

CZECH TECHNICAL UNIVERSITY IN PRAGUE

FACULTY OF MECHANICAL ENGINEERING

The Department of Process Engineering

**BALANCE OF HYDROGEN PRODUCTION TECHNOLOGY
USING GASIFICATION OF WOOD CHIPS**

MASTER'S THESIS

Author: Bc. Nikita Kozlov
Study program: Process engineering
Supervisor: Doc. Ing. Lukáš Krátký, Ph.D.

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I. OSOBNÍ A STUDIJNÍ ÚDAJE

Příjmení: **Kozlov** Jméno: **Nikita** Osobní číslo: **419587**
Fakulta/ústav: **Fakulta strojní**
Zadávající katedra/ústav: **Ústav procesní a zpracovatelské techniky**
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Balance of hydrogen production technology using gasification of wood chips

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Vypracujte bilanční model výroby vodíku, jehož základem je technologie zplyňování dřevní štěpky. V práci se detailně zaměřte zejména na tyto body.

- 1) Proveďte rešerši možností výroby vodíku pomocí gasifikace odpadů s lignocelulózovým základem.
- 2) Zpracujte bilanční model výroby vodíku s využitím technologie zplyňování dřevní štěpky - PFD schéma, hmotnostní bilance, energetické bilance, základní ekonomika provozu.
- 3) Diskutujte potenciál výroby s konvenčními metodami a proveďte citlivostní analýzu procesu.

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Jméno a pracoviště vedoucí(ho) diplomové práce:

doc. Ing. Lukáš Krátký, Ph.D., ústav procesní a zpracovatelské techniky FS

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doc. Ing. Lukáš Krátký, Ph.D.
podpis vedoucí(ho) práce



prof. Ing. Tomáš Jirout, Ph.D.
podpis vedoucí(ho) ústavu/katedry



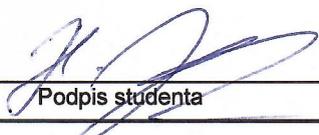
prof. Ing. Michael Valášek, DrSc.
podpis děkana(ky)

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This is to certify that to the best of my knowledge, the content of this thesis is my own work. I certify that the intellectual content of this thesis is the product of my own work and that all the assistance received in preparing this thesis and sources have been acknowledged.

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Signature

Acknowledgement

I would first like to thank my thesis supervisor, **Doc. Ing. Lukáš Krátký, Ph.D.**, for the useful comments, remarks and engagement through the learning process of this master thesis.

Anotace

Tato diplomová práce je zaměřena na vypracování literární a patentové rešerše na téma výroby vodíku z biomasy. Uvádí se zde principy zplyňování a následné způsoby separace vodíku z vyrobeného plynu a zároveň rešerše na téma paliv z odpadů, obsahujících lignocelulózu. V praktické části je uveden návrh modelu zplyňovací jednotky založené na technologiích z rešerše.

Klíčová slova

Vodík, zplyňování, membrány, odpad, lignocelulóza, bilancování

Anotation

This master thesis is aimed at preparing a theoretical framework around hydrogen production out of biomass. Gasification and gas separation principles are amongst the crucial topics along with study on lignocellulosic waste. The practical application focuses on creating a model of a plant that produces hydrogen based on the technologies from the theoretical framework.

Key words

Hydrogen, gasification, membranes, waste, lignocelluloses, balancing

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1 Introduction

Hydrogen is considered by many scientists to be the fuel of the future. Unfortunately for the mankind, pure hydrogen does not exist as a molecule H_2 in nature, making impossible to acquire it the usual way, as, for example, acquiring iron, because it only exists in combination with other elements as, for example, water. Current state-of-art technologies, such as steam reforming and electrolysis, that provide the industry with pure hydrogen, are inefficient from an energy, economic and ecological perspectives.

Hydrogen is used a rising number of applications and thus, the demand for pure (liquid or gas form) is constantly rising. Therefore, researchers and engineers are trying to find more energy and economically efficient ways to produce hydrogen. Currently, the gasification of waste and further membrane separation technology of gases and biotechnological production are the most promising technologies.

2 Waste generation

The humankind produces with each year more and more waste, which can be used for a good purpose. This thesis is mainly focused on wood waste, but this does not mean, that there are no other hydrogen sources in waste. Following three subchapters provide a scope of contemporary waste management situation in Czech Republic, European Union and in the world.

According to World Bank Website [35], the food/green waste and wood waste do a combined 50% of waste. This type of waste contains lignocellulose that is considered to be a promising source hydrogen. The distribution of other waste types incl. food/green and wood waste is shown in the Figure 1.

The wood waste mainly consists of wood chips and sawdust that are residues in wood processing facilities. This type of waste is the primary target for the gasification process because wood is built out of lignin and cellulose. Food and green waste is current object of studies whether it is to use as lignocellulose containing material efficient in hydrogen production. The gasification of lignocellulose containing materials and further hydrogen production is described in following chapters.

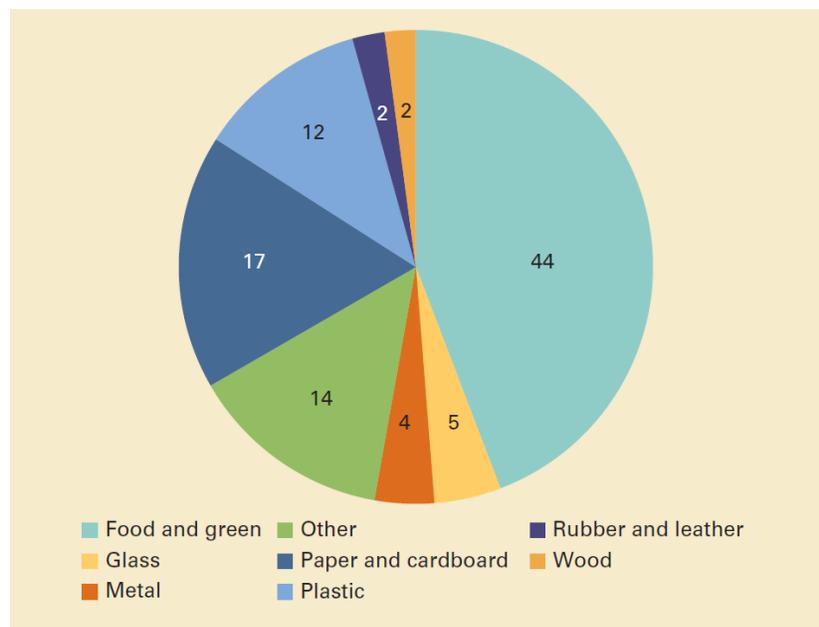


Figure 1 Global waste distribution, %wt DM [35]

2.1 Lignocellulosic waste generation in the Czech Republic

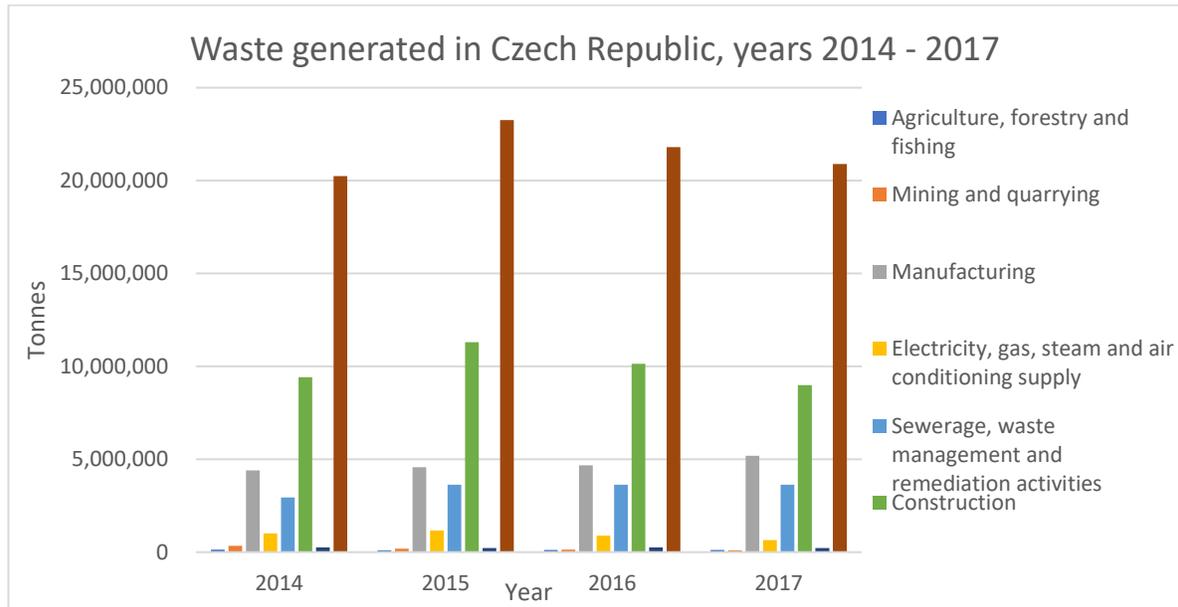


Figure 2 Waste generated in Czech Republic, years 2014 – 2017 [1]

The lignocellulosic waste is presented in the agriculture, forestry and fishing in this case. According to Czech Statistical Office [1]. The agricultural industry, forestry and wood processing industries produced 404 492 tons of waste wood in the year 2017. This number represents only 1.9% of total industry waste created in Czech Republic in the year 2017. The tendency has kept the same for the years 2002-2017 with a little growth as there have been opened more and more industries. For comparison purposes United Kingdom produces annually 7.5 million tonnes of waste wood of which only 16% is recycled and 4% is incinerated, leaving 80% of unused waste wood, which is disposed to landfills all around UK [2].

2.2 Lignocellulosic waste generation in the European Union

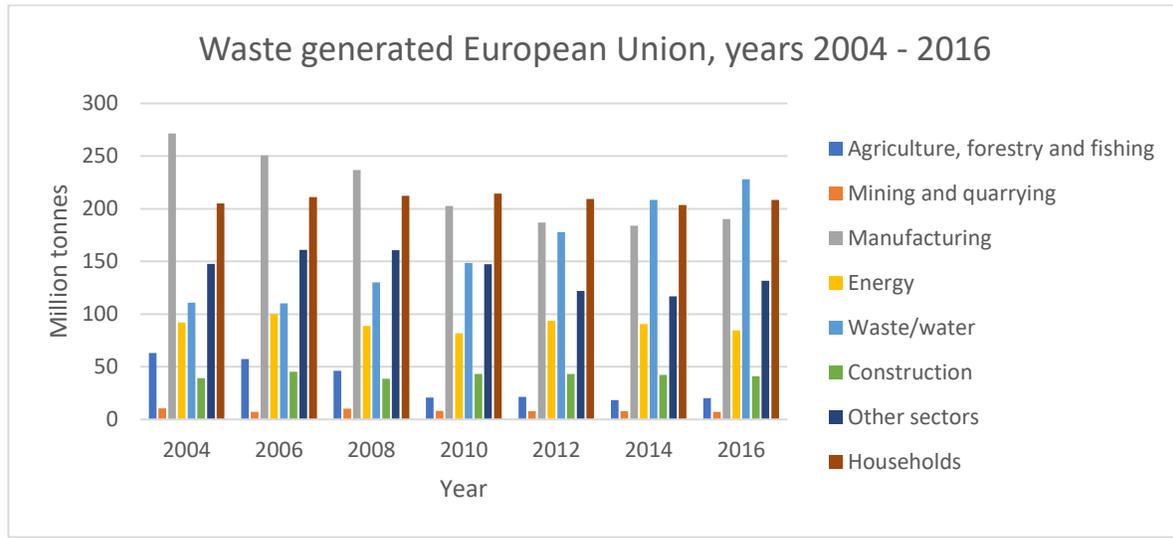


Figure 3 Waste generated in EU, year 2004 – 2016 [4]

According to Eurostat [4] and [5], European Union (all its 28 members together) produced 63.23 million tonnes of wood and paper & cardboard waste in the year 2008, but this number kept constantly declining. The three main waste wood producers in EU are United Kingdom, France and Italy, where Germany, for example, has the highest recycling ratio. This means that European Union is theoretically able to produce roughly 63 million Nm^3 of biogas that is rich in hydrogen. And only forestry waste is taken into consideration which presents a research challenge to search for other lignocellulose waste suitable to produce hydrogen.

