies are deemed to have a certain development potential such as opportunities for optimizing jet fuel production with different feedstocks or possibly feedstock blends and therefore there is a considerable level of uncertainty related to future price and performance. On the other hand, the process steps that are used are all well know refinery process units which limits the potential improvements.

Examples of Market Standard Technology
The leading commercial supplier of the technology is Honeywell UOP. Their corporate office is:

UOP World Corporate Offices
25 E. Algonquin Road (Bldg A)
P.O. Box 5017

Des Plaines, IL 60017-5017
Toll Free: +1 (800) 877-6184
Phone: +1 (847) 391-2000

They have a European office at:

Noorderlaan 147
B-2030 Antwerp, Belgium
Phone: +32-3-540-9911

Predication of Performance and Cost

The production of jet fuel from fats and oils is being practised commercially in one facility, however very little information on the actual performance is available in the public domain. Therefore, it is uncertain where the technology sits on the development curve.

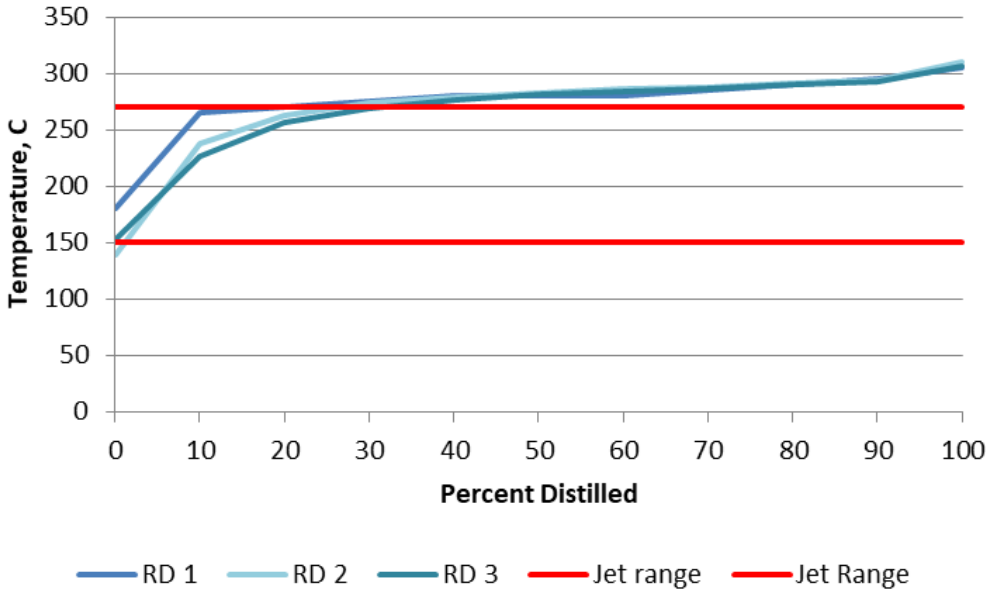
Uncertainty

There are relatively high levels of uncertainty with the performance parameters for this technology as the data in the public domain is not based on the performance of the one operating plant. That information is confidential. The information that is in the public domain from peer reviewed studies is sometimes contradictory especially with respect to product yields and energy consumption.

Additional Remarks

It is possible to produce some jet fuel from HVO plants that focus on the production of diesel fuel. There is a portion of the HVO that can be recovered by distillation that will meet the fuel specifications of the jet fuel. This is the "light" end of the diesel HVO and removing it makes the remaining product heavier. However, if there is a demand for diesel and jet fuel it is one way to produce HVO Jet without a loss of yield and large amounts of LPG or naphtha materials being produced. The distillation curves for three commercial renewable diesel fuels are compared to the typical upper and lower limit for jet fuel in the following figure. It can be seen that 10 to 20% of the HVO diesel fuel is within the jet fuel range. There is no quantitative public information on the flexibility of the operations to alter the diesel/jet fuel ratios.

Figure 5: Renewable Diesel Distillation Curves



Quantitative Description

There is very little actual data available on this process in the public domain. The information in the peer reviewed literature is often contradictory so much of the information presented below has a high degree of uncertainty.

Typical Plant Size

The output of the primary product will be 20 to 40% lower in a jet HVO plant compared to one that produced diesel fuel. The co-product output however will be higher. It is assumed that the typical plant size will be 400,000 tonnes, 80% of the typical diesel HVO plant reflecting the lower primary product yield.

Input and Output

The estimated input and output is shown in the following table. This is based on the GREET model parameters which were used for Table 1. The output is one litre of HVO jet fuel.

Table 1: Mass and Energy Flows HVO Jet

	Inputs	Outputs
Feedstock	1.07 kg	
Hydrogen	0.035 kg	
Natural gas	6.06 MJ	
Electricity	0.048 kWh	
LPG		0.142 kg
Naphtha		0.097 kg
HVO Jet (one litre)		0.77 kg

It has been assumed that no changes to the inputs and outputs are achieved over the present technology. The data in the table represents a very good mass balance with limited opportunity for improvement.

Forced and Planned Outage

HVO jet fuel plants should operate in a similar manner to HVO diesel fuel plants. Forced shutdowns should be minimal and longer turnaround should happen every four years with an annual rate of 15 days per year.

Technical Lifetime

Plant lifetimes should exceed 25 years, similar to what would be expected in a petroleum refinery.

Construction Time

Construction time of 24 to 30 months should be expected. This is longer than a FAME plant but the HVO plants are generally larger and have process vessels that operate at higher pressures and temperatures with an expectation of longer fabrication periods.

Financial Data

The financial data has been estimated based on the available information for HVO plants and the process differences to produce jet fuel instead of diesel fuel.

Investment Costs

Investment costs are 25% higher than the HVO plant due to the lower yield of the primary product. It is estimated that a 400,000 tonne per year plant would have a capital cost of €550 million. This is the same cost but lower throughput than used for the HVO plant. This is €1.06/litre.

De Jong [7] reported a range of capital costs of €200 to 644 million for a 350,000 tpy HVO Jet plant. This is €0.57 to €1.84 per litre.

The future investment cost is lower due to the increased scale of the plants, the same approach is used as was used for the HVO plants, a scaling factor of 0.6 is applied to the ratio of the plant size in the future to the current plant size.

Operating and Maintenance Costs

The operating and maintenance costs will be allocated across the jet fuel and the co-products. It is assumed that they will be the same per litre as were determined for the HVO plant.

The fixed O&M costs are €0.04 per litre and the variable O&M costs are €0.08 per litre. These costs are held constant for the future years.

Start-up Costs

The start-up costs are included in the operating costs.

Technology Specific Data

The physical properties of HVO Jet are shown in the following table.

Table 2: HVO Jet Properties

Property	Value
Density, kg/m^3	770
LHV, MJ/kg	44.0
LHV, MJ/litre	34.4
Oxygen content	0
Freezing point, C	-63 °C

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104 HVO Renewable Diesel

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Date	Ref.	Description
12 2018		Datasheet revised

Qualitative Description

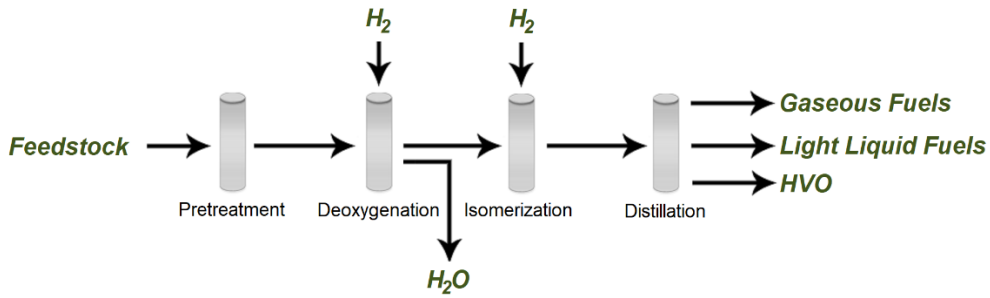
Hydrogenated vegetable oil (HVO) is also known as renewable diesel. HVO plants use the same feedstocks that are used for FAME plants including used cooking oils and animal fats, however rather than reacting the feedstock with methanol the feedstock is reacted with hydrogen in the presence of a catalyst. The operating temperatures and pressures are higher than in FAME plants.

There are small differences in hydrogen demand between the different feedstocks that are used but the variation is generally less than what is seen between process developers and between plants so only one data sheet is presented. The resulting product is oxygen free and has better fossil diesel blending properties. The product is generally isomerized to improve the cold weather properties of the fuel.

Brief Technology Description

HVO is a mixture of straight chain and branched paraffins – the simplest type of hydrocarbon molecules from the point of view of clean and complete combustion. Typical carbon numbers are C15 to C18. Paraffins exist also in fossil diesel fuels which additionally contain significant amounts of aromatics and naphthenics. Aromatics are not favorable for clean combustion. HVO is practically free of aromatics and its composition is quite similar to GTL and BTL diesel fuels made by Fischer Tropsch synthesis from natural gas and gasified biomass. The general HVO production steps are shown in the following figure.

Figure 1: HVO Production Process



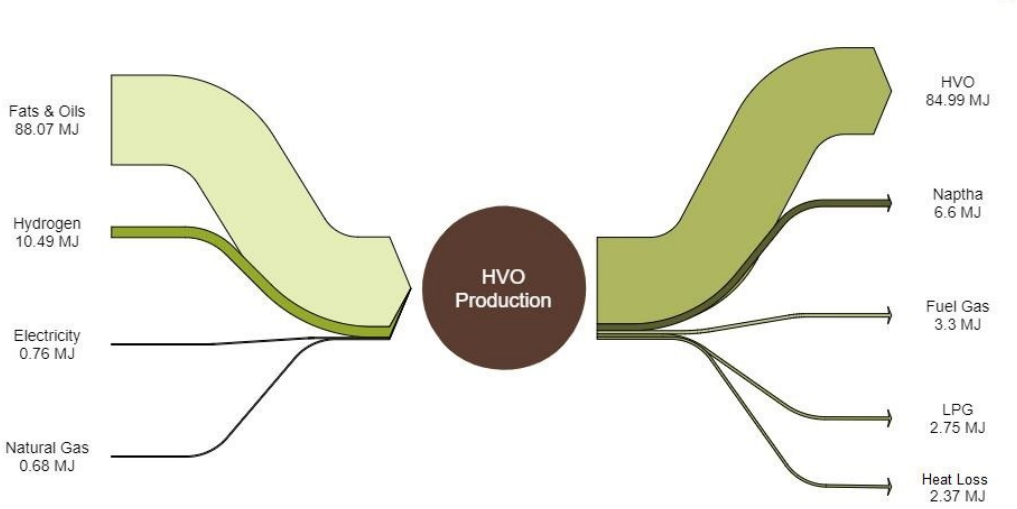
Almost all operating HVO plants have some degree of integration with nearby facilities to provide energy or hydrogen or to process co-product streams. This integration can reduce capital and operating costs and risks for new technologies.