2.3 Waste gasification – “dirty energy”

Waste incineration is a common process to acquire heat energy. However, the outlet gases are not used in any way and are usually released into atmosphere.

Another way to recycle, or better said “re-use”, the waste is to gasify it. The gasification process creates a so-called syngas (produced gas) which is usually rich in methane (CH_4), hydrogen (H_2) and carbon oxides (mono- and dioxide) (CO and CO_2). A typical, modern biomass gasification plant scheme is provided in Figure 4. The plant on the scheme is a

Biomass Gasification Project in Sweden, called GoBiGas [6], that is producing biomethane. The GoBiGas [6] plant is the largest plant of its kind and is the first to convert solid biomass to high-quality biomethane [6].

The fuel (wood pellets or woody biomass) enters the gasification reactor (Figure 4, Equipment 1). During the gasification process certain solid and gas compounds are produced. The gas composition depends stark on the operating conditions like temperature, flow rates and catalyst type.

A certain amount of char is produced during the gasification, where part of the char is gasified and the rest is transported to the combustor (Figure 4, Equipment 2), where heat is produced. The syngas (or what is going to be the biomethane afterwards) undergoes several cleaning, purification and reaction steps, amongst which are for example removal of H_2S , water-gas shift reaction, methanation, removal of CO_2 and drying.

The biomass is gasified in a DFB (dual fluidized bed) reactor at relatively low temperature (low for that type of process) of $900\text{ }^\circ\text{C}$. The fluidisation steam is supplied at the temperature of $345\text{ }^\circ\text{C}$ and at a flowrate of $4255\text{ Nm}^3 \cdot h^{-1}$.

The plant is capable to produce 20 MW of biomethane with 8000 operating hours per year, 65% biomass to biomethane efficiency. The total cost of the project has reached 165 M€. The high efficiency is achieved through maxed out usage of process waste like chars and tars. Those compounds are still of some energy value which is clearly visible in the Figure 5 where the process lines show that recirculation used a lot. It is claimed that the overall cold gas efficiency has reached 71.7% LHV. [6]

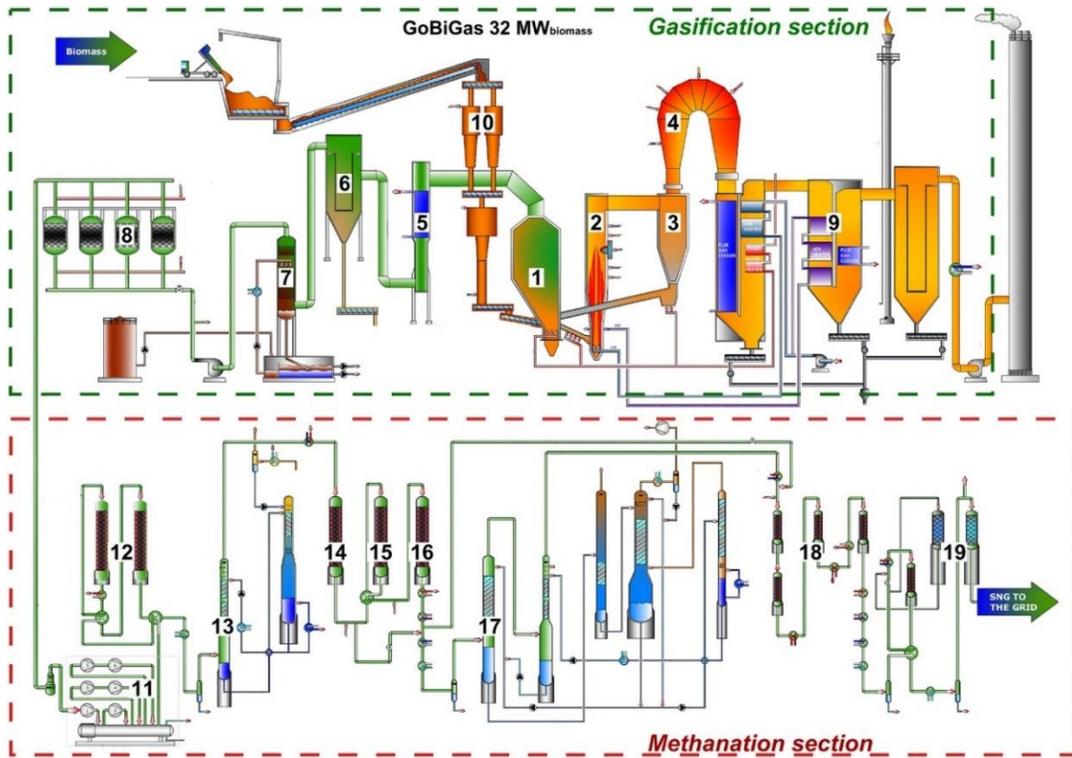


Figure 4 Biomass gasification process scheme [6]

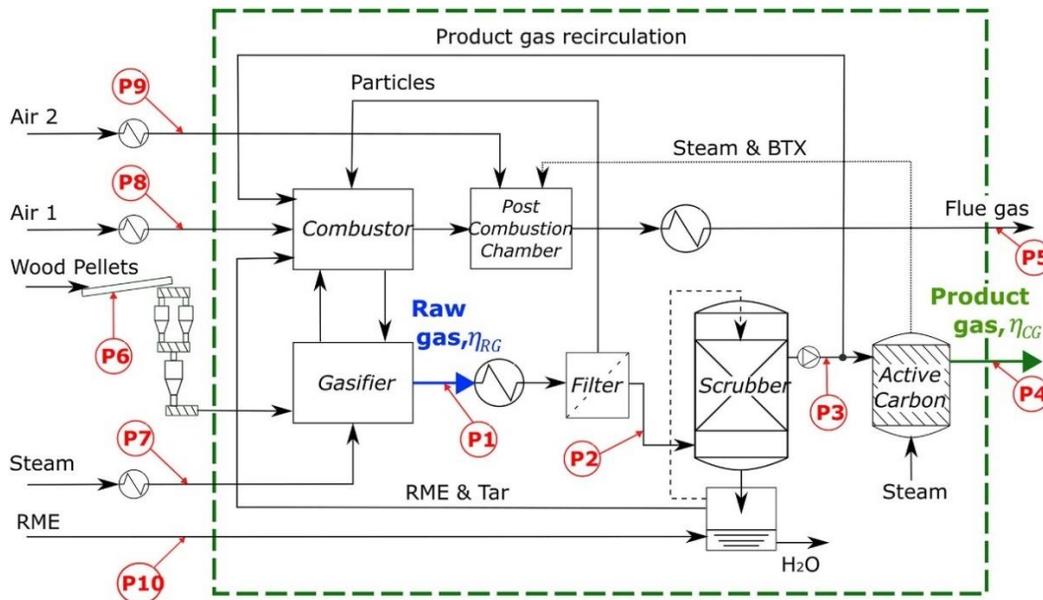


Figure 5 Gasification and recirculation schematic [6]

2.4 Hydrogen production out of lignocellulosic waste

2.4.1 Why produce hydrogen out of waste?

The humankind produces a constantly increasing amount of waste which is either burnt or put to a dump. The flue gas out of the waste incineration is usually cleaned and sent back to atmosphere whereas there might be some useful gases present in the flue gas, like methane or hydrogen. Therefore, scientists do research on waste flue gas composition and seek ways to separate the gases of interest from the rest compounds

2.5 Current state-of-art technology to produce hydrogen

2.5.1 Steam reforming

Steam reforming is the state-of-art technology and it is currently the most used method to produce hydrogen.

A modern process of steam reforming consists of several steps. First of all, the natural gas flow has to be purified from any undesirable elements like sulphur that can poison the catalyst in the reformer. Hydrogen from the process as a recycle gas is mixed natural gas at inlet. Afterwards the desulphurized gas together with steam enters the pre-reformer. Adiabatic steam pre-reforming uses a highly active nickel catalyst to partially reform a desulfurized hydrocarbon feed, using waste heat (480°C) from the convection section of the reformer. The use of waste heat lowers the steam production in the convection section of the reformer furnace, reducing the primary reformer duty and hence gas consumption. While also reducing energy consumption, in general, the installation of a pre-reformer can reduce the size of the primary reformer by up to 25% [8]. Next, the pre-reformed gas mixture enters the main reformer, again mixed up with new steam.

Steam reforming is highly endothermic reaction. The process has better yield if carried out at high temperature, low pressure, and high steam-to-hydrocarbon ratio in order to achieve maximum conversion. The pressures are typically 20–40 bar [9]. The steam reforming process involves reaction described in Table 1 Steam reforming reactions.

Reaction	Reaction enthalpy change $\Delta H_{298} [kJ \cdot mol^{-1}]$
$CH_4 + H_2O \rightleftharpoons CO + 3H_2$	-206
$CO + H_2O \rightleftharpoons CO_2 + H_2$	41

Table 1 Steam reforming reactions

At last, the syngas is cooled down, thus preheating the boiler feed water and then the syngas flows into Gas Purification Section where pure H_2 is obtained.

The price of hydrogen produced this way is dependent on local prices of resources (natural gas, electricity, water, etc.) but for Europe the price is roughly around 4\$/kg of produced hydrogen. [11]

The mean power consumption per produced kg of hydrogen is $0.315 kWh \cdot kg^{-1}$, which approximately 160 times lower than the electric energy consumption when producing hydrogen via electrolysis. But not to forget is the natural gas consumption when producing hydrogen via steam reforming process. The mean natural consumption per produced kg of hydrogen. is $4.43 kg(natural\ gas) \cdot kg^{-1}(hydrogen)$. These data have been taken from a real plant that is daily capable of producing $1\,500\,000 Nm^3 \cdot day^{-1}$ [12]. A usual setup of a steam reforming plant is visualized in the Figure 6.

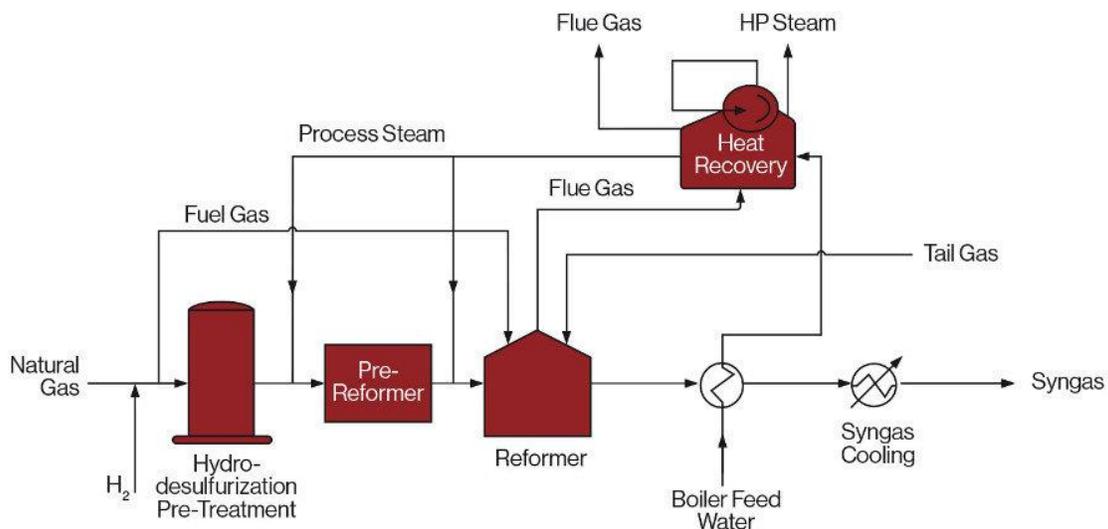


Figure 6 Steam reforming schematic [7]

2.5.2 Electrolysis

Electrolysis is another state-of-art technology to produce hydrogen out of the most abundant resource: water. The process uses electricity to split water molecule into hydrogen and oxygen molecules. An electrolysis unit is built of a tank, where an anode and a cathode are immersed into conducting electrolyte. A direct current is applied to the electrodes (positive terminal to anode and negative terminal to cathode). In general, the hydrogen ions are dragged towards the cathode and the hydroxide ones are dragged towards the anode. A diaphragm is used to separate the two compartments. The gas flow from the electrolyte is then collected by gas collector above the process tank. The overall basic process scheme is shown on the Figure 7. [15]

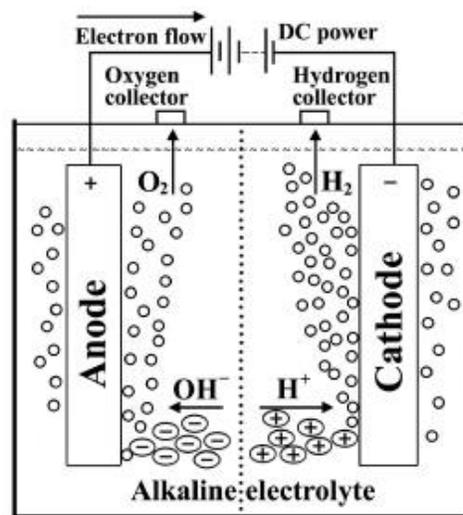
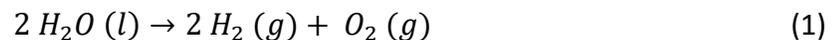


Figure 7 Basic scheme of electrolysis process [15]

There're different electrolyser types that function slightly differently, but two most used technologies are the polymer electrolyte membrane electrolysis and the alkaline electrolysis. The overall electrolysis reaction is shown in (1). [10], [11]



The price of hydrogen production via electrolysis is strongly dependent on local electricity ($\$ \cdot kW^{-1}$) price and water price. The estimated energy spent to produce 1 kg of hydrogen is 50 kWh and 9 liters of water spent. The CAPEX index is decreasing with the plant capacity

and stabilizes at $2600\$ \cdot kW^{-1}$ of installed electrolyzers. The OPEX index is estimated at $45 - 70\$ \cdot kW^{-1} \cdot year^{-1}$. The CAPEX and OPEX are valid for Multi-MW plant with PEM electrolyzers. For alkaline electrolyser reaches the CAPEX index $1350 - 3375 \$ \cdot kW^{-1}$ and OPEX index $20 - 60 \$ \cdot kW^{-1} \cdot year^{-1}$ respectively [13],[14]. Not to forget is that these costs are valid for year 2010 – 2020 and are predicted to fall rapidly within the next 20 – 30 years.

The two following subchapters give a more detailed description of alkaline and PEM electrolysis.

2.5.2.1 Alkaline Electrolyzers

The water electrolysis process requires minerals to be present in the electrolyte so that the reaction occurs. The alkaline electrolyzers have additionally other minerals solved in the electrolyte. Normally potassium hydroxide (KOH) or sodium hydroxide (NaOH) are used. One must be careful when using acidic electrolytes which can cause corrosion of electrodes thus leading to their disfunction. Despite being a well-established and developed technology it has several drawbacks, compared to other technologies, like low degree of purity, corrosive liquid electrolyte and low operational parameters. [15]

2.5.2.2 PEM Electrolyzers

The PEM Electrolysis cell is mainly consisting of Proton Exchange Membrane (PEM) on which electrodes are bonded. The anode is usually coated with iridium oxide (IrO_2) and the cathode is usually coated with platinum (Pt). These materials are used as electrocatalysts to support the electrochemical reaction. The hydrogen is produced so that water is supplied to the anode where a decomposition of water into oxygen, protons and electrons occurs.

The protons are carried through the PEM to the cathode where they combine with electrons to create hydrogen gas. The principle of this technology is shown in the Figure 8. [16]

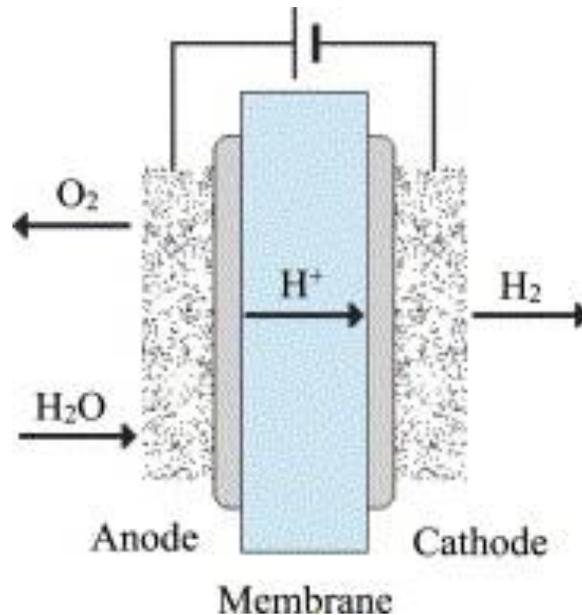


Figure 8 PEM Technology principle [16]

2.5.3 Other technologies

There are several other methods to produce hydrogen, but they are not used very much due to the complexity of the process. Mainly the remaining technologies can be divided into two groups.

The first group consists of gasification and reforming of certain fossils like coal or derived liquids. These processes are quite similar to the steam-methane reforming technology that has been described earlier.

The second group consists of biological processes, where certain bacteria and microalgae can produce hydrogen under beneficial conditions through biological reactions. These processes are currently the objects of research at very early stage, but some to be very promising, like the dark fermentation technology.

2.5.4 Technical assessment of hydrogen production out of lignocellulosic waste

There is a lot of research being conducted on how to produce bio-hydrogen. Unfortunately, the Technology Readiness Level (TRL) is not yet high enough, being at levels starting at Pilot Scale (TRL5) up to Technology Demonstration (TRL6). This means it will take 3-5 more years for the technologies to be research so that they would be applicable to full-scale production lines and 1-3 additional years to build the plants.

3 Aims of thesis

This thesis is aimed at acquiring theoretical and experimental data in order to create a model of hydrogen producing plant based on promising technologies. This sums up what shall be done and what will be the result of the theoretical framework:

- Get an overview about gasification and membrane technologies regarding their technical and economic data.
- Create a Process Flow Diagram with mentioned technologies with all support equipment required.
- Create a model to perform technical and economic assessment.

4 Theoretical framework

4.1 Gasification

Gasification is a process during which heat, pressure and steam convert input material into gas compounds, primarily consisting of hydrocarbons, carbon and hydrogen and their oxides.

4.1.1 Gasification – basic theory

4.1.1.1 Gasification – basic principles

The gasification consists of 5 main processes. Firstly, the entering product (in our case biomass) undergoes a drying process, where bound moisture is released in the form of water vapor (steam).

After that, the raw biomass is pyrolyzed at 200 to 500 °C and is decomposed, breaking down into solids, liquids and gasses in the absence of air or oxygen. The released gasses are hydrogen, carbon monoxide and dioxide and methane.

Then, with increasing temperature, several reactions occur. The first group of reactions are the gas-solid reactions, mentioned in Table 2.

Gas-solid reaction	Reaction enthalpy change $\Delta H_R [MJ \cdot kmol^{-1}]$
$C + \frac{1}{2}O_2 \rightleftharpoons CO$	-110.5
$C + CO_2 \rightleftharpoons 2CO$	-172.4
$C + H_2O \rightleftharpoons H_2 + CO$	131.3
$C + 2H_2 \rightleftharpoons CH_4$	-74.8

Table 2 Gas-solid reaction overview [31]

There is also another type of reactions occurring – gas-phase reactions. These reactions are mentioned in Table 3. [30], [31]

Gas-phase reaction	Reaction enthalpy change $\Delta H_R [MJ \cdot kmol^{-1}]$
$CO + H_2O \rightleftharpoons H_2 + CO_2$	-41.1
$CO + 3H_2 \rightleftharpoons CH_4 + H_2O$	-206.1

Table 3 Gas-phase reactions overview [31]

4.1.1.2 Reactor types, operating conditions etc.

There're four main groups of reactors used as gasifiers.

The first group are the gasifiers with a fixed bed. These are constructional simple and are usually used for small to medium sized plants. Currently the industry uses three different

fixed bed gasifiers: updraft (counter-current), downdraft (co-current) and crossdraft (cross-current). The principle of different Fixed Bed Gasifiers is to be seen in the Figure 9.

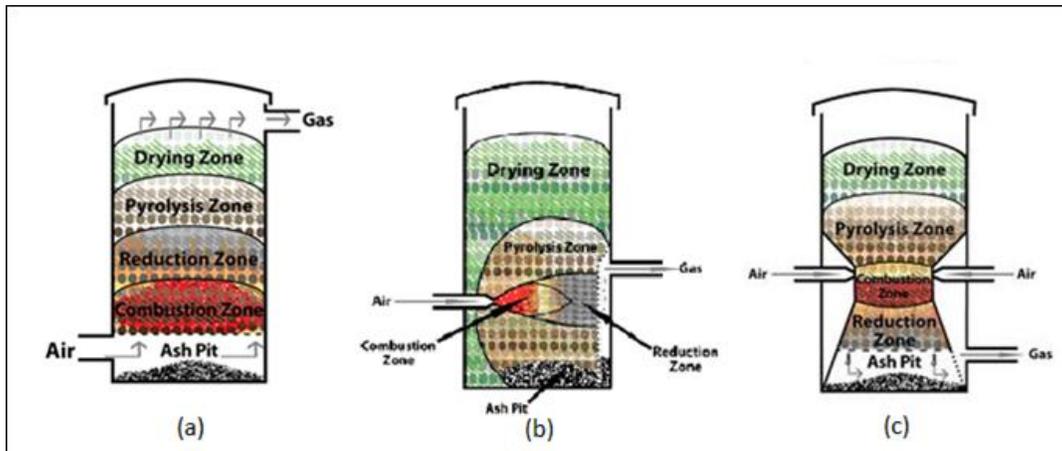


Figure 9 Fixed-Bed Gasifier variations [31]

The second group of reactors are the Fluidized Bed Gasifiers. A fluidizing medium (air, oxygen or steam) is used to create a fluid bed. The bed can consist of sand or some catalyst to bolster the reaction. After the bed is created and operation temperature is reached, the feedstock is supplied. These reactors have better temperature control, independency on feedstock (from sawdust to pellets). The Fluidized bed Gasifiers are to be seen in the Figure 10. [32]

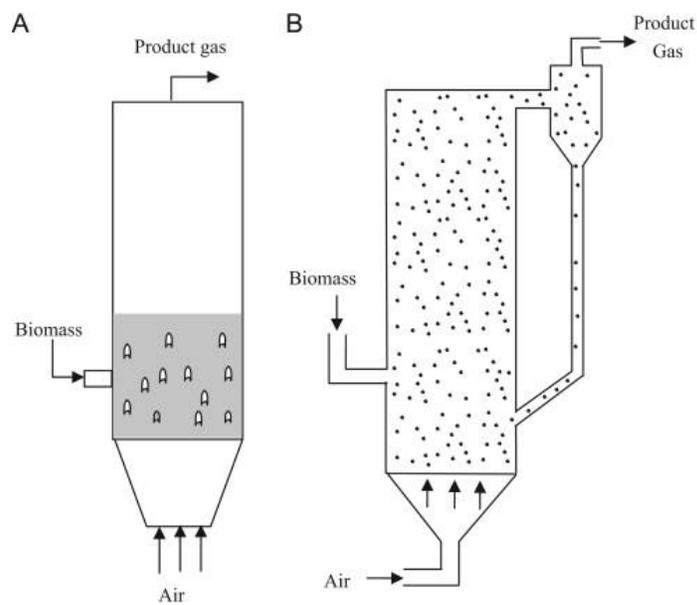


Figure 10 Fluidized Bed Gasifier [32]

The third type of gasifier are the entrained flow gasifiers. They have higher operating conditions, meaning higher temperature level around 1200 °C and at pressure level of 20 bar. This allows the product gas to be directly used in further processes, like synthesis, and results in high carbon conversion (98-99.5%). The schematic of such gasifier is shown in the Figure 11. [31]

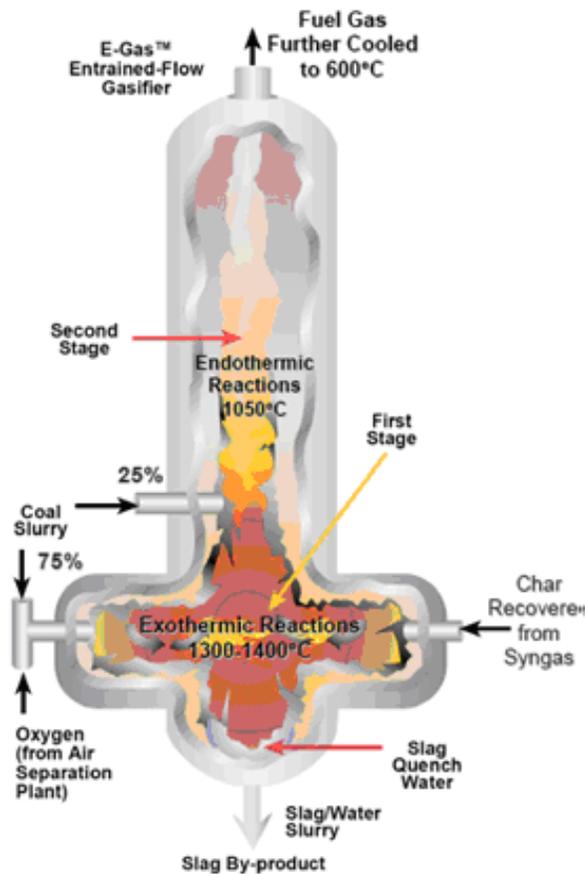


Figure 11 Entrained Flow Gasifier [31]