Input
The primary inputs to the production process are the fats and oils and hydrogen, along with some electricity, thermal energy, and chemicals.

Output
The outputs from the process are the HVO, some fuel gas (a mixture of hydrogen and light hydrocarbons), propane, and a light liquid stream that can be used as fuel or as a renewable gasoline blending component.

Energy Balance
High quality public information on the energy balance of the HVO process is very limited. The following figure shows the typical energy balance based on a composite of information in the public domain [1] [2] [3]. The sum of the inputs totals 100 MJ.

Figure 2: HVO Energy Balance



There is the potential to recover about 15% of the heat loss as high temperature heat for district heating applications.

Typical Capacities
Stand-alone HVO plants range in size from 150 million litres/year (AltAir, USA) to move than 1,000 million litres per year (Neste Singapore and Rotterdam) (170 to 1,125 MW).

Regulation Ability
Very little information is available on the turn down ratios of the process. It is a process with a high operating temperature and pressure and thus will likely have limited capacity to regulate throughputs. The existing HVO plants are generally running at capacity.

Space Requirements
The Neste plants in Singapore and Rotterdam and the Diamond Green Diesel and REG plants in Louisiana are greenfield plants that have some integration with nearby utility suppliers. The space required for the plants are in the 0.02 to 0.025 ha/million litres of production range (220 m2/MW).

Advantages/Disadvantages
HVO has a number of advantages over biodiesel as a fuel and as a production process. The process is not restricted by the free fatty acid content of the feedstock and most plants process a wide range of feedstocks from vegetable oils to UCO and animal fats.

HVO can be blended with petroleum diesel fuel without any limit or labelling requirements because it is a hydrocarbon. HVO has a higher cetane value than biodiesel and can be produced with better cold weather properties as a result of the secondary isomerization step in the process. Isomerization is also known as dewaxing. This step is often, but not always, included in HVO process designs.

The fuel is sulphur free and has very low aromatics content (although this can cause issues with some engine seals).

The market price of HVO is higher than that of biodiesel and of petroleum diesel.

Environment

The process has a low ratio of energy consumed to fuel produced (~0.02) excluding the hydrogen but drops to about 0.15 when the hydrogen energy is considered.

The hydrotreating catalysts have a long life and can be re-generated. There is limited waste generated as part of the process, although this can depend on the feedstock and the need to pre-treat the feedstock to remove minor contaminants.

Research and Development Perspective

Hydrogenated vegetable oil (HVO) is a Category 4 technology, a commercial technology with large scale deployment. While there are fewer HVO facilities in the world compared to FAME facilities, the basic processing technology have been practiced for many years in petroleum refineries. The price and performance of the technology today is well known and normally only incremental improvements would be expected. Therefore, the future price and performance may also be projected with a relatively high level of certainty.

Examples of Market Standard Technology

The leading commercial supplier of the technology is Honeywell UOP. Their corporate office is:

UOP World Corporate Offices
25 E. Algonquin Road (Bldg A)
P.O. Box 5017
Des Plaines, IL 60017-5017
Toll Free: +1 (800) 877-6184
Phone: +1 (847) 391-2000

They have a European office at:

Noorderlaan 147
B-2030 Antwerp, Belgium
Phone: +32-3-540-9911

Haldor Topsøe offers their HydroFlex™ technology. Topsøe has been researching and developing innovative solutions for renewable fuel production since 2004. They claim that many licensed HydroFlex™ units are in full operation today.

Haldor Topsøe A/S
Haldor Topsøes Allé 1
DK-2800
Kgs. Lyngby
Denmark

Neste and REG have developed their own technology and have not sold it to any other producers at this time.

Predication of Performance and Cost

The production of HVO from triglycerides is a commercial technology that is deployed at scale. Only small incremental improvements on performance can be expected in the future. The current price and performance information presented has a relatively high level of certainty, although with a limited number of participants in the sector much of the information is considered confidential by the producers.

Future capital costs can be expected to improve as the plant size increases, however the future inputs and outputs are expected to be very stable due to the high levels of performance of the existing facilities. These assumptions are confirmed by the work of de Jong [4] who determined that the technology had a high capital growth factor indication a small difference between the capital costs of the pioneering plants and the nth plant.

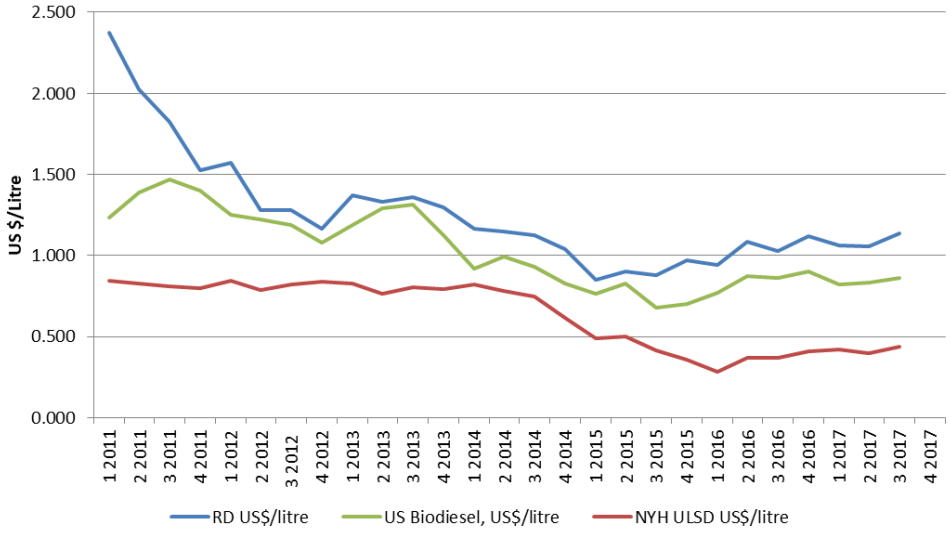
Uncertainty

There are relatively low levels of uncertainty with respect to this pathway as it is a commercially available process. However, only a limited number of companies throughout the world offer this technology. With exception of Neste, these companies provide little information on their operations.

Additional Remarks

As noted above Neste does provide some information on the performance of their renewable products division as part of their quarterly financial reporting [5]. The reported HVO revenue is their world-wide average value. The following figure is developed from those reports and benchmark price data on diesel fuel and biodiesel reported by the US Energy Information Administration [6] and Iowa State University [7].

Figure 3: HVO Selling Prices vs Biodiesel and Fossil Diesel

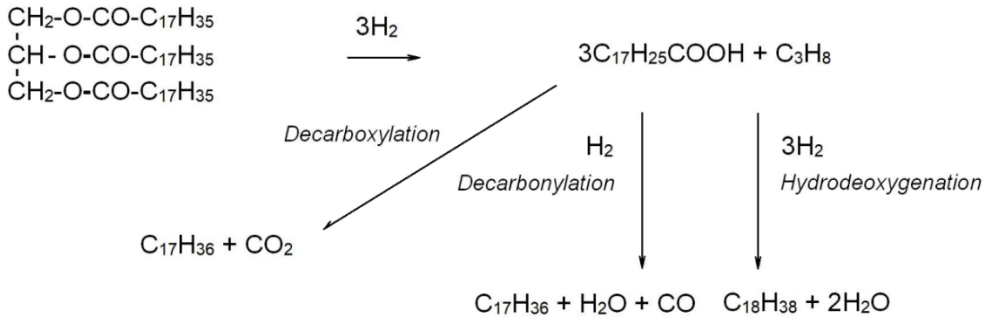


RD, biodiesel and ULSD represents selling prices for HVO, FAME biodiesel and ultra low sulphur diesel, respectively.

Quantitative Description

The quantitative data is presented below for the technology. The HVO chemistry is shown in the following figure [8].

Figure 4: HVO Chemistry



The oxygen from the triglyceride feedstock is removed as CO, CO_2 and some as H_2O. Unlike an ethanol plant the CO_2 is not available as a concentrated co-product stream. Some of the feedstock is also converted to propane and because there is some cracking of the chains a portion of the liquid product is in the jet fuel or gasoline pool boiling range.

Typical Plant Size

There is almost an order of magnitude difference between the output of the smallest plant and the largest plants. It is assumed that the typical plant size will be in the middle of the range at 500,000 tonne per year of primary product (640 million litres per year).

Input and Output

The input and output data shown in the following table is the same information used for the energy balance shown in Figure 2. It is typical data and not representative of any specific plant. The output is one litre of HVO.