The last type is the indirect gasifier where the reactions occur in the absence of oxygen and thus the feed is not directly combusted. This means that an external heat source is needed. This might be either a combustion chamber or the heat can be directly sourced by the gasifying agent, as shown on the Figure 12. [31]

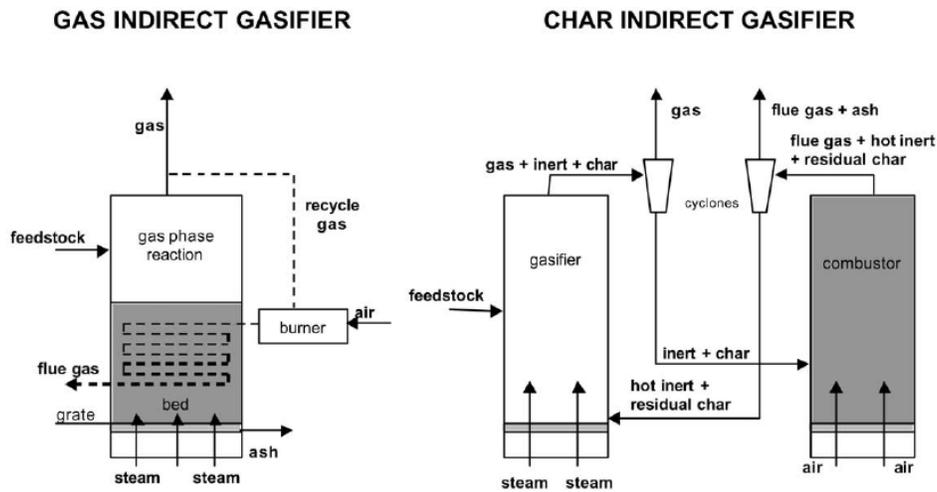


Figure 12 Indirect gasifiers [33]

There has been a serious amount of research done on a new type of gasifiers, the so-called dual fluidized bed gasifiers. These gasifiers achieve high hydrogen content in the producer gas. The main concept is to separate combustion reaction (heat supply) from the gasification reaction (produced gas) so that it is not diluted with nitrogen. The heat carrier is usually the catalyst (e.g. Olivine) or conventional silica sand. The principle is shown in the Figure 13.

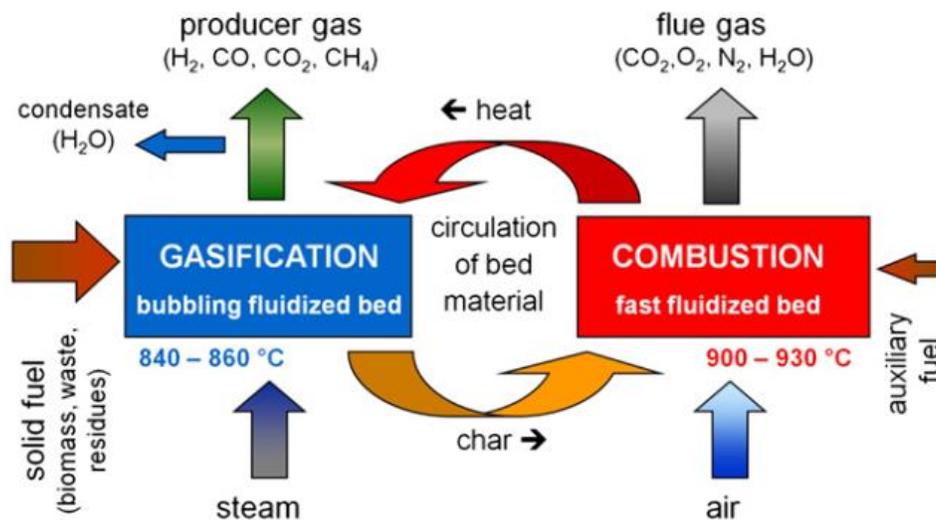


Figure 13 Dual Fluidized Bed System [34]

4.1.2 Practical application of DFB system aimed at hydrogen production

Researchers in [37] were able to apply DFB technology and create a pilot plant with a setup shown in the Figure 14. As it is visible the system is very similar to the plant in Sweden, mentioned in the Chapter 2.3. The bed material is circulating between the combustion and the gasification areas and plays the role of heat carrier. This pilot plant built to consume $(0.7 \div 2.2) \text{ kg} \cdot \text{h}^{-1}$ of biomass fuel, beech pellets in this case.

The aim was to experiment with steam-to-biomass ratio, with different bed materials and with temperature levels. As main bed material Fe-olivine has been chosen because of its catalytic properties and its regeneration ability. All these changes resulted in different syngas composition. These compositions based on the changes are shown in the Table 4 and Table 5.

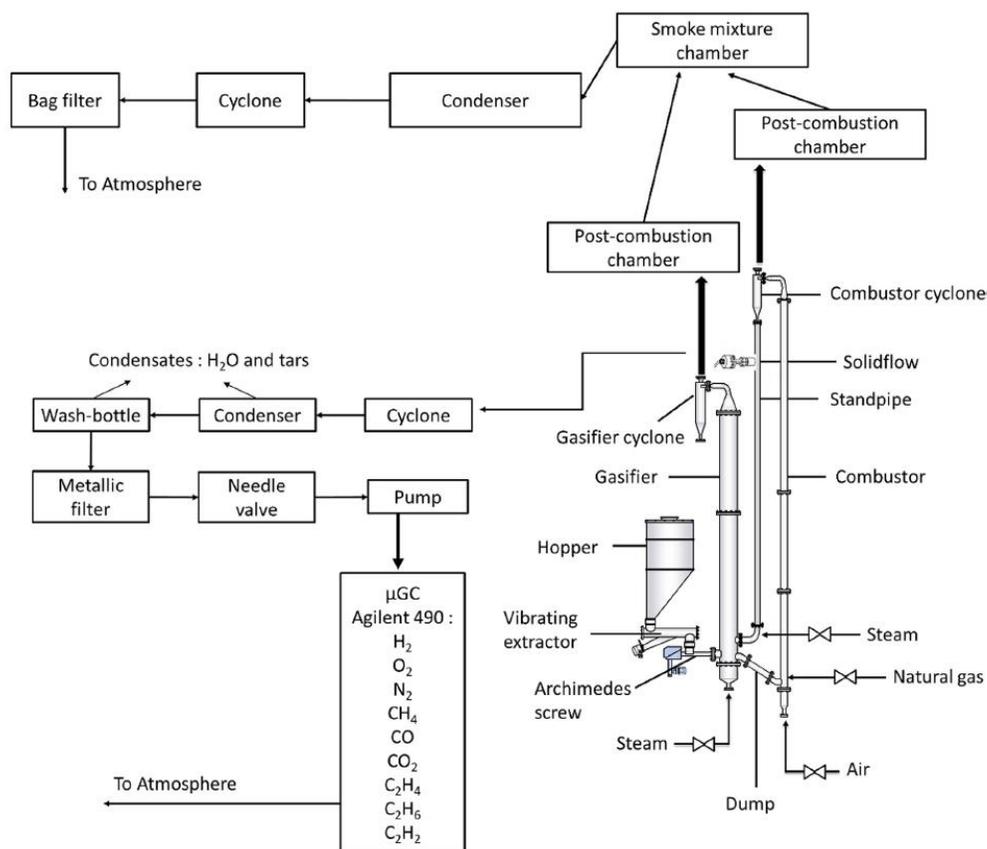


Figure 14 Pilot DFB plant [37]

Test n°	T_B [°C]	$\frac{F_{STEAM}}{F_{BIOMASS}}$ [kg · kg ⁻¹]	$F_{BIOMASS}$ [kg · h ⁻¹]	Bed material
1	850	0	1	Olivine
2	800	0	1.6	Olivine
3	800	0	1.6	Sand
4	850	4.1	1	Olivine
5	850	2	2	Olivine
6	850	2	2	Olivine
7	750	6.5	1	Olivine
8	800	5.7	1	Olivine
9	800	5.2	1	Olivine
10	850	1.4	2.9	Olivine
11	850	1.7	2.4	Olivine
12	850	2.0	2.0	Olivine
13	850	2.6	1.6	Olivine
14	850	2.6	1.6	Olivine
15	850	3.7	1.1	Olivine
16	850	4.1	1.0	Olivine
17	850	5.6	0.7	Olivine
18	820	1.6	2.2	Olivine
19	820	2.0	2.2	Olivine
20	820	2.5	2.2	Olivine
21	820	2.8	2.2	Olivine
22	800	5.7	1.0	Olivine
23	800	5.2	1.0	Olivine

Table 4 Pilot plant results, Part 1 – Beech Pellets [37]

Test n°	y_{H_2} [mol %]	y_{CO} [mol %]	y_{CO_2} [mol %]	y_{CH_4} [mol %]	$y_{C_2H_X}$ [mol %]	P_G [Nm ³ · kg ⁻¹]
1	36.2	43	9.6	8.6	2.6	0.86
2	31	40	13.1	11.6	1.3	0.78
3	29.9	40.9	12.1	12.7	4.3	0.67
4	48.2	16.1	26.2	7.1	2.4	1.4
5	46.2	18.8	25	7.3	2.7	1.2
6	45.5	20.1	25.3	6.9	2.2	1.4
7	49.4	13	28.9	6.4	2.3	1.3
8	46.8	15.6	28.4	6.9	2.3	1.5
9	48.2	13.7	29.4	6.9	2.1	1.5
10	43.1	24.6	21.3	8.4	2.6	1.1
11	43.9	23.1	22.0	8.4	2.6	1.17
12	46.2	18.8	25.0	7.3	2.7	1.2
13	46.3	17.9	25.9	7.6	2.3	1.26
14	46.4	17.6	26.2	7.6	2.2	1.3
15	47.5	16.2	26.4	7.7	2.2	1.37
16	48.3	15.9	26.9	6.7	2.2	1.4
17	48.7	13.6	29.1	6.6	2.0	1.5
18	37.9	28.1	20.7	9.8	3.5	1
19	47.0	18.7	25.2	6.9	2.2	1.1
20	42.8	21.9	24.3	8.0	3.0	1
21	43.6	21.4	24.0	8.1	2.9	0.9
22	46.8	15.6	28.4	6.9	2.3	1.5
23	40.2	25.7	21.8	9.0	3.3	1.2

Table 5 Pilot plant results, Part 2 – Beech Pellets [37]

4.1.3 Membrane gas separation – basic theoretical background

The gas transport is driven by the solution-diffusion mechanism which is governed by the pressure gradient and is divided into three different steps. A sorption on the feed side (high pressure stream) is occurring in the first place. The gas then fluxes through the membranes due to the diffusion mechanism and due to concentration difference and lastly the gas fluxes into the low-pressure stream due to desorption. The process can be described with following equation:

$$P = D \cdot S \quad (2),$$

where P is gas permeability [$Barrer = 10^{-10} \cdot cm^3_{Standard} \cdot cm \cdot cm^{-2} \cdot s^{-1} \cdot cmHg^{-1}$], D stands for diffusion coefficient [$cm^2 \cdot s^{-1}$] and the S is the solubility coefficient [$cm^3_{Standard} \cdot cm^{-3} \cdot cmHg^{-1}$] [22]. The selectivity $\alpha_{A,B}$ is a ratio of individual gas permeability P_A and P_B as per (3)

$$\alpha_{A,B} = \frac{P_A}{P_B} \quad (3).$$

In the next chapters different membranes are presented with their appropriate properties. [22]

4.1.4 Membrane gas separation – currently available and developed technologies

There are currently four different gas membrane types that are commercially available or are under development – polymeric membranes, porous (ceramic, carbon, metallic) membranes, dense metal membranes and ion-conductive membranes. These different membrane types are discussed in the following subchapters. [16] Generally speaking, the palladium (Pd) based membranes, protonic conducting ceramic membrane and oxygen-permeable ceramic membrane are amongst the most potential technologies to acquire high-purity hydrogen. E.g., the palladium-based membranes have an exceptional hydrogen selectivity but suffer from hydrogen embrittlement and lose stability (hydrogen as a catalytic poison) in an atmosphere containing sulphur or carbon and their mixtures. [18]

4.1.4.1 *Polymeric membranes*

Recently serious efforts have been made to produce membranes out of microporous material like carbon molecular sieve, zeolites, metal and covalent organic frameworks and microporous polymers. The last-mentioned show compelling advantages of good processability and low cost against their inorganic and crystalline competitors. [19]

They are usually produced from glassy polymers and are used in industry acquire hydrogen from hydrocarbon streams. However, their main drawback is instability in solvent or under high temperature and pressure. But recent research has shown that combining the organic polymeric membranes and inorganic membranes, thus creating so-called metal-organic membranes, leads to overcoming of mentioned drawbacks of both membrane types. [20], [21]

4.1.4.2 *Porous membranes (ceramic, carbon, metallic) and dense metal membranes*

As it has been pointed out in the previous chapter that the inorganic membranes have greater chemical and thermal stability and resistance, respectively. This type of membranes can be further divided into porous membranes (silicas, carbon, etc.) and dense metal, composite and ceramic membranes.

Porous ceramic membranes possess high permeability, moderate to high selectivity, and are chemically and thermally stable. These membranes can be operated at promising temperature levels of 200 °C and 600 °C.

Porous membranes such as silica membranes have decent hydrogen permeability and selectivity but have limited structure stability in steam-containing atmosphere. A modification made of zeolite are more stable in the presence of steam, but they have lower hydrogen selectivity. Carbon membranes are the final representatives of porous membranes. These membranes share similar properties and can be used at high temperatures of 500 °C to 900 °C. Nevertheless, they brittle and require special handling, special and expensive manufacturing processes thus increasing their price.

Dense metal membranes, like palladium-based membranes, have high hydrogen selectivity and permeability. These membranes can operate at elevated temperatures of 300 °C to 500 °C. Nonetheless there is a danger that the palladium membranes are damaged by hydrogen at lower temperatures (hydrogen embrittlement). This problem can be overcome by alloying the palladium membranes with silver or copper. Additionally, these membranes are highly sensitive to poisoning with sulphur, chlorine and carbon monoxide and can even prevent correct function of the membrane. [21]

4.1.5 Membrane separation – schemes, pros and cons etc. – a discussion

In this paragraph the results of the theoretical framework are discussed, in particular the membrane separation framework. Table 6 and Table 7 presents a scope of achieved results. It is clear that commercially available polymers are not to be considered as promising material for membrane separation. Each of presented materials has its advantages and disadvantages at different properties, but the metal-based membranes appear to be the most suitable ones. They have high permeance, very high hydrogen selectivity, thus producing high purity hydrogen, and can withstand high temperatures. Their only disadvantage is the hydrogen embrittlement and affinity to catalytic poisons from the syngas. Several steps have been taken forward in the means of creating a working membrane separation system, which is claimed by authors in [28]. The palladium-based membrane is shown in the Figure 15.

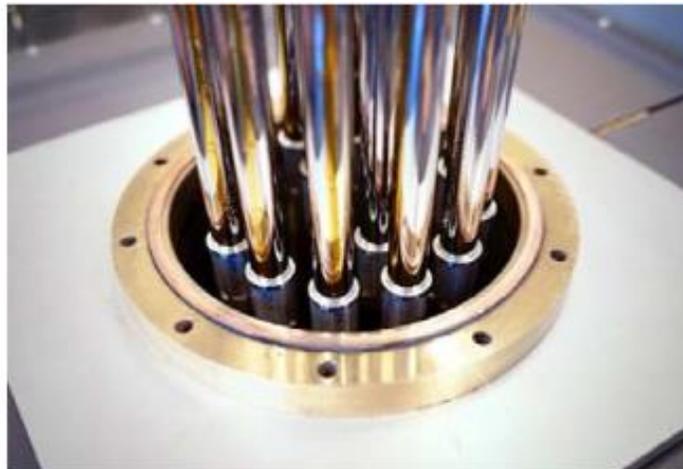


Figure 15 Palladium membrane module [28]

Researchers currently develop methods how to deal with this the metallic membranes' drawbacks, meanwhile other set of material might be promising in not so far future – nanocomposites.

Their main problem is the price. However, some nanocomposite membranes, like graphene oxide coated titanium tubes, show competing results in the means of hydrogen separation.

An overview of metallic and metallic alloy membranes is given in the Table 6 and an overview of inorganic membranes is given in the Table 7.

Material	Permeability [$mol \cdot s^{-1} \cdot m^{-1} \cdot Pa^{-1/2}$]	Temperature [K]	H₂ pressure [kPa]
Pure Metal			
Cu	$5 \cdot 10^{-12}$	773	<0.2
Ni	$2 \cdot 10^{-12}$	500	
Fe	$2 \cdot 10^{-9}$	260-700	20-500
Pd	$9 \cdot 10^{-9}$	500	13
Pt	$8.1 \cdot 10^{-7}$	705-827	100
Alloy			
Ni-Pd	$2 \cdot 10^{-12}$	500	20
Fe3Al	$(6 - 10) \cdot 10^{-11}$	300	101
Zr24Gf12Ni64	$(6 - 20) \cdot 10^{-10}$	473-673	100-300
Nb29Ti31Ni40	$(1.5 - 7) \cdot 10^{-9}$	523-673	100-200
Nb21Ti50Ni29	$(0.9 - 10) \cdot 10^{-9}$	523-673	100-200
VCr4Ti4	$(0.01 - 10) \cdot 10^{-6}$	773-923	
V85Ni15	$(2 - 3) \cdot 10^{-8}$	473-673	<300
(Ni0.6Nb0.4)45Zr50Cu5	$2.3 \cdot 10^{-8}$	673	<300
(Ni0.6Nb0.4)45Zr50Co5	$2.5 \cdot 10^{-8}$	673	<300
Nb20Zr45Ni45	$2.7 \cdot 10^{-8}$	623	100-550
V85Ni15	$(3 - 4) \cdot 10^{-7}$	423-673	10-200
V85Ni10.5Al4.5	$(5 - 7) \cdot 10^{-8}$	423-673	10-200
Pd-Ni	$7 \cdot 10^{-6}$	723	53
Pd-Cu	$6 \cdot 10^{-5}$	723	670
Pd-Ag	$5 \cdot 10^{-5}$	723	20-75
Hf34Ni61Cu5	$1.5 \cdot 10^{-3}$	623	100-300

Table 6 Metallic and metallic alloy membranes [36]

Inorganic	Permanence [GPU]	H ₂ /CO ₂ Selectivity[-]	Temperature [°C]	H ₂ pressure [kPa]
Silica (hydrophilic)	20	11	200	101
Carbon	50	29	250	600
PDMS Coated Carbon	80	42	250	600
SiO ₂ on alumina	300	4	600	0.2
ZSM-5 Zeolite	300	N/A	185	N/A
Zeolite-A	300	10	35	101
Zirconia	300	3.5	25	150
MFI Zeolite	560	18	450	101
AlPO ₄ -5 Zeolite	100	24	35	N/A
Silica (Si600)	1500	N/A	200	N/A
Silica on porous silica	4000	8	300	200
Silica on zirconia	4000	4	300	200
Silica (hydrophobic)	4500	6	200	101
Silica (Si400)	6000	7	200	-
Silicon composite	22500	2	130	103

Table 7 Inorganic membranes [36]

According to authors in [28] they've integrated a membrane module into natural gas power plant, which exactly is integrated into water-gas-shift process, thus creating membrane-water-gas-shift process. Such membrane technology is applied in Tokyo in the natural gas reformer, capable of producing $40 \text{ Nm}^3 \cdot \text{h}^{-1}$.

Table 8 presents commercial metal membrane applications for hydrogen all around the globe. It is apparent that current applications cannot handle a large-scale flow of hydrogen containing gas. Theoretically a battery of such membrane modules would do its job, but a price of such battery would be very high at this moment. Additionally, there are still certain research challenges and drawbacks to make them commercially effective. [29]

Location	Membrane material	Application	Performance
Netherlands	Pd-Au/YSZ/SS	Coal to fuel project	3.5-6 $Nm^3 \cdot h^{-1}$ with 99.5–99.995% hydrogen at 21 Bar pressure difference
Tokyo	Pd-Cu trimetallic alloy	Coal gasification	0.23 $mol \cdot m^{-2} \cdot s^{-1}$ with 99.9999% hydrogen
United States of America	Pd-Y(Gd)-Ag/SS	Reforming	40 $Nm^3 \cdot h^{-1}$ with 99.99% hydrogen
Netherlands	Pd and Pd-alloy	Experimenting	40 – 70 $Nm^3 \cdot h^{-1}$ with 99.99% hydrogen
United States of America	Pd and Pd-alloy	Fluidized bed membrane reactor	0.2 $mol \cdot m^{-2} \cdot s^{-1}$ on Pd-Cu alloy membrane at 673K, syngas presence

Table 8 Membranes application all over the world [29]

4.2 Another approach – membrane reactor

Authors in a recent article [23] have been discussing a potential of membrane reactor to produce hydrogen out of biogas. They propose to build a reactor with in-built palladium-based membrane, which highly permeable and selective towards hydrogen. Biogas is mixed with steam and water at the inlet and then this mixture is autothermal reformed. The reactor itself is a fluidized bed membrane reactor. Authors claim that the price of hydrogen produced this way is lower than via conventional processes, reaching $5\$ \cdot kg_{H_2}^{-1}$. The principle is described in the Figure 16 below.

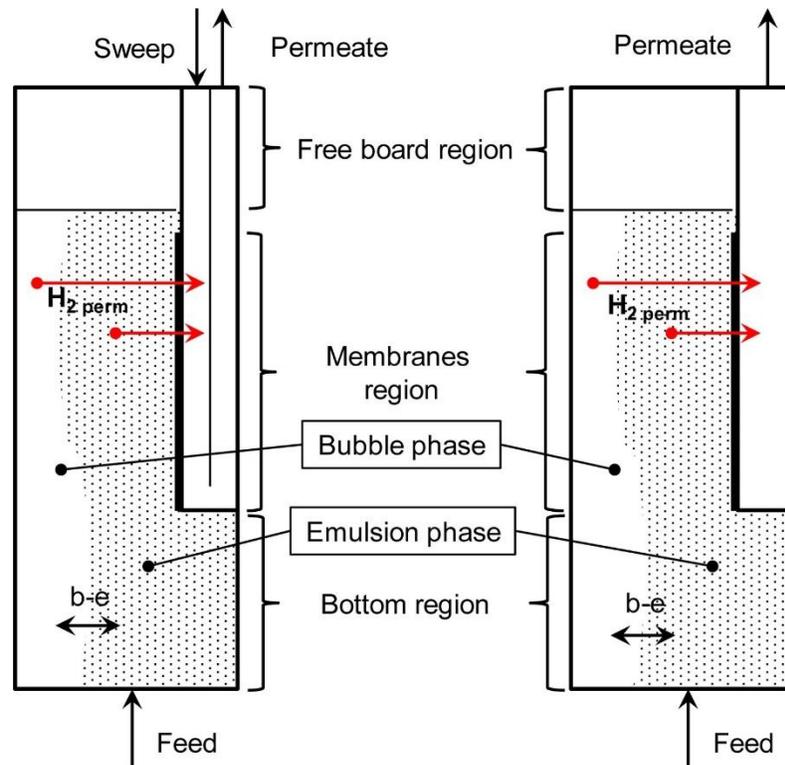


Figure 16 Membrane reactor [23]

5 Practical application

In the practical part of this thesis theory from the previous part has been applied to find a modern, energy and ecologically efficient way to produce hydrogen.

5.1 Hydrogen production technology waste

It was important to choose technology which gives a gas mixture with high hydrogen concentration. So, the DFB (Dual Fluidized Bed Reactor) system has been chosen due to its high hydrogen output (up to 50% vol. in the output syngas). Afterwards a separation method is needed. Although PSA is a state-of-art technology, it is a very complex and cost ineffective and would be replaced with new, emerging method – membrane gas separation. A palladium membrane alloyed with copper (Pd-Cu, see data in Table 6) has been chosen due to reasonable price-to-value ratio and suitable process parameters.