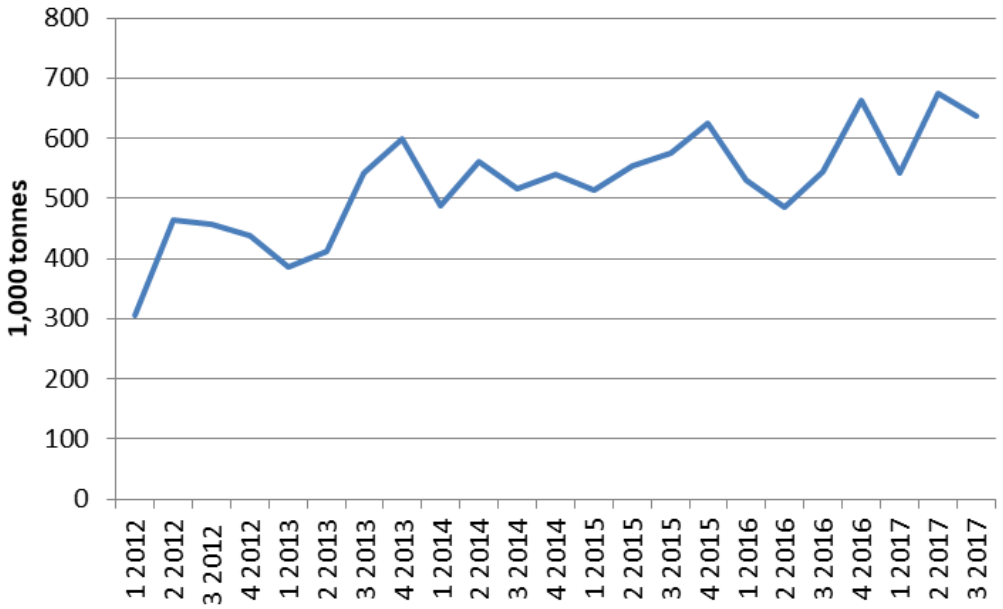
Table 1: Mass and Energy Flows HVO

	Inputs	Outputs
Feedstock	0.97 kg	
Hydrogen	0.035 kg	
Natural gas	0.27 MJ	
Electricity	0.085 kWh	
Fuel gas		0.03 kg
LPG		0.025 kg
Naphtha		0.06 kg

Forced and Planned Outage

Neste report their quarterly sale rates. Since the last quarter of 2011, the plants in Finland, Singapore, and Rotterdam have been operational with an original nameplate capacity of 500,000 tonne per quarter. Neste now claims a quarterly production capacity of 650,000 tonnes. The sales performance is shown in the following figure and it is assumed that the production rates closely follow the sales rates. Since 2013 the plants have operated above the original nameplate capacity indicating minimal unplanned outages.

Figure 5: Quarterly Performance – Neste HVO



Neste report that the plants undergo a major turnaround lasting 8 weeks every four to five years. This is equivalent to 15 days per year.

Technical Lifetime

Plant lifetimes should exceed 25 years, similar to what would be expected in a petroleum refinery.

Construction Time

Construction time of 24 to 30 months should be expected. This is longer than a FAME plant but the HVO plants are generally larger and have process vessels that operate at higher pressures and temperatures with an expectation of longer fabrication periods.

Financial Data

Financial data from public data sources is presented below. Where primary public data is not available two peer reviewed techno-economic analyses have been used [9] [10].

Investment Costs

Diamond Green Diesel is undergoing a 330,000 tonne/year expansion in Louisiana. The reported cost is $190 million (US) [10]. They reported that this was about 50% of the cost of a greenfield plant.

Converting this cost to Europe, the location factor, exchange rate, and greenfield factor is applied as shown below.

$$US\ capital\ cost \cdot greenfield\ factor \cdot currency\ factor \cdot location\ factor$$

$$\$190{,}000{,}000 \cdot 2.0 \cdot 0.85 \cdot 1.21 = €390\ \text{million}$$

It is estimated that a 500,000 tonne per year plant would have a capital cost of €550 million using a scaling factor of 0.8. This corresponds to €0.86/litre.

The EU Sub Group on Advanced Biofuels [12] reported that the capital costs for a 500,000 tonne/year plants would be between 192 and 577 million Euros. The Diamond Green data would be at the upper end of the range.

A one million tonne per year plant is expected to cost 20.6 times (1.5) that of the 500,000 tonne per year plant or 0.65 €/litre. This value is used for the 2050 capital cost. The same approach is used for the 750,000 tonne plant in 2030.

Operating and Maintenance Costs

Neste publishes some information on their production costs in their quarterly financial reports. The available information for 2016 is shown in the following table.

Table 2: HVO Operating Costs

Item	Value	
	€/tonne	€/litre
Revenue,	1,211	0.93
Variable production costs	117	0.09
Fixed Costs	58	0.04
Depreciation	50	0.04
EBIT	21	0.02
Implied Feedstock costs	876	0.67

Operating costs are not expected to change over time due to the high efficiency level of the current technology. The variable operating costs are reduced by 0.01 €/litre to account for the energy costs that would be included in the reported variable costs.

Start-up Costs

The start-up costs are included in the Neste financial information.

Technology Specific Data

The physical properties of HVO are shown in the following table.

Table 3: HVO Properties

Property	Value
Density, kg/m^3	770
LHV, MJ/kg	44.1
LHV, MJ/litre	34.4
Oxygen content	0
Cetane number	>70
Cloud point, C	As low as -40 °C[15]

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105 First Generation Ethanol

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Publication date

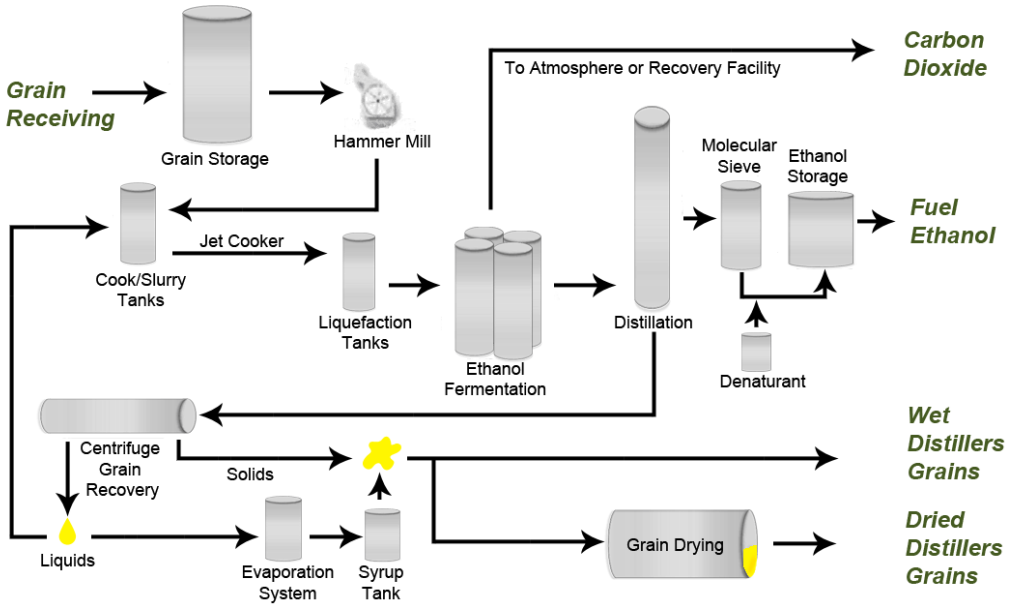
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Amendments after publication date

Date	Ref.	Description
12 2018		Datasheet revised

Qualitative Description

Figure 1: 1st Generation Ethanol Production Process



Brief Technology Description

Cereal crops, with corn and wheat being the most common, are milled to reduce the particle size. The milled grain is mixed with water and a liquefying enzyme and heated to 90 to 100°C to liquefy the starch in the grain. The slurry is cooled to 30 to 35°C, a saccharification enzyme to convert the starch to fermentable sugars and yeast is added to ferment the sugars to ethanol and carbon dioxide. After about 48 hours the "beer" is sent to distillation columns to separate the ethanol and some of the water from the remaining solids. The ethanol stream has the remaining water removed in a molecular sieve or an anhydrous distillation column. The ethanol is suitable for fuel blending at this stage. The distillation bottom columns are centrifuge to separate some of the solids from the liquid. The solids can be sold or sent to a dryer. The liquid can have corn oil removed and is then concentrated through a vacuum evaporation system to produce a syrup which can also be sold or sent to the dryer. The distillers' grains are used as a high protein animal feed.

Input

The primary inputs into the system are the corn or wheat feedstock, water, power and thermal energy. There are secondary inputs of enzymes, yeast, and acids and bases for pH adjustment.

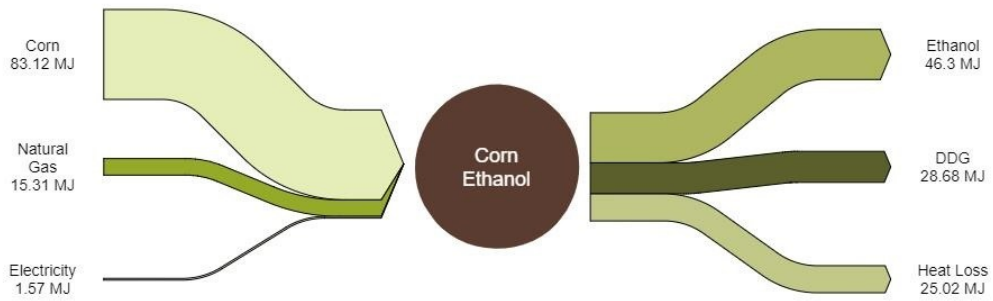
Output

The primary outputs are ethanol and distillers' grains. The distillers' grains can be sold wet or dry, and either combined (where the solids and syrup are combined) or separately (where the solids and syrup streams are sold as individual products. Corn plants can also separate a small volume of corn oil which can be used for animal feed or as a FAME feedstock. The corn oil can have a much higher value when sold separately instead of as part of the distillers' grains. The oil content of wheat is too low make the extraction practical.

Energy Balance

The energy balance of a 1st generation corn ethanol plant is shown in the following figure. This figure is based on an ethanol yield of 415 litres per tonne of corn. The ethanol yield is a function of the starch content of the grain and the plant conversion efficiencies, the typical range is 405 to 430 l/tonne [1]. The ratio of the energy of ethanol to Distillers' Dried Grains DDG will vary with the starch content of the feedstock. Wheat generally has a slightly lower ethanol output but higher DDG output. The natural gas and power requirements are typical values for a corn ethanol plant that dries all of the distillers' grains. Wheat plants will use more energy than corn plants as more DDG is produced that must be dried and the evaporation systems are generally not as efficient as the viscosity of the syrup is higher for a given solids content.

Figure 2: 1st Generation Ethanol Energy Balance



Most of the unit operations in the process are undertaken at relatively low temperatures so the heat losses are low quality. Most of the heat is lost either through a cooling tower or in the exhaust stack of the distillers' grain dryer. A few plants have started to recover some of the latent heat of vapourization in the dryer exhaust and use the energy to reduce the purchased natural gas. The practice is not yet widespread.