5.1.1 Basic description of designed plant

The PFD in Figure 17 (available in full-scale as Annex 2) represents the designed plant concept. The concept is divided into three main blocks.

The first block is the input of all process material, meaning combustion air, feed water, natural gas, steam, heat carrier/catalyst (olivine in this case) and wooden pellets. In this block, air is filtered of possible impurities and blown into the combustion chamber. Natural gas is supplied to support the combustion process. Steam is mixed together with recovered steam from the process. Wooden pellets and olivine are transported by screw conveyor and then by feeder. The feeder has a shutter built in to prevent any fire getting into the conveyor or even into material storage in hoppers.

The second block represents the main production section where the wooden pellets are gasified in the gasification reactor. The reaction heat is supplied from the combustion chamber.

The third block represents the regeneration section and membrane separation process. A more detailed description of the regeneration process is described in the Chapter 5.1.4.1 and in the Chapter 5.1.4.2, respectively. This block features the most crucial and promising technology of this plant – the membrane separation module. The module is supposed to separate 99.995% of hydrogen from the syngas flow. The residual gas is then blown into the combustion chamber where it helps to decrease the natural gas consumption.

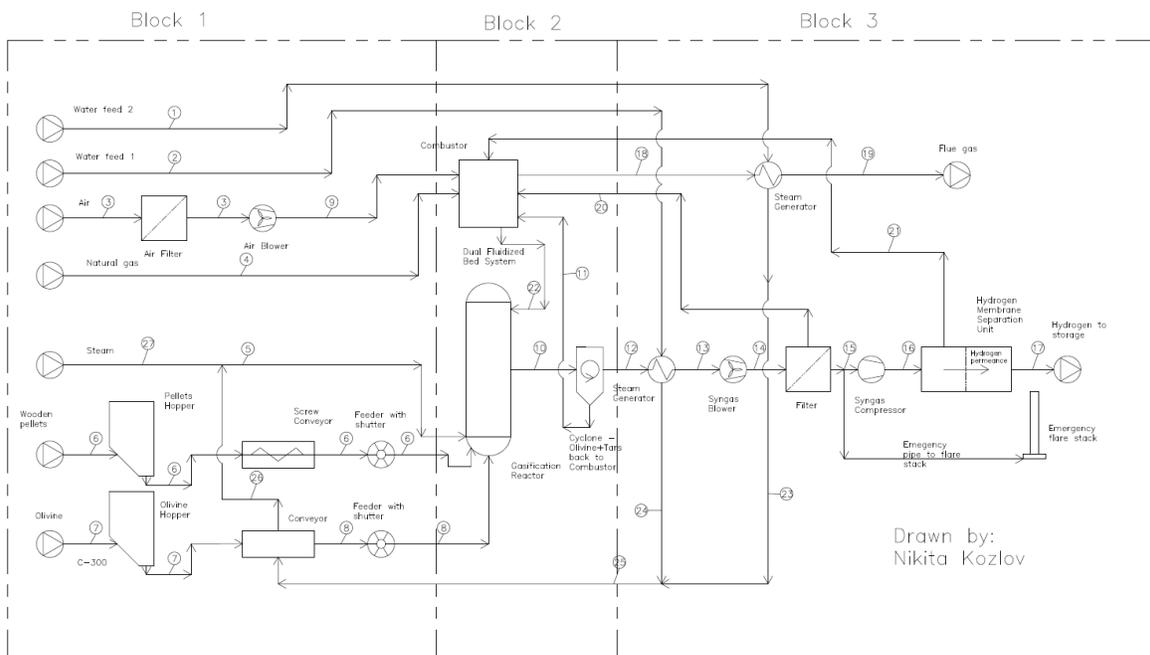


Figure 17 Designed PFD Scheme

5.1.2 Inputs (gasified material)

Many inputs can be considered as suitable for this specific application. But recent researches and experiments, such as [37], have studied wood waste, wood chips and wood pellets as potential source of hydrogen if gasified. This study has been using the DFB system as well, as mentioned in 5.1.

Therefore, the first chapters of this thesis deal with wood waste in Czech Republic, Europe and in the world. It is clear that the wood waste can be a reliable source of hydrogen until other waste sources will be proved as reliable, too.

5.1.3 Energy and mass balance assumptions

Prior to balancing itself few assumptions have to be stated:

- Energy (Conservation of energy) and mass (Conservation of matter) laws are complied
- Test results No. 7 in chapter 4.1.2 has been chosen as the ones to be used in balancing. This means compound concentrations, steam to biomass ratio.
- The energy (air flow, additional fuel flow etc.) balance has been modelled according to chapter 2.3 where the gasification system is very similar to the one here modelled.

5.1.4 Energy and mass balance modelling

To know precise energy and mass balances one would need to build at least a laboratory sized plant and then perform a scale-up process. Fortunately, the suggested technologies exist (although at TRL 5 – 6) and have been tested at least as pilot plants or even small-scale plants and there're publications that presented techno-economic analyses of similar plants. The key difference is that these articles have not been assuming usage of membrane to separate the hydrogen out of produced gas.

The main compounds to be balanced were: water, steam, olive, water, flue and produced gas and their components' concentration. The produced gas composition was based on experiment from [37]. The amount of steam, olive, combustion air and natural gas feed have been determined from the same article.

Table 9 presents the ratios expressed in required feed per $1 \text{ kg} \cdot \text{h}^{-1}$ of biomass fed.

Feed	Unit	Rate per 1 $kg \cdot h^{-1}$ of biomass
Steam	$kg \cdot kg^{-1}$	1
Olivine	$kg \cdot kg^{-1}$	0.0897
Combustion air	$Nm^3 \cdot kg^{-1}$	1.13
Natural gas	$Nm^3 \cdot kg^{-1}$	0.0167

Table 9 Feed per biomass feed [37]

Following two tables present a scope of inputs and outputs from the plant modelled plant.

Table 10 represents the inputs and Table 11 represents the outputs.

Input	Unit	Amount
Steam	$Nm^3 \cdot h^{-1}$	1000
Combustion air	$Nm^3 \cdot h^{-1}$	1120
Natural gas	$Nm^3 \cdot h^{-1}$	15.42
Wooden pellets	$kg \cdot h^{-1}$	1000
Olivine	$kg \cdot h^{-1}$	89.7

Table 10 Model input

Output	Unit	Amount
Flue gas	$Nm^3 \cdot h^{-1}$	2366
Syngas	$Nm^3 \cdot h^{-1}$	1300
Hydrogen	$Nm^3 \cdot h^{-1}$	642.2
Hydrogen-free gas	$Nm^3 \cdot h^{-1}$	657.8

Table 11 Model output

Due to everything being bound to biomass feed, the model is simplified this way, but a certain error should be expected during the calculation as those are average ratios to present the concept of such hydrogen producing plant. So, a model has been set up with the help of Microsoft EXCEL in order to perform a technical and economical balancing and assessments. Figure 18 is a snip of the EXCEL model, which available in Annex 1 as a whole.

Stream	15	16	17	18	19	20
Description	Filtered syngas [Nm ³ /h]	Compressed syngas [Nm ³ /h]	Hydrogen (99.995% purity) [Nm ³ /h]	Hot flue gas [Nm ³ /h]	Cooled flue gas [Nm ³ /h]	Filter residues
Flow	1300	1300	642.2	2366	2366	
Temperature [C]	300	456.6091771	456.6091771	1000	100	
Pressure [kPa]	120	2500	110	120	120	
Concentration [vol%]						
O ₂	-	-	-	0.0435	0.0435	-
N ₂	-	-	-	0.569	0.569	-
CO	0.13	0.13	-	0.0002	0.0002	-
CO ₂	0.289	0.289	5.0E-05	0.1153	0.1153	-
CH ₄	0.064	0.064	-	-	-	-
H ₂	0.494	0.494	0.99995	-	-	-
H ₂ O	-	-	-	0.272	0.272	-
CxHy	0.023	0.023	-	-	-	-
Tar	0	0	-	-	-	-

Figure 18 Model in EXCEL – example

5.1.4.1 Modification and optimization of the process – heat

As shown in Annex 1, the flue gas and syngas have very high temperatures, reaching over 1100 K. Moreover, these gases have high energy due to their high mean heat capacity. It is very appropriate to install water-flue gas exchangers to produce steam and then additionally install steam superheaters because the gasification steam has to be supplied at 345 °C. Thus, it is possible to recover up to 88% of steam required in the process.

Firstly, it is needed to determine how much energy the water will absorb to be transformed into steam and then be superheated to the required operating temperature. The equations (4) shows how much energy is required to create the steam with processes parameters.

$$Q_{TOT} = Q_W + Q_L + Q_S \quad (4),$$

where Q_{TOT} is the total required heat energy, Q_W is the heat energy required to heat water to 100 °C (boiling point), Q_L is the latent heat energy required to convert water to steam and Q_S is the heat energy required to heat steam up to desired temperature. Following equations (5) – (9) show exact calculation procedure for each for the separate energies.

$$Q_W = c_p \Delta T_B \quad (5),$$

where c_p is the water heat capacity, that is equal to $c_p = 4200 \text{ J} \cdot \text{kg}^{-1} \cdot \text{K}^{-1}$ and ΔT_B is the temperature difference to reach the boiling point.

$$Q_L = l_{23} \quad (6),$$

where l_{23} is the water latent heat, that is equal to $l_{23} = 2300000 \text{ J} \cdot \text{kg}^{-1}$

$$Q_S = a \cdot \Delta T_{SH} \quad (7),$$

where a is the heat required to superheat the steam to the desired temperature and is equal to $a = 2000 \text{ J} \cdot \text{kg}^{-1} \cdot \text{K}^{-1}$. ΔT_{SH} is the temperature difference from the boiling point up to desired temperature.

In the case of this thesis the overall energy required is shown in Equations (XX) – (YY), where $\Delta T_B = 102 \text{ K}$ (due to feed water supplied at $20 \text{ }^\circ\text{C}$ and boiling point being at $122 \text{ }^\circ\text{C}$ and 130 kPa) and $\Delta T_{SH} = 223 \text{ K}$.

$$Q_{TOT} = Q_W + Q_L + Q_S \quad (8),$$

$$Q_{TOT} = c_p \Delta T_B + l_{23} \cdot \Delta T_{SH} \quad (9),$$

$$Q_{TOT} = 4200 \cdot 102 + 2.3 \cdot 10^6 + 2000 \cdot 223 = 3\,174\,400 \text{ J} \cdot \text{kg}^{-1} \quad (10).$$

The last Equation defines how much energy it is required to produce 1 kg of process steam, but in order to effectively estimate the heat exchangers and superheaters the energy duty is required, which can be calculated as per Equation (11).

$$\dot{Q} = Q_{TOT} \cdot \dot{m} = 3\,174\,400 \cdot \frac{1000}{3600} = 881\,777 \text{ W} \quad (11),$$

where \dot{m} is amount of the process steam per hour. This means that to recover all steam the plant needs a heat duty source of nearly 900 kW if $1000 \text{ kg} \cdot \text{h}^{-1}$ of biomass is gasified. This heat, as mentioned before, can be found in flue and produced gases.

The component's concentration on gases is known due to experiments published in [6] and [37]. Thus, it is able to determine the gases mean heat capacity and so acquire energy transported by these gases. Table 12, Table 13, Table 14 and Table 15 present calculation preparation and its results based on above-mentioned references.

Gas compound	Concentration vol. c^V [%]
CO	0.13
CO_2	0.289
CH_4	0.064
H_2	0.494
C_2H_6	0.023

Table 12 Syngas compound concentration

Gas compound	Concentration vol. c^V [%]
O_2	0.0435
N_2	0.569
CO	0.0002
CO_2	0.1153
H_2O	0.272

Table 13 Flue gas compound concentration

Heat capacity	Value [$kJ \cdot kg^{-1} \cdot K^{-1}$]
Syngas	8.1984
Flue gas	1.3968

Table 14 Resulting heat capacities on the temperature levels

Prior to the next table it is important to mention that an efficiency of 70% has been assumed regarding the heat exchangers, steam superheaters and of the piping system losses. Table 15 shows heat available in the output gases prior to any cleaning and compression processes. The heat duty for syngas has been calculated for a temperature drop from 750 °C down to 215 °C. The flue gas is cooled down from 900 °C to 100 °C.

Name of measure	Unit	Amount
Syngas heat duty	kW	508.65
Flue gas heat duty	kW	273.94
Total provided heat energy	kW	782.59

Table 15 Available heat in output gases

The result presents that it is possible to recover up to 88% of steam which greatly reduces operational costs but increases capital expenditure. Some additional heat can be recovered by burning ashes that are filtered from the syngas and tar that occur the olivine. Other possible modification to recover heat could be reduce losses in the system, e.g. shorter piping systems, more efficient heat exchangers etc.

In other words, available heat produces $887.512 \text{ kg} \cdot \text{h}^{-1}$ of steam, decreasing the amount of supplied steam to $112.488 \text{ kg} \cdot \text{h}^{-1}$ from an external source.

5.1.4.2 Modification and optimization of the process – membrane separation

To maintain the reaction in additional fuel is required so natural gas is fed, and its amount is provided in

Table 9. However, this parameter can be lowered. After the hydrogen is successfully separated in the membrane module, the residue gas still has high concentration of potential compounds to be burned, like methane and carbon monoxide. If applied to the thesis case, the flow of methane (main natural gas compound) is equal to $85.21 \text{ Nm}^3 \cdot \text{h}^{-1}$. The carbon monoxide has a flow rate of $168.98 \text{ Nm}^3 \cdot \text{h}^{-1}$, but has significantly lower heating value, 3.2 times less to be precise. Together this hydrogen-free gas has a heating value of $28.21 \text{ MJ} \cdot \text{Nm}^{-3}$ compared to natural gas, that has a heating value of $33.5 \text{ MJ} \cdot \text{Nm}^{-3}$. If we calculate heat duty provided by the natural gas in (12)

$$\dot{Q} = \dot{V} \cdot LVH_{NG} = \frac{1000}{3600} \cdot 0.0167 \cdot 33.5 = 0.155 \text{ MW} \quad (12),$$

we can then estimate how much of the residual syngas is needed to maintain the heat, meaning the rest can be stored and sold as residual gas. The equation (13) calculates the amount of necessary syngas.

$$\dot{V}_{TO\ BURN} = \frac{\dot{Q}}{LVH_{SYNGAS}} = \frac{0.155}{28.21} \cdot 3600 = 19.78 \text{ Nm}^3 \cdot \text{h}^{-1} \quad (13),$$

This means that now we can estimate the percentage of carbon monoxide and methane in the syngas-to-combustion followingly:

$$c_{CH_4}^V = \frac{\dot{V}_{CH_4 TOTAL}}{\dot{V}_{CH_4 TOTAL} + \dot{V}_{CO TOTAL}} = \frac{85.21}{85.21 + 168.98} = 0.3352 \quad (14),$$

$$c_{CO}^V = \frac{\dot{V}_{CO TOTAL}}{\dot{V}_{CH_4 TOTAL} + \dot{V}_{CO TOTAL}} = \frac{168.98}{85.21 + 168.98} = 0.6648 \quad (15),$$

Now it is visible that we have to determine the amount of carbon monoxide taken from the total hydrogen-free syngas, then calculate the ratio and subtract this percentage from the syngas as a whole. The ratio of carbon monoxide to subtracted from the syngas flow:

$$\dot{V}_{CO SYNGAS} = c_{CO}^V \cdot \dot{V}_{TO BURN} = 0.6648 \cdot 19.78 = 13.15 \text{ Nm}^3 \cdot \text{h}^{-1} \quad (16),$$

$$r_{CO SYNGAS} = \frac{\dot{V}_{TO BURN}}{\dot{V}_{CO TOTAL}} = \frac{13.15}{168.98} = 0.078 = 7.8\% \quad (17)$$

This means that at least 7.8% of the total syngas shall be subtracted and the rest may be stored. To keep things safe, I would raise this percentage up to 20% so that there is enough heat to maintain the gasification process. This in total leaves following amount of syngas to be stored.

$$\begin{aligned} \dot{V}_{SYNGAS STORAGE} &= \dot{V}_{TOTAL} - r_{CO SYNGAS} \cdot \dot{V}_{TOTAL} = 657.8 - 0.2 \cdot 657.8 \\ &= 526.24 \text{ Nm}^3 \cdot \text{h}^{-1} \end{aligned} \quad (18)$$

However, this particular part is an object of discussing meaning this is only a suggestion how to possibly decrease operational expenditure or find another source of revenue. This modification has not been implemented into the model and is only based on its calculations, but the benefits and drawbacks are discussed in the Discussion section.

5.1.4.3 Modification and optimization of the process – olivine circulation

The olivine ratio mentioned in

Table 9 is so low because it only circulates between the reactor and the combustion chamber. It has a lifetime of 150 hours and then it needs to be either exchanged or regenerated. This case presumes that fresh olivine will be used but a regeneration section could be built at the plant in the future.

5.1.5 Energy and mass balance of the process

For the purpose of this thesis the amount of fuel, namely wooden biomass, has been set to $1000 \text{ kg} \cdot \text{h}^{-1}$. As per

Table 9 the resulting input material consumptions are shown in the Table 16 with previously mentioned optimization.

Input material	Unit	Feed rate
Steam	$\text{kg} \cdot \text{h}^{-1}$	112.488
Olivine	$\text{kg} \cdot \text{h}^{-1}$	89.7
Combustion air	$\text{Nm}^3 \cdot \text{h}^{-1}$	1120
Natural gas	$\text{Nm}^3 \cdot \text{h}^{-1}$	16.716

Table 16 Feed rates for thesis case

For this case an operating time per year of $7500 \text{ hours} \cdot \text{year}^{-1}$ has been taken thus resulting in yearly consumption presented in table

Input material	Unit	Feed rate
Steam	$\text{kg} \cdot \text{year}^{-1}$	775 845
Olivine	$\text{kg} \cdot \text{year}^{-1}$	1 422 750
Combustion air	$\text{Nm}^3 \cdot \text{year}^{-1}$	8 400 000
Natural gas	$\text{Nm}^3 \cdot \text{year}^{-1}$	125 370

Table 17 Annual material consumption

5.1.6 Technical and economical assessment

5.1.6.1 Economical assessment – operational expenses

In order to perform economical assessment prices of input materials and electricity are required. As for electric energy the main consumer is the syngas compressor. The gas has to be fed to the chosen membrane (see Chapter 5.1) at temperature level of 723 K and pressure level of 670 kPa . The equation (19) shows how to calculate the power duty of a compressor and thus its electricity consumption and Table 18 shows all the necessary parameters.

$$P = \frac{P_{ad}}{\eta_{ad}} = \frac{\kappa}{\kappa - 1} \cdot p_1 \cdot \dot{V}_1 \cdot \left[\left(\frac{p_2}{p_1} \right)^{\frac{\kappa-1}{\kappa}} - 1 \right] \cdot \frac{1}{\eta_{ad}} \quad (19),$$

Property	Name	Unit	Rate
κ	Adiabatic index	–	1.2914
p_1	Inlet pressure	<i>kPa</i>	120
p_2	Outlet pressure	<i>kPa</i>	670
η_{ad}	Adiabatic efficiency	–	0.7
T_1	Inlet temperature	K	490

Table 18 Compressor duty parameters

The syngas is fed at the rate of $1300 \text{ Nm}^3 \cdot \text{h}^{-1}$ and thus it is needed to recalculate this rate to $\text{m}^3 \cdot \text{h}^{-1}$ which is done in Equation (20) below. The equation assumes Standard Conditions for Temperature and Pressure, meaning temperature of $T_{STP} = 273.15 \text{ K}$ and pressure of $p_{STP} = 100 \text{ kPa}$

$$\dot{V}_1 = \dot{V}_{SYNGAS\ FEED} \cdot \frac{T_1}{T_{STP}} \cdot \frac{p_{STP}}{p_1} = \frac{1300}{3600} \cdot \frac{490}{273.15} \cdot \frac{100}{120} = 0.54 \text{ m}^3 \cdot \text{s}^{-1} \quad (20).$$

Afterwards compressor duty can be calculated as per (21)

$$P_{COMPRESSOR} = \frac{1.2914}{1.2914 - 1} \cdot 120000 \cdot 0.54 \cdot \left[\left(\frac{670000}{120000} \right)^{\frac{1.2914-1}{1.2914}} - 1 \right] \cdot \frac{1}{0.7} \quad (21).$$

$$= 194.62 \text{ kW}$$

Additionally, there is a combustion air blower that has been calculated the same way with data provided in the Table 19 and according to Equations 19, 20 and 21.

Property	Name	Unit	Rate
κ	Adiabatic index	–	1.4
p_1	Inlet pressure	<i>kPa</i>	101.325
p_2	Outlet pressure	<i>kPa</i>	120
η_{ad}	Adiabatic efficiency	–	0.7
T_1	Inlet temperature	K	293.15

Table 19 Air blower duty parameters

$$\dot{V}_{AIR} = \dot{V}_{AIR FEED} \cdot \frac{T_1}{T_{STP}} \cdot \frac{p_{STP}}{p_1} = \frac{1120}{3600} \cdot \frac{293.15}{273.15} \cdot \frac{100}{101.325} = 0.33 \text{ m}^3 \cdot \text{s}^{-1} \quad (22).$$

$$P_{BLOWER} = \frac{1.4}{1.4 - 1} \cdot 120000 \cdot 0.33 \cdot \left[\left(\frac{120000}{101325} \right)^{\frac{1.4-1}{1.4}} - 1 \right] \cdot \frac{1}{0.7} = 8.27 \text{ kW} \quad (23).$$

Together with all other necessary plant equipment it is appropriate to raise the plant consumption to 225 kW. Table 20 presents annual electricity consumption of the modelled plant.

Property	Unit	Value
Annual operating hours	<i>hour</i>	7500
Electricity consumption	<i>kW</i>	225
Total annual electricity consumption	<i>kWh · year⁻¹</i>	1 687 500

Table 20 Total annual electricity consumption

To finalize the annual expenditure estimation personal costs are required. Such technology module would require 3 operators to secure correct function. Table 21 showcases the personal costs with wages valid for Czech Republic.

Property	Unit	Value
Operators	<i>Person</i>	3
Yearly wage	<i>USD · year⁻¹</i>	21 419
Total personal costs	<i>USD · year⁻¹</i>	64 257

Table 21 Annual personal costs

Now everything is prepared to estimate OpEx (Operational Expenditure) in $USD \cdot kg_{Hydrogen}^{-1}$ ratio. Table 22 sums up all prices per material, that are valid for Czech Republic, and electrical energy.

Material	Unit	Value
Steam	$USD \cdot kg^{-1}$	0.035
Olivine	$USD \cdot kg^{-1}$	0.18
Wooden pellets	$USD \cdot kg^{-1}$	0.103
Natural gas	$USD \cdot Nm^{-3}$	0.75
Electrical energy	$USD \cdot kWh^{-1}$	0.06
Water	$USD \cdot kg^{-1}$	0.003

Table 22 Prices per material unit

By combining the data from Table 17, Table 20 , Table 21 and Table 22 total annual operational expenditure can be estimated which is presented in Table 23 and for better view is visualized in Figure 19.