Typical Capacities

In North America, the most recent new plants were built with 400 million litres of ethanol production capacity per year (320,000 tonnes ethanol per year, 280 MW). These plants generally had two production trains of 200 million litres per year each (160,000 tonnes ethanol per year).

Twenty-five years ago, plants were typically 60 million litres per year. Over time capacities grew to take advantage of economies of scale.

Regulation Ability

Plants generally operate continuously with one or two shutdowns per year for maintenance. The process is biological and it is important to minimize the possibility of infections that can reduce yield and negatively impact the performance of equipment. This limits the potential for operations below about 80% of nameplate capacity. Lower production rates can only be achieved with more frequent start-up and shutdown cycles which increase the operating costs.

Space Requirements

Pannonia Ethanol in Hungary is a 430 million litre per year corn ethanol facility designed and built by the American company Fagen. It covers 22 ha on the banks of the Danube River. It was a greenfield plant.

The Vivergo Fuels 420 million litre per year wheat ethanol plant in the UK occupies 10 ha in an existing industrial complex. The site was chosen due to its existing infrastructure and utilities.

Space requirements will range from 0.024 to 0.052 ha/million litres (350 to 775 m^2/M. Smaller plants may require more land per million litres than larger plants. The degree of integration with other infrastructure will also impact the land requirements.

Advantages/Disadvantages

1st generation ethanol is a well proven technology. It is the largest volume renewable fuel produced in the world today. Ethanol is a high-octane fuel that can be used to increase the octane of the gasoline that it is blended with. This has the potential to reduce the GHG emissions of the petroleum refinery and increase the refinery energy efficiency, factors that are not often considered when the benefits of ethanol are discussed.

Ethanol does contain about 35% oxygen which limits the quantity that can be blended with gasoline. In Denmark, ethanol is blended at the 5% by volume level [2]. Europe, ethanol blended at the 10% by volume level is a relatively recent development. E10 is currently available in Belgium, Finland, France and Germany [3].

Environment

The lifecycle GHG emissions of 1st generation ethanol is about equally split between the feedstock production and fuel production [4]. Feedstock emissions are geographic specific and can vary significantly from country to country due to production practices.

Research and Development Perspective

First generation ethanol produced from corn and wheat is a category 4 technology; a commercial technology with large scale deployment. The price and performance of the technology today is well known and normally only incremental improvements over time would be expected. Therefore, the future price and performance may also be projected with a relatively high level of certainty.

Examples of Market Standard Technology

There are many suppliers of 1st generation ethanol technology. More than 50% of the US production capacity was designed by ICM Inc. They have also designed plants in Europe, Brazil, Argentina, and Africa.

ICM, Inc.
P.O. Box 397

310 N. First Street
Colwich, KS 67030-0397

The German company GEA is one of the largest suppliers of process technology for the food industry and a wide range of other industries, including fuel ethanol. They also have expertise with respect to wheat starch production.

GEA Group Aktiengesellschaft

Peter-Müller-Str. 12

40468 Düsseldorf

Germany

Predication of Performance and Cost

The corn ethanol industry has been studied from the perspective of the experience curve [5] [6]. There have been significant improvements in the performance of the technology over time and the improvements are in line with the expectations based on the production growth of the industry.

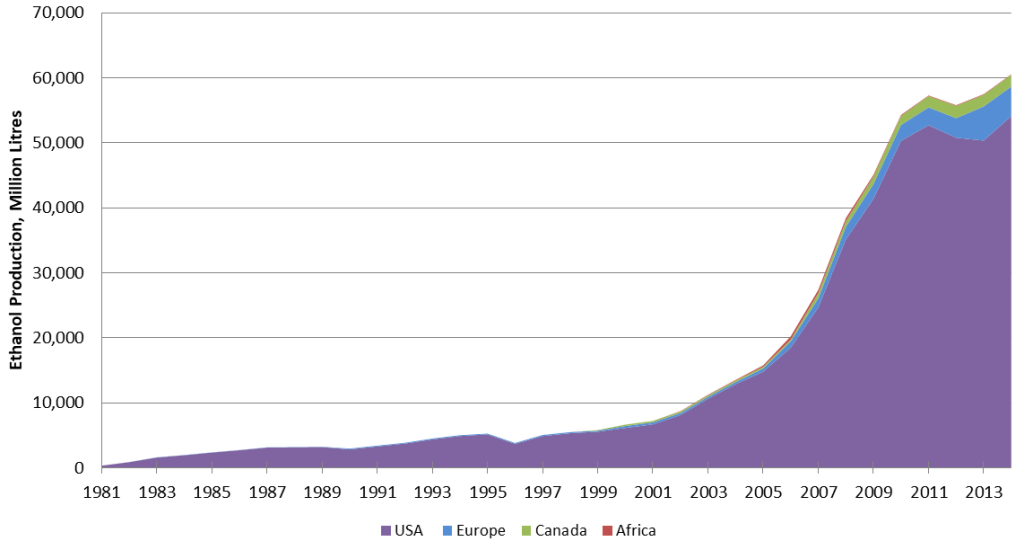
Uncertainty

Given the level of maturity of the industry there is a low level of uncertainty associated with information collected on the technology.

Additional Remarks

The world starch ethanol production is shown in the following figure [6]. The United States is the dominant producer and that is almost all produced from corn. Production in Canada and Europe is a mixture of wheat and corn plants.

Figure 3: World Starch Ethanol Production



Production levels in the post 2010 period have not increased as rapidly as they did in the previous decade, as a result the rate of change in the technology has slowed down on a year over year basis since it takes longer to double the production in this decade than it did in the last decade.

Quantitative Description

There is a significant amount of quantitative information available for this technology since it is the largest renewable pathway in the world.

The process involves two chemical reactions. In the first, starch is hydrolyzed to glucose as shown below.

$$(C_6H_{10}O_5)_n + nH_2O \rightarrow nC_6H_{12}O_6$$

Starch Water Glucose

The glucose is then fermented with yeast to ethanol and carbon dioxide.

$$C_6H_{12}O_6 \rightarrow 2C_2H_5OH + 2CO_2$$

Glucose Ethanol Carbon Dioxide

Typical Plant Size

Plant sizes can range up to 400 million litres per year (300 MW), although plants of this size often have two production trains. The dryers for the plants are often the limiting size of the production train, although the size of individual fermenters can also be a limiting factor. Most plants operate with four fermenters and a 48 to 60-hour fermentation cycle so fermenters can be quite large. The average plant size in Europe is smaller than this.

In their 2014 report on the industry, ePure [8] reported the production capacity of beverage and fuel ethanol plants and the number of plants by country. The fuel ethanol production capacity was reported to be 7 billion litres. The fuel ethanol industry apparently operated at 65% of capacity in 2014. That information along with the fuel ethanol production reported by Eurostat for 2014 is shown in the following table. Not all of these plants are currently operating.

Table 1: European Ethanol Plants

Country	Production Capacity (million litres)	No. Plants	Average Size	2014 Fuel Production
			Million litres	
France	2300	19	121	975
Germany	1400	12	117	897
United Kingdom	900	5	180	519
Poland	750	14	54	181
Spain	600	5	120	486
The Netherlands	575	2	288	0
Hungary	520	3	173	372

Belgium	500	3	167	325
Czech Republic	350	6	58	132
Italy	300	6	50	1
Sweden	275	6	46	175
Austria	250	2	125	262
Slovakia	240	1	240	134
Romania	200	3	67	15
Lithuania	100	2	50	13
Latvia	50	2	25	0
Bulgaria	50	2	25	27
Finland	50	4	13	24
Ireland	40	1	40	0
Denmark	30	1	30	0
Total	9480	99	100	4,538

ePure reported that 31% of the feedstock in 2016 was corn, 32% was wheat and the remainder was sugar based and other cereals [9]. Ethanol production from corn in 2016 was 1.61 billion litres and from wheat was 1.66 billion litres.

Input and Output

The primary input is the grain feedstock and the electric power and thermal energy. The primary output is the ethanol, the DDG for animal feed, and some plants capture the CO2 from the fermentation vessels for use in the industrial gas market. The inputs and outputs are shown in the following table.

Table 2: Corn Ethanol Inputs and Outputs

	Inputs	Outputs
Feedstock	2.4 kg	
Power	0.20 kWh	
Natural gas	7 MJ	
Ethanol		1 litre
DDG		0.72 kg

CO_2 (if captured)		0.75 kg
Corn Oil		0.03 kg

The inputs and outputs for a wheat plant are slightly different since wheat usually has a lower starch content than corn. Feedstock consumed may be 2.6 kg/litre of ethanol and the DDG production may be 0.94 kg/litre. The energy consumption may be 9 to 10 MJ of natural gas per litre of ethanol. There can be variations in the feedstock starch contents from year to year and since wheat is a winter crop and corn is planted in the spring for a fall harvest, the ratio of the starch contents of corn and wheat can vary from year to year.

Future improvements can be expected in ethanol yield and energy use. In the US the ethanol yield has been increasing at about 0.2% per year. This is expected to continue. Natural gas energy use has been declining by 1.8% per year. This rate is expected to start to decline as the time to double production increases. It has been assumed that gas consumption declines by 20% over the next 30 years. Electric power consumption has shown little improvement in the past decade, which is not that surprising as the power consumption tends to be more a function of the original design rather than operating practices. No change in power production is forecast.

Forced and Planned Outage

Ethanol plants are generally designed to operate 350 days per year with 15 days for maintenance. Some plants take a maintenance break once per year and others will take two shorter breaks per year.

Since the process is a generally a combination of batch fermentation with intermediate storage tanks at the front and back ends of the plant, individual components can often be taken offline for cleaning or maintenance without impacting production rates. The industry generally has very high production utilization if the markets for the products are available. The European industry generally operates at lower rates of capacity utilization because of the smaller and less mature market than in North America.

Technical Lifetime

Plant lifetimes should exceed 25 years, as the US and EU starch ethanol industries have demonstrated.

Construction Time

Plants can generally be constructed in 18 to 24 months. There are no very high pressures or temperatures involved in the process which reduces the required lead time for process components.

Financial Data

There is more financial data available for the North America industry than there is for the European industry. Other than feedstock costs and selling prices the other categories of financial costs should be relatively similar for the two industries.

As with the FAME industry, Iowa State University maintains a financial model and monthly operating data for a 380 million litre corn ethanol plant located in Iowa [10].

Investment Costs

There have been relatively few 1st generation ethanol plants built since 2010. So, the capital cost data is not as current as it used to be.

The Iowa State model has capital costs of $181 million dollars excluding working capital. There are two corn ethanol plants that are under construction in the United States [11] [12]. One will produce 300 million litres and has a reported cost of $150 million [12]. This included $10 million in working capital. The other has a capacity of 450 million litres and has a cost of $190 million [12].

Table 3: Capital Costs

	Iowa State	Ringneck Energy	Elite Octane
Capacity, million lpy	380	300	450
Capital cost, $ Million	181	140	190
$/annual litre	0.48	0.47	0.42

The three cost estimates are quite close. Converting this data to a European value using the methodology provided in the Introduction, the capital cost for this large plant is estimated as follows:

$$\$0.50/\text{litre} \cdot 0.85\ €/\$ \cdot 1.21\ \text{Europe factor} = 0.51\ €/\text{litre}$$

Given the slow pace of new plant construction it is expected that the future capital costs are a function of the plant scale and a 0.8 scaling factor and not of technology improvements.

Operating and Maintenance Costs

The operating and maintenance costs in the Iowa State model are used as the basis for the data table. The 2016 data has been used and the power and natural gas costs have been adjusted to the Danish values reported in section 1.4. The capital cost has been adjusted to 206 million Euro to match the value above. The results for 2016 are shown in the following table.

Table 4: Production Costs 1st Generation Ethanol

	Iowa State
Natural gas	0.043
Power	0.012
Other Variable	0.049
Fixed Costs	0.048
Total Costs ex feedstock	152

Fixed operating costs are not expected to change over time but the variable costs decrease as the energy use improves.

Start-up Costs

Start-up costs are included in the previous table.

Technology Specific Data

The properties of ethanol are shown in the following table. The ethanol molecule contains about 35% oxygen which reduces the volumetric energy density of the fuel.

Table 5: Ethanol Properties

Property	Value
Density, kg/m^3	790
LHV, MJ/kg	26.9
LHV, MJ/litre	21.3
Oxygen content	35%
Blending Octane number	~115
Flash point, C	13

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106 Cellulosic Ethanol

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Amendments after publication date

Date	Ref.	Description
12 2018		Datasheet revised

Qualitative Description

Cellulose is an important structural component of the primary cell wall of green plants. It is an organic compound with the formula $(C_6H_{10}O_5)_n$, a polysaccharide consisting of a linear chain of several hundred to many thousands of β(1→4) linked D-glucose units. The basic chemical formula is the same as for starch with the difference being that the "n" is much larger for cellulose.

If the cellulose can be broken down into the glucose molecules, then those can be fermented into ethanol and the rest of the process is similar to the 1st generation starch ethanol process.

Brief Technology Description

The cellulosic feedstock, which could be straw, corn stover, bagasse, or wood residue is generally subjected to a pretreatment stage to separate the cellulosic portion of the material from the lignin and make the feedstock amenable to hydrolysis. The pretreatment stage may use acid and or steam to accomplish the goal.

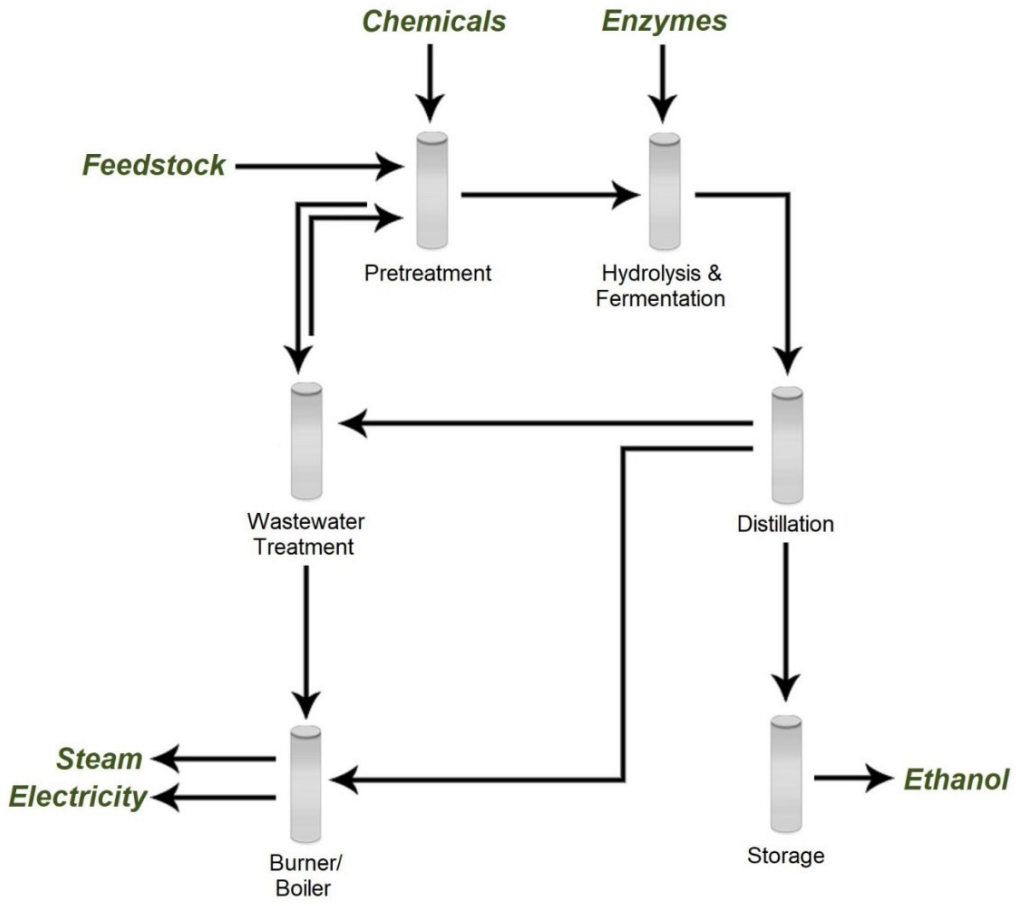
The hydrolysis stage breaks down the cellulose molecules into smaller polysaccharides or completely into glucose units; this is a hydrolysis reaction. Hydrolysis can be carried out with cellulase enzymes or with acid. Most process developers are using enzymes.

Following the hydrolysis stage, a yeast is added to convert the sugars into ethanol and carbon dioxide. The "beer" is then distilled to remove ethanol and water from the unfermented solids. A second distillation column to reach 95% ethanol and 5% water is followed by a molecular sieve to produce anhydrous ethanol by most process developers.

The unfermented solids have little value as animal feed and process developers generally convert the material to energy (thermal and electricity) to be used in the process with any excess being exported to the grid.

A typical process flow diagram is shown in the following figure.

Figure 1: Cellulosic Ethanol Process



Input

The primary input is the cellulosic feedstock, along with water and the chemicals or enzymes used in the process. The quantity and type of feedstock vary between process developers.

Output

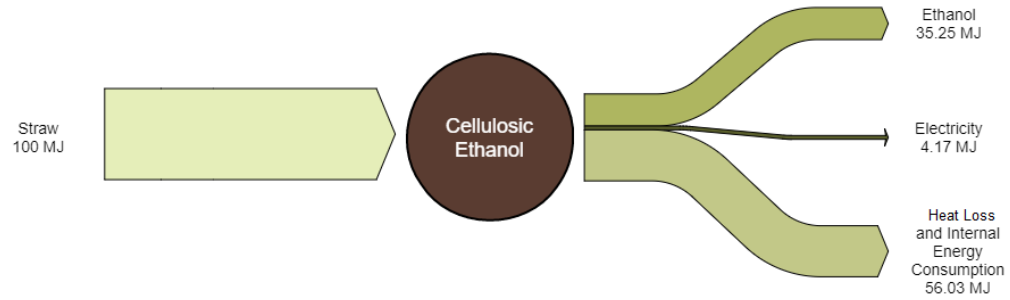
The output from the system is ethanol and in most cases some electricity. With most developers the lignin and unfermented solids are burned and/or anaerobically digested to produce energy for the process. Some integrated biorefinery designs would use vinasse (stillage) from the ethanol production for biogas production, which can be used directly for (on-site) CHP or upgraded to be sold or added to the local gas grid. Lignin and other combustibles would be used for CHP and any excess electricity and heat could be sold externally. Some developers try to sell the stillage (unfermented solids) as a fertilizer for field application. This will return the nitrogen, phosphorus and phosphates in the straw to the fields and replace the synthetic fertilizers that would be used to replace these nutrients.

Energy Balance

There are a number of technology developers working on cellulosic ethanol processes. There are different approaches to the basic system that are being employed. As this is a new technology, developers do not release too much information to the public. The most detailed techno-economic information available is from the US National Renewable Energy Laboratory [1]. This information is now dated and likely doesn't represent the current state of the art but it is still the most complete analysis available. A Sankey diagram based on

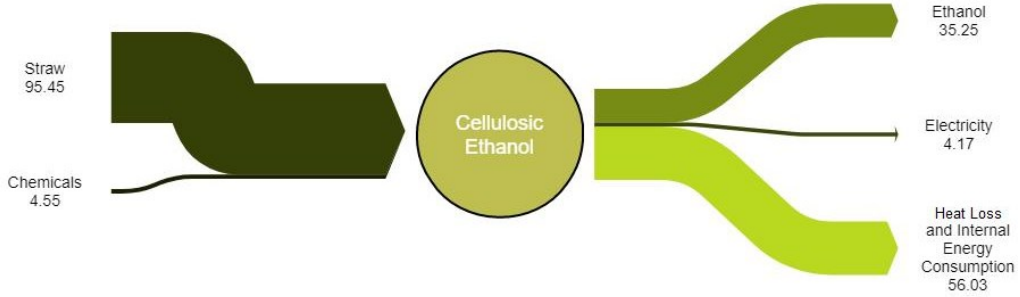
published data from the US National Renewable Energy Laboratory is shown in the following figure.