Material	Unit	Value
Steam	$USD \cdot year^{-1}$	29 746
Olivine	$USD \cdot year^{-1}$	121 095
Wooden pellets	$USD \cdot year^{-1}$	773 468
Natural gas	$USD \cdot year^{-1}$	93 933
Electrical energy	$USD \cdot year^{-1}$	101 146
Water	$USD \cdot year^{-1}$	19 969
Personal costs	$USD \cdot year^{-1}$	64 257
Total annual costs	$USD \cdot year^{-1}$	1 203 615

Table 23 Annual material expenses

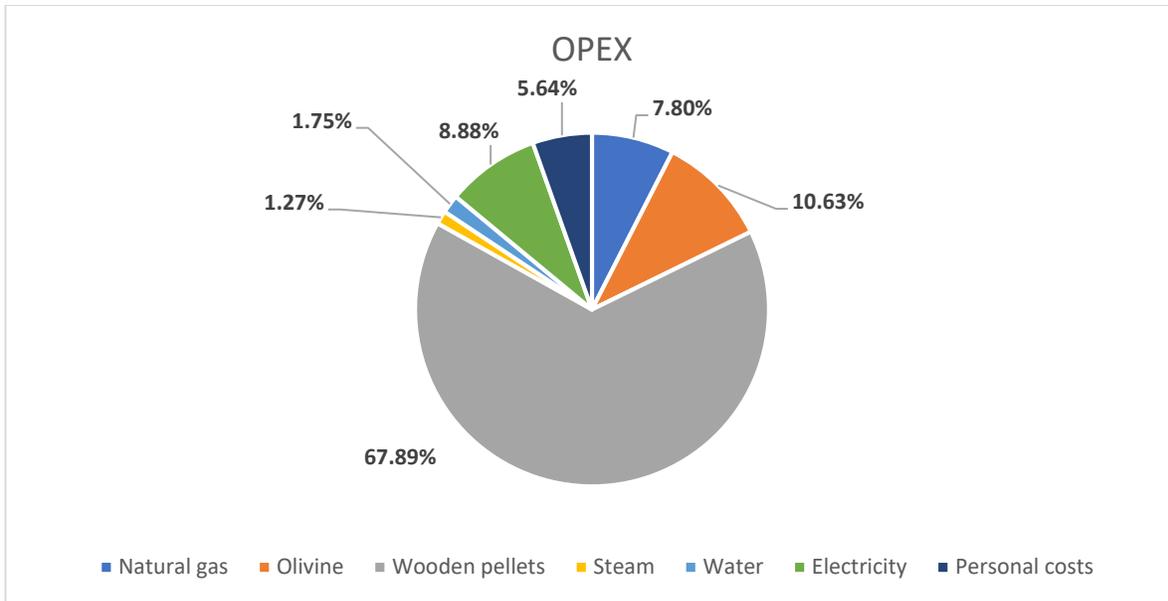


Figure 19 OpEx visualization

Totally, the OpEx index can be calculated with the Equation (24)

$$\begin{aligned}
 OPEX &= \frac{\text{Operational expenses [USD} \cdot \text{year}^{-1}]}{\text{Annual production [kg}_{\text{Hydrogen}} \cdot \text{year}^{-1}]} = \frac{1\,203\,615}{400\,732,8} \quad (24). \\
 &= 3.004 \text{ USD} \cdot \text{kg}_{\text{Produced Hydrogen}}^{-1}
 \end{aligned}$$

It is important to mention that depreciation and other economic indexes have not been included into the OpEx and CapEx.

The Figure 19 clearly visualizes that the major impact on OpEx has the price of wooden pellets and is seconded by olivine prices. The impacts are described in detail in the Chapter 5.1.7.

5.1.6.2 Economical assessment – capital expenses

In order to estimate technology capital expenses equipment prices are required. A cost estimation website [42] has been used. It requires process data or additional calculated data to estimate the expenses.

Firstly, the heat exchangers' heat transfer area shall be calculated. Thanks to heat duty calculations in Chapter 5.1.4.1 it is possible to estimate are. In following equations syngas heat exchanger will have index 1 and flue gas heat exchanger will have index 2.

Heat transfer coefficient has been set up to $k = 32.5 \text{ W} \cdot \text{m}^{-2} \cdot \text{K}^{-1}$ [41]. The area can be calculated from Equation 25 as follows

$$\dot{Q} = kS\Delta T \rightarrow S = \frac{\dot{Q}}{k\Delta T} \quad (25),$$

where $\dot{Q}[\text{kW}]$ is the heat duty, $S[\text{m}^2]$ is the effective heat transfer area and $\Delta T[\text{K}]$ is the temperature gradient. The temperature for syngas and flue gas is $\Delta T_1 = 532 \text{ K}$ and $\Delta T_2 = 900 \text{ K}$. Thus, resulting in

$$S_1 = \frac{\dot{Q}_1}{k\Delta T_1} = \frac{508.65 \cdot 1000}{32.5 \cdot 532} = 29.78 \text{ m}^2 \quad (26),$$

$$S_2 = \frac{\dot{Q}_2}{k\Delta T_2} = \frac{274.94 \cdot 1000}{32.5 \cdot 900} = 9.37 \text{ m}^2 \quad (27).$$

This data has been used to estimate heat exchanger price.

Secondly, membrane price has been estimated. The chosen Palladium alloyed with copper membrane price is calculated as stated in Table 24.

Property	Unit	Value
Membrane flux	$\text{mol} \cdot \text{m}^{-2} \cdot \text{s}^{-1}$	0.8
Gas flux	$\text{mol} \cdot \text{s}^{-1}$	7.42
Price per area	$\text{USD} \cdot \text{m}^{-2}$	10 763.92
Basic module cost	<i>USD</i>	20 500
Total membrane cost	<i>USD</i>	120 350

Table 24 Membrane cost estimation

The resting technology equipment as per PFD and their cost are listed in Table 25 [42].

Equipment	Price [USD]
Heat exchanger 1 – Syngas	29 100
Heat exchanger 1 – Flue gas	15 500
Syngas compressor	77 600
Cyclone	6 900
Blower – Air	5 400
Blower – Syngas	5 100
Air filter	21 800
Reactor	67 900
Syngas filter	21 800
Hopper – Pellets	3 000
Hopper – Olivine	3 000
Screw conveyor – Pellets	8 300
Screw conveyor – Olivine	8 300
Feeder – Pellets	3 000
Feeder – Olivine	3 000
Membrane	120 350
Total equipment cost	400 050

Table 25 Total equipment cost [42]

The equipment costs presented in Table 25 are only the basic expenses. The additional costs bound with installation of equipment and other cost categories are presented in Table 26 Additional costs overview as per [43].

Additional costs category	Percentage of total equipment cost [%]
Equipment installation	45
Instrumentation and controls	18
Piping	16
Electrical systems	10
Building	15
Yard improvements	15
Service facilities	40
Engineering and supervision	33
Legal expenses	39
Construction expenses	4
Contractor's fee	17
Contingency	35
Total additional cost percentage of equipment cost	287

Table 26 Additional costs overview [43]

By summarizing Table 25 and Table 26 total project investment cost is estimated, which is given in Table 27.

Additional costs category	Costs [USD]
Equipment cost	400 048,47
Additional costs	1 148 139.1
Total project investments	1 548 187.57

Table 27 Total project investments

5.1.6.3 Economical assessment – Payback time

The Payback time has been calculated for four different cases. These cases differ in hydrogen selling price ranging from $3.30 \text{ USD} \cdot \text{kg}^{-1}$ to $5.1 \text{ USD} \cdot \text{kg}^{-1}$. The Payback Time is static in this case and presumed a selling tax of 19,0%.

In order to estimate Payback period, static cumulative Cashflow has to be calculated. The methodology is shown in Equations (28) and (29).

$$CCF = -INV - OPEX \cdot t + C_{H_2} \cdot N \cdot t - C_{H_2} \cdot N \cdot t \cdot Tax \quad (28),$$

$$t = \frac{CCF + INV}{C_{H_2} \cdot N - C_{H_2} \cdot N \cdot Tax - OPEX} \quad (29),$$

where CCF [USD] is the cumulative cashflow, which has to be equal to 0 if calculating Payback time, INV [USD] are the project investments, $OPEX$ [USD · year⁻¹] are the total operational expenses, C_{H_2} [USD · kg⁻¹] is the hydrogen selling price, N [kg_{H₂} · year⁻¹] is the hydrogen annual production capacity and t [year] is the payback time. Table 28 scopes the results for different selling prices resulting in different Payback periods and Figure 20 Payback times gives a graphical visualization of such estimation.

Selling price C_{H_2} [USD · kg ⁻¹]	Payback Time t [year]
3.30	Never
3.90	24.27
4.5	6
5.1	3.41

Table 28 Payback times

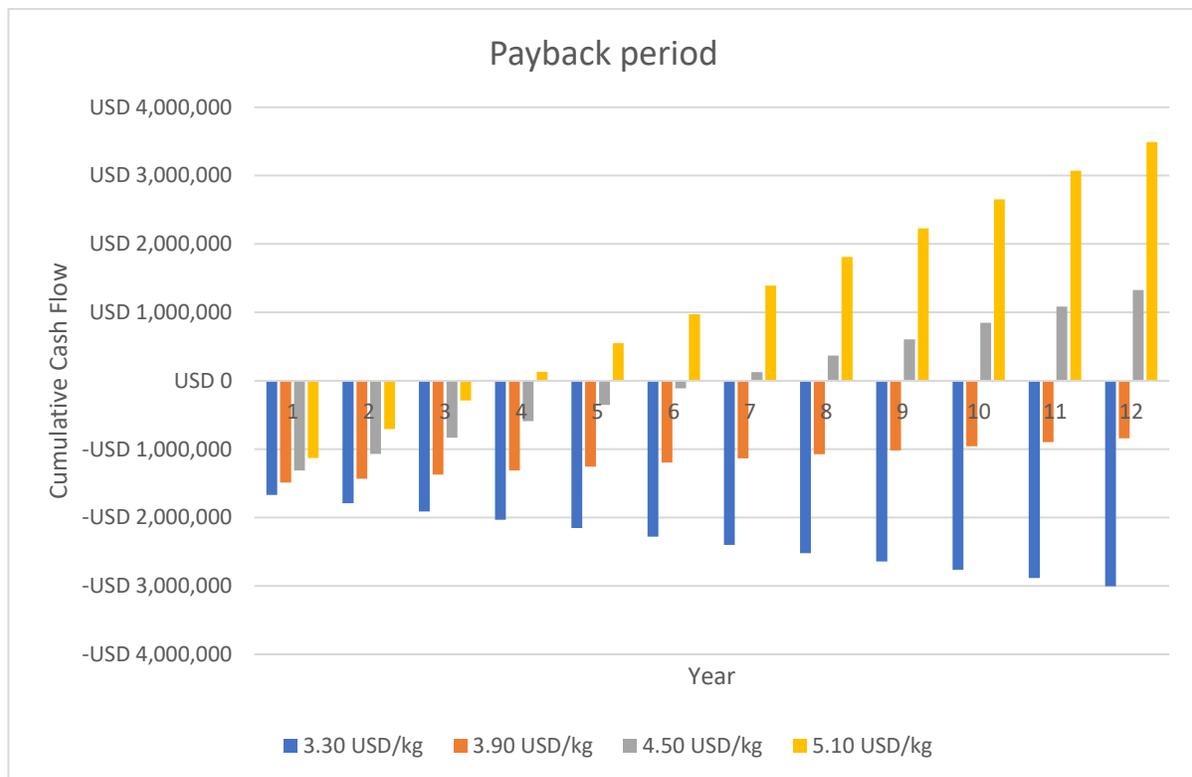


Figure 20 Payback times

5.1.6.4 Technical assessment

As mentioned in Chapter 2.5.4, the process chain based on Dual Fluidized Bed Gasification is at level TRL 5. The full process has been simulated and tested in laboratory conditions but there're a lot of challenges to overcome before stating full-scale production. However, the second technology, implemented into modelled plant, decreases the TRL furthermore – membrane separation.

The membrane separation is known technology but has not been yet used in such application, especially speaking about metallic alloyed membranes and nanocomposite membranes. Such modules only exist as laboratory-scale prototypes and are at TRL level of maximum 3. For better understanding of TRL, Figure 21 has been added to visualize the Technology Readiness Levels.

Overall the TRL is set to level 3 because this kind of set up is an object of current research and feasibility studies.

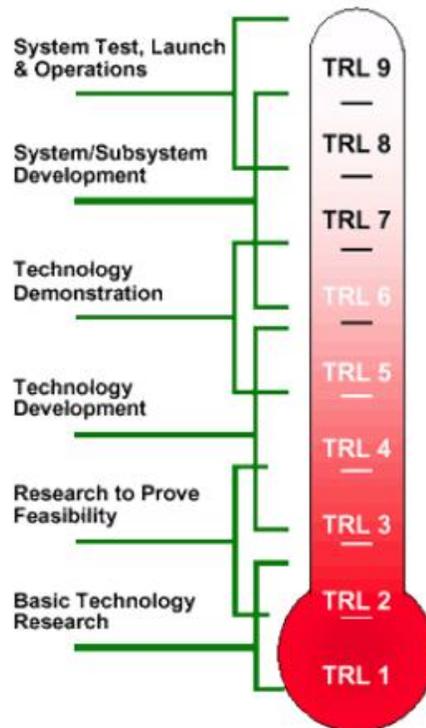


Figure 21 TRL Level diagram [44]

5.1.7 Sensitivity analysis, risk analysis