Figure 2: Cellulosic Ethanol Energy Balance



This process also consumes a significant quantity of chemicals to assist with the process and waste treatment. The Sankey diagram when these chemicals are included is shown in the following figure. The chemical energy is the energy required to produce and transport the chemicals.

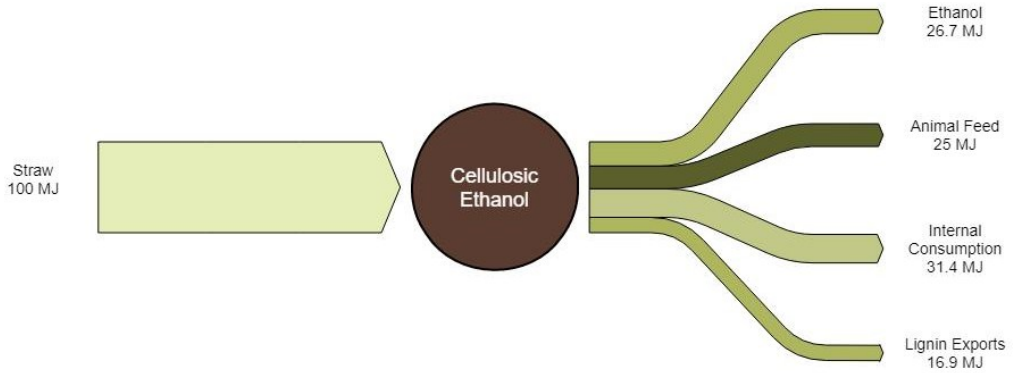
Figure 3: Cellulosic Ethanol Energy Balance with Chemicals



The process generates its own energy for use in the process. This is reflected in the high internal energy consumption shown in the figure. The process also consumes significantly more energy than the 1st generation ethanol processes.

For comparison the Sankey diagram for the Inbicon technology is shown in the following figure [2].

Figure 4: Inbicon Cellulosic Ethanol Energy Balance



This technology has much lower ethanol production but also produced animal feed and lignin for energy applications after some of the lignin was used to supply the heat and power for the facilities. The Inbicon plants were designed to be integrated with power plants, which also allows the diversion of the lignin and animal feeds streams to power generation depending on the location specific opportunities.

Typical Capacities

There have been about a half dozen commercial demonstration plants built in Europe, Brazil and the United States over the past five years. The plants have been sized in the 60 to 115 million litre per year range (45 to 80 MW), however few of the plants have been able to operate at capacity for significant periods of time but several plants are making progress towards operating at design capacity. The plants consume 50,000 to 100,000 tonnes of feedstock per year when operated at capacity.

Regulation Ability

Since most of the plants have not been able to operate continuously for extended periods of time it has been difficult to assess their regulation capabilities. In theory, the regulation capabilities should be similar to those of the 1st generation plants but the added complexity of the process and the fact that the process supplies the fuel for the process may limit the regulation capacity.

Space Requirements

The 50 million litre per year Beta Renewables Plant in Italy covers an area of 14 ha including some feedstock storage. The 110 million litre per year DuPont facility in Nevada Iowa has an area of 18.5 ha, excluding the storage area. The storage area appears to be twice the size of the plant. The 80 million litre per year POET DSM covers an area of 120 ha including feedstock storage.

The space requirements for the plant do vary significantly, some of which is due to the need to store a year or more of feedstock on the site. Space requirements of 0.5 ha/million litres (7000 m2/MW) would appear to be a reasonable value which allows room for feedstock storage.

Advantages/Disadvantages

The cellulosic ethanol process utilizes a feedstock that is largely underutilized in most agricultural regions of the world, thus there is no perceived competition for feed and food crops. The GHG emissions are similar to those of sugarcane ethanol, where crop residues provide the energy for the process.

The technology is complex and capital costs will be higher than for 1st generation ethanol plants. The technology complexity has also negatively impacted the time required to commercialize the process. The POET DSM plant has only recently announced that they have been able to solve the pre-treatment challenges and have been able to run the plant at 80% of capacity [3]. This is three years after the plant opened.

Other large scale first of kind plants have also had commercial and technical challenges, for example the DuPont plant is now for sale, the Beta Renewables plant in Italy has been idled and the Abengoa plant in Kansas was sold in a bankruptcy sale. The exact reasons for these decisions are not public and in some cases the decisions may have been due to other challenges that the companies may have faced or changes in business priorities. Operations at some of these plants may resume with new owners.

POET DSM [4] reported that in addition to resolving the technical challenges of pretreatment that they have decided to build an on-site enzyme production facility.

Presumably this is a commercial decision; on-site production would not need to concentrate the enzyme to reduce transportation costs, saving capital and operating costs.

Van der Meij [5] recently presented a venture capitalist's view of the technology and stated that he could not see the business case for the technology citing feedstock logistics costs, scale, high pretreatment costs and margin uncertainty as issues. He also stated that there was still significant technology risk present. Nevertheless, new plants are being announced [9] [10].

Environment

In some regions of the world there is concern that removing too much straw from the field will have negative impacts on soil health. Reduced soil carbon and increased erosion potential are potential outcomes if the crop residue harvest is not undertaken in a sustainable manner.

Research and Development Perspective

Cellulosic ethanol produced from straw is a category 2 technology, a pioneer phase technology with limited applications to date. The technology has been proven to work through demonstration facilities or semi-commercial plants. However, due to the limited application, the price and performance is still attached with high uncertainty, since development and customization is still needed. The technology still has a significant development potential.

This technology has probably had more research undertaken on it than all of the other technologies in this report. As the technology has been scaled to near commercial scale many of the process developers have experienced significant challenges with materials handling through the front end of the process.

The challenges of operating the first of a kind cellulosic ethanol facility were acknowledged by Merritt [6]. He stated that three years after start-up the plant is operating at 80% of capacity, converting 95% of the fermentable sugars to ethanol at the design solids loading. The engineering focus is now on how to simplify the process and reduce capital costs, exactly the activities that create the learning curve. He also highlights advances in enzyme effectiveness in reducing future operating costs.

Examples of Market Standard Technology

The cellulosic ethanol technology has recently suffered some setbacks with respect to the commercialization of the technology.

One of the early movers on the commercialization of cellulosic ethanol technology was the Danish company Inbicon. However, they have had difficulty in developing a commercial project without government guarantees.

The Italian company, Gruppo Mossi Ghisolfi, which developed the Beta Renewables plant in Crescentino Italy and supplied the technology for the GranBio plant in Brazil, has recently been granted protection from creditors [7].

DuPont has recently put their Nevada Iowa plant up for sale [8].

The one company that has recently moved ahead with new projects is Clariant. They have announced a license agreement with Enviral [9] to add a 50 million litre per year plant to an existing facility in Slovakia and a similar sized plant to be built in Romania [10].

Clariant Produkte Deutschland GmbH

Biotech & Renewables Center

Staffelseestrasse 6

D-81477 Munich, Germany

Predication of Performance and Cost

This technology is at the early stages of commercialization. There are some reported capital costs for the first commercial scale plants (pioneer plants) but there is no published, independently verified information on the actual performance of the plants.

It is expected that capital costs and operating costs will improve significantly once the existing plants achieve consistent operations and new plants are built based on the learnings from operating the first commercial plants.

Uncertainty

There remains significant uncertainty with respect to the commercialization of the technology. The plants built 3 to 5 years ago have experienced some unforeseen technical and economic challenges which has slowed the development progress.

Additional Remarks

The significant cellulosic plants that have been built are shown in the following table [11].

Table 1: Cellulosic Ethanol Plants

Name	Annual Capacity	Feedstock	Comment
Abengoa Bioenergy Biomass of Kansas	90 million litres	corn stover, wheat straw, switch grass	Idle
Beta Renewables	50 million litres	wheat straw, rice straw, arundo donax, poplar	Idle
DuPont	110 million litres	Corn Stover	Idle
GranBio	80 million litres	Sugarcane bagasse and straw	Operational
POET-DSM	80 million litres	Corn Stover and corn cobs	Operational
Raizen Energia	40 million litres	Bagasse	Operational

Quantitative Description

The available quantitative data that is available on the technology is mostly from third parties and not from the technology providers or plant operators. Actual plant data is considered confidential by the process developers.

The process itself is essentially identical to the starch ethanol process with the primary difference being the length of glucose polymer chain.

The process involves two chemical reactions. In the first, cellulose is hydrolyzed to glucose as shown below. One of the technical challenges with the process is making the cellulose component of the lignocellulosic feedstock accessible for the hydrolysis process.

$$(C_6H_{10}O_5)_n + nH_2O \rightarrow nC_6H_{12}O_6$$

Cellulose Water Glucose

The glucose is then fermented with yeast to ethanol and carbon dioxide.

$$C_6H_{12}O_6 \rightarrow 2C_2H_5OH + 2CO_2$$

Glucose Ethanol Carbon Dioxide

Typical Plant Size

The plants that have been built have been in the 50 to 100 million litres per year range. Some of the techno economic studies that have been done use plant sizes of 200 million litre per year (140 MW) [1]. As with the 1st generation plants it is likely that plant sizes will increase as more plants are built and experience is gained with the technology.

Input and Output

The primary input is the feedstock. Most technology developers (but not all) utilize the non fermentables to provide the thermal energy and electric power for the plant and some have excess power available for sale.

The plants also consume enzymes, yeast, and a number of process chemicals. Enzymes can be purchased or produced on site using a portion of the sugar produced by hydrolysis. Enzyme consumption is much higher than it is in a 1st generation ethanol plant. In the 2012 state of the art case [1] reported the consumption of 0.38 kg of chemicals for every kilogram of ethanol produced. It is likely that this has been decreased with additional development work but NREL have not published an updated state of the art report.

The plants produce ethanol and in some cases electricity. Some developers propose to sell a fertilizer product instead of producing electricity. The data used for the Sankey diagram is shown in the following table [1].

Table 2: Cellulosic Ethanol Inputs and Outputs

Parameter	Input [kg]	Output
Feedstock	3.38	
Chemicals	0.38	
Sulfuric acid	0.072	
Sodium hydroxide	0.082	
Ammonia	0.042	
Corn steep liquor	0.048	
Diammonium phosphate	0.005	
Sorbitol	0.002	
Glucose	0.103	
Host nutrients	0.002	

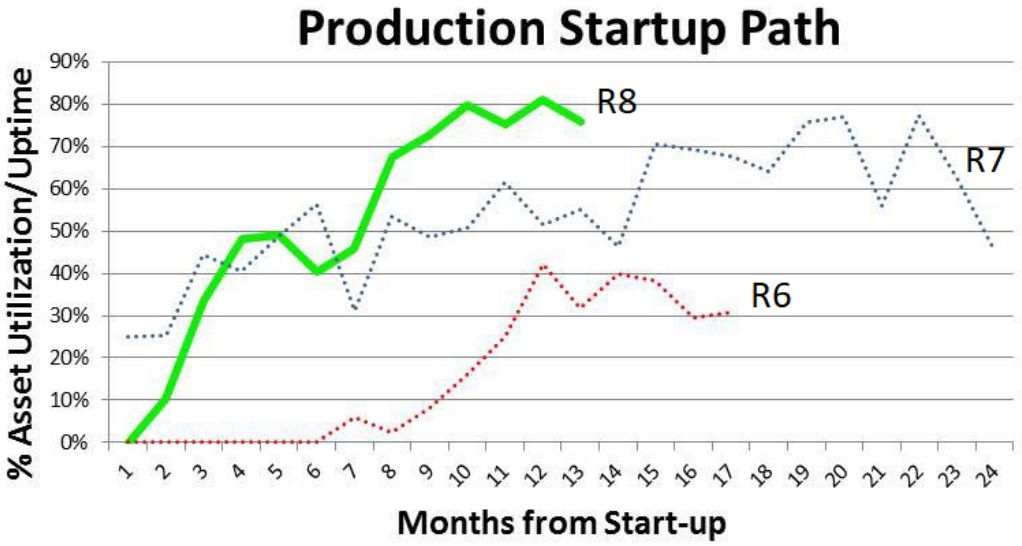
Sulfur dioxide	0.001	
Ethanol		1 litre
Power		0.70 kWh

The current performance is not likely at this level. It has been assumed that the current yield is 80% of the value shown, the value can be achieved in 2030 and the 2050 value is 10% better.

Forced and Planned Outage

The existing operating plants have struggled with forced outages. The following figure shows the production rate versus time for three generations of Iogen's cellulosic technology [12]. The figure shows the difficulty in achieving design rates but also the progress that can be made with experience.

Figure 5: Cellulosic Ethanol Production Rates. The legends refer to different production lines and R8 is at the latest stage of development.



Both Iogen and Poet have stated that the initial operational challenges have been in the pre-treatment stage of the process. Severe fouling and flow restrictions have been reported in this section. Eventually it is expected that plants will be able to operate for 330 to 350 days per year with no forced downtime.

Technical Lifetime

Plant lifetimes should exceed 25 years, as the US and EU starch ethanol industries have demonstrated.

Construction Time

Due to the added complexity of the process construction times are expected to be 24 to 30 months.

Financial Data

There is very limited financial data available on the process. The available information is a combination of techno-economic studies and the published capital costs of the commercial plants.

Investment Costs

Lynd et al (2017) reported the size and capital expenditures for the pioneering plants [13]. The information is shown in the following table.

Table 3: Capital Costs

Company	Location	Feedstock	Capacity, million litres	Capital Expenditure, Million US$	Capital, $/ litre
Abengoa	Kansas, USA	Straw	95	444.6	4.7
Beta Renewables	Crescentino, Italy	Straw	51	171	3.4
DuPont	Iowa, USA	Corn stover	114	500	4.4
Granbio	Alagoas, Brazil	Bagasse, straw	82	265	3.2
POET/DSM	Iowa, USA	Corn stover	76	275	3.6
Raizen	Piracicaba, Brazil	Bagasse, straw	40	102	2.5
Average			76	293	3.9

The capital costs for the pioneering plants have been about six to seven times higher than 1st generation plants. While de Jong [14] did not report a capital growth factor for this technology, the factors for other biomass feedstock technologies was in the range of 0.4 suggesting that the nth plants might have capital costs of $1.60/litre.

The NREL capital cost for the nth plant [1] was $381 million for a 207 million litre plant ($1.84/litre), in line with the capital growth factor adjusted costs of the pioneering facilities.

Irena [15] reported capital costs on the range of 2,000 to 3,500 US$/kW ethanol (1.5 to 2.6 US $/annual litre). They see this being reduced to 1,200 to 2,300 US $/kW ethanol by 2045.

The EU Sub Group on Advanced Biofuels [16] reported capital costs 3,300 €/kW but noted that the first plants ranged from 2,380 to over 6,700 €/kW.

The capital costs are 3 to 5 times higher than 1st generation plants and vary widely between plants. The current estimated range is €1.5 to €2.5/litre (€3,300 to 5,600/KW).

Operating and Maintenance Costs

The NREL operating and maintenance costs are summarized in the following table [1].

Table 4: Operating and Maintenance Costs

Parameter	Euro/litre
Variable operating costs, ex feedstock	0.083
Fixed costs	0.04

These are nth plant estimates which are unlikely to be achieved for the pioneering plants although there is no published information on the performance of the pioneering plants.

Start-up Costs

Start-up costs are included in the operating costs.

Technology Specific Data

The ethanol properties of cellulosic ethanol are identical to the 1st generation ethanol properties.

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107 Upgrading of Biocrudes

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Qualitative description

Brief technological description
Direct thermochemical liquefaction technologies, such as fast pyrolysis, catalytic pyrolysis, and hydrothermal liquefaction (HTL), convert biomass into biocrudes that require further upgrading to become drop-in fuels or direct substitutes for fossil fuels[16]. The extent of upgrading depends on the desired end use, as fuel specifications vary across applications. Biocrudes are typically viscous, high-boiling-point liquids with a polar nature caused by their high oxygen and nitrogen content. Their composition depends on both biomass feedstock and the specific liquefaction route. For transportation fuels, the removal of oxygen and extensive upgrading are essential to meet quality standards.

While Biocrudes have been used at a commercial level for heating applications, where the fuel quality requirements are less stringent [1], they are not yet widely adopted for transport fuels. In upgrading efforts, fast pyrolysis oil (FP bio-oil) derived from woody biomass has received particular attention due to its relatively consistent composition and lower levels of contaminants.

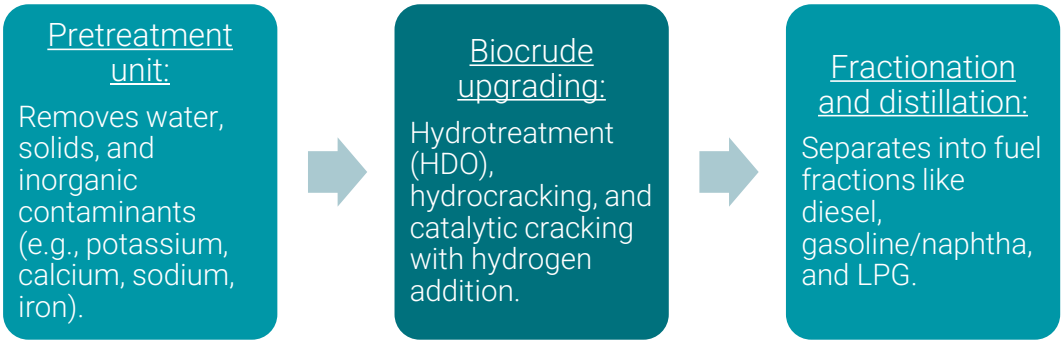
The upgrading process remains under development and faces several challenges. These challenges include the variability and complexity of biocrude composition, high oxygen and water content, acidity, and chemical instability. Such properties contribute to catalyst inhibition, increased risk of coking, and difficulties in achieving stable and efficient conversion. Additionally, the heterogeneity of feedstocks complicates process optimisation and standardisation across different upgrading technologies. [1]

Co-processing of biocrudes in existing petroleum refineries is an intermediate solution that allows utilisation of existing refinery infrastructure. Successful tests have been conducted with blend ratios between 1–5% in fluid catalytic cracking (FCC) units. In such cases, a limited amount of remaining oxygen in the product can be tolerated while maintaining fuel quality. However, this approach dilutes the biogenic fraction, meaning final products do not qualify as fully renewable.

To produce fully renewable fuels exclusively from biocrudes, more advanced processing is required. This involves both pretreatment and upgrading steps. Pretreatment is essential to remove impurities, reduce oxygen and water content, and stabilise the oil before it enters downstream upgrading units. These steps help mitigate challenges such as acidity, instability, and the presence of reactive compounds that can inhibit catalysts and promote coking. [1]

[16] This chapter does not include a description of HEFA (Hydroprocessed Esters and Fatty Acids) / liquid-to-biofuel processes. These technologies are already fully commercial and described in another chapter of this catalogue. The main barriers to further scale-up are not technological but rather the high cost of lipid feedstocks, their limited availability, and concerns about sustainability.

The following steps are typically required to prepare biocrudes for use as a drop-in fuel:



Pretreatment unit: Before advanced bio-oils can undergo upgrading processes, a pretreatment step is essential to ensure compatibility with downstream technologies. This initial step focuses on reducing the presence of water, solids, and inorganic contaminants that can interfere with catalysts and compromise process efficiency [1]. Elements such as potassium, calcium, sodium, and iron must be lowered to trace levels to avoid operational issues. Although several lab-scale studies have demonstrated effective removal techniques, including acid washing [2] and adsorption using materials like zeolites and resins [3], the transition to industrial-scale applications remains limited.

Biocrude upgrading: Once pretreated, advanced bio-oils can undergo upgrading to transportation fuels. Upgrading routes include hydrotreatment/processing, hydrocracking and catalytic cracking. The most common pathway is hydroprocessing, which typically involves the following steps [4]:

Hydroprocessing unit:

[1] Guard unit
    a. Removal of residual contaminants.
[2] Multi-stage hydrotreater
    a. High pressure (>150 bar), long residence time (LHSV < 0.5), and high temperatures (>375°C).
    b. Removes oxygen molecules.
    c. Removal of remaining oxygen, sulphur, and nitrogen, and optionally aromatics.
    d. Conversion of alkanes to isoalkanes.
[3] Hydrocracker
    a. Two-stage hydrocracker (pressure >200 bar and temperature >425°C).

Fractionation and distillation: The final step is fractionation and distillation, separating it into fuel fractions such as diesel, gasoline/naphtha, and LPG. Most drop-in biofuel technologies produce a liquid product that is a mixture of different types of fuels, such as naphtha, jet, and diesel, which all go through distillation processes to separate them into different fractions. [1]

Inputs

The characteristics of the biocrudes from each technology (HTL, FP and CFP) differ substantially. At the same time, the type of feedstock used will also significantly impact the characteristics of the biocrude. Wood-based feedstocks are preferred for their low ash and stable profile, while residues like agricultural waste or sewage sludge introduce nitrogen,

sulphur, and inorganic impurities that deactivate catalysts and complicate operations. All these considerations add to the challenges in upgrading the biocrudes into drop-in biofuels. [1, 5].

Figure 6: Properties of typical Bio-oil and biocrude feedstocks compared with petroleum [5]

	Woody FP Bio-Oil	Woody CFP Bio-Oil	Sewage Sludge HTL Biocrude	Petroleum
H/C	1.4-1.6	1.0-1.1	~1.5	1.5-1.8
O	36-52	15-36	~2-8	0.1-1.0
S	~0	~0	~0.5	0.1-6
N	<0.2	<0.2	~5	0.1-2
H_2O	17-30	~5-10	~2-12	0.02-0.1

Pyrolysis oil is a complex and highly oxygenated feedstock that differs substantially from petroleum-based oils, which has major implications for upgrading and refinery integration. Bio-oils produced via fast pyrolysis (FP) or catalytic fast pyrolysis (CFP) typically contain 35–52 wt% oxygen (FP) or 15–36 wt% oxygen (CFP) and 17–30 wt% water, compared to petroleum which has less than 1 wt% oxygen and less than 0.1 wt% water. Higher oxygen and water content reduces energy density, increases chemical instability, and complicates storage and processing. [5]

Elemental composition also varies: FP and CFP bio-oils have very low sulphur (~0 wt%) and nitrogen (<0.2 wt%), while hydrothermal liquefaction (HTL) biocrudes can contain up to ~5 wt% nitrogen and 2 wt% sulphur. These differences affect catalyst selection and process design when bio-oils are co-processed with fossil feedstocks or used as a replacement. [5]

Acidity is another critical property. Instead of pH, the refinery industry uses Total Acid Number (TAN) as an indicator. Pyrolysis oils typically exhibit TAN values between 50–100 mg KOH/g, compared to heavy fuel oil with TAN <0.5 mg KOH/g. This high acidity makes pyrolysis oils highly reactive and corrosive, posing challenges for storage, transport, and integration into existing refinery infrastructure. [5]

Hydrogen demand for upgrading biocrudes is influenced by several factors. The primary determinant is the chemical composition of the biocrude, particularly its oxygen, nitrogen, and sulphur content, as these require removal through hydrotreating. Higher concentrations of these elements increase hydrogen consumption. The end use also plays a significant role, since full conversion to drop-in fuels requires substantially more hydrogen than mild stabilisation. [6, 7]

Outputs

The primary goal of upgrading biocrudes is to produce a drop-in fuel, i.e., a refined biofuel that can be used directly in existing engines and infrastructure designed for fossil fuels. This makes the technology particularly attractive for sectors that are hard to electrify, such as maritime transport. There is growing interest in using upgraded biocrudes to replace

heavy fuel oil in shipping, where the need for low-emission alternatives is urgent and the compatibility with current systems is crucial.

Most upgrading technologies yield a mix of fuel types, including fractions suitable for diesel, jet fuel, and sometimes naphtha. These are separated through distillation based on boiling point ranges, which is similar to conventional refinery processes. While the exact distribution depends on feedstock and upgrading conditions, most pathways inherently produce multiple fuel types. [1]

Energy balance

Experimental results show that 79% to 87% of the initial energy from the reactants, converted hydrogen and bio-oil, was retained in the upgraded oil phase, depending on the upgrading pathway and feedstock characteristics. [8]

Typical capacities

As the technology is not yet commercially established, there is no standard or typical capacity for standalone upgrading of biocrudes. Transitioning from co-processing to dedicated upgrading requires a series of investments. For example, installing a new oxygen removal unit becomes economically viable at an input scale of approximately 50,000 tonnes of oil per year. However, increasing the scale beyond this may lead to disproportionately high capital expenditures. It is therefore expected that initial facilities will be designed to reach this scale, with capacities gradually increasing as the technology matures and demand for upgrading capacity grows. [4]

Regulation ability

Standalone upgrading of pyrolysis oils has not yet been demonstrated at full commercial scale. The dynamic regulation ability can therefore only be estimated based on experience from conventional refinery hydrotreating units, where operational flexibility is constrained by catalyst stability, hydrogen partial pressure, and thermal management.

Hydrodeoxygenation (HDO) reactors generally operate under steady-state conditions with slow load changes (typically 2–5 %/min) to avoid catalyst deactivation caused by temperature fluctuations and coke formation. Similar or more restrictive constraints are expected for pyrolysis-oil upgrading, due to the higher impurity load and polymerisation tendency of oxygenates. Studies indicate that rapid load cycling increases catalyst fouling, requiring more frequent regeneration intervals.

Space requirements

There are currently no commercial reference plants for standalone upgrading of pyrolysis oil. However, estimates from industry actors suggest that a facility processing approximately 100,000 tonnes per year (≈2,000 barrels per day) may require a footprint of around 60 x 40 meters, excluding hydrogen supply, office buildings, and warehouse. When including auxiliary infrastructure such as tank farms, offices, and storage, the total area requirement may increase by a factor of 2 to 3. Input storage typically includes two stirred tanks to ensure continuous feedstock supply.

Advantages/disadvantages

Advantages

One of the principal advantages of upgrading pyrolysis oil is its potential to serve as a renewable alternative to fossil-based fuels. Once upgraded, the bio-oil can meet the technical specifications required for conventional engines, thereby facilitating integration into current fuel supply chains without the need for extensive infrastructure modifications.

Disadvantages:

A key limitation is the immature development stage of the upgrading technologies, which restricts their commercial deployment and scalability. Furthermore, several upgrading pathways require an input of hydrogen, which must be produced from renewable sources to maintain the sustainability profile of the final fuel. The production of renewable hydrogen via electrolysis is energy-intensive, thus introducing a supplementary technical challenge and additional costs to the overall upgrading process.

To be used as transportation fuels, especially for aviation, which has strict fuel quality standards, the upgraded products must meet demanding specifications in terms of energy density, stability, and purity. This places high requirements on the upgrading process.

Environment

The environmental performance of pyrolysis oil upgrading depends significantly on the overall system efficiency and product yield; high yields are crucial for minimising climate and resource impacts per unit of fuel produced. Co-location of upgrading units with pyrolysis plants or hydrogen supply can reduce energy losses and transport-related emissions.

Regarding air emissions, NO_x and particulate matter emissions from combustion of upgraded pyrolysis fuels are comparable to those from fossil fuels.

Research and development

The upgrading of pyrolysis oil into drop-in biofuels remains under active research, with several technical challenges preventing full-scale commercialisation. Currently, biocrude is primarily utilised for heating applications due to the complexity of upgrading processes. [1]

Although two-step hydrotreatment of fast pyrolysis biocrude has been successfully demonstrated, the overall commercialisation of this technology is still in progress. Key challenges, such as catalyst inactivation and long-term operational stability, have yet to be fully resolved. [1]

In Sweden, tests have shown that pyrolysis oil from Pyrocell can be co-processed with fossil crude in an existing catalytic cracker at the Preem refinery on the west coast. However, due to technical limitations, only a small fraction, up to 5%, of the total feed could be substituted with pyrolysis oil. (Energiforsk, 2016). Achieving higher substitution levels will require hydrodeoxygenation using hydrogen, which has proven difficult in practice. Additionally, other impurities, including metals, acids, and phosphorus, need to be removed to enhance the feasibility of upgrading pyrolysis oil to commercial fuels. These pretreatment processes, however, remain in the research stage [10].

It is estimated that the actual initiation of standalone upgrading of pyrolysis oil will realistically take place between 2030 and 2035, as both the decision-making process and facility construction are expected to require approximately five years. Initial implementation is likely to occur at a modest scale, with capacity expected to accelerate from 2040 onwards. [11]..

Example of Market Standard Technology

The technologies for converting bio-oil (pyrolysis oil) are currently at Technology Readiness Levels (TRL) 2–5, meaning they range from early development to technology demonstration (Hultgren A. E., 2022; study, 2025) (Hultgren A. E., 2022). As such, there are no commercial-scale plants, and there is therefore currently no established market standard technology.

Prediction of Performance and Cost

Due to the early stage of technological development, comprehensive cost data for the upgrading of biocrudes is currently not available but techno-economic models provide a range of estimates for performance and cost.

Analyses by the Pacific Northwest National Laboratory (PNNL) on algae-based HTL report techno-economic estimates for wet biomass conversion and subsequent upgrading. The model indicates that the price required to cover all production costs, could be in the range of US$ 5.4 per gallon ~ 1.2 Euro/litre. [13]

Uncertainties

Several uncertainties remain regarding the large-scale implementation of biocrude upgrading. First, the technical feasibility of long-term, continuous operation at commercial scale is still unproven. Challenges such as catalyst deactivation, coking, and the management of impurities (metals, acids, and phosphorus) could impact both process efficiency and operational stability.

Second, economic uncertainties are significant. The high costs associated with hydrogen consumption, multi-step upgrading, and pretreatment of the bio-oil make the overall economics sensitive to both feedstock prices and market fuel prices.

Third, regulatory and policy frameworks will influence adoption. Incentives for renewable fuels, carbon pricing, and environmental regulations could accelerate or delay commercialisation, while their future evolution remains uncertain.

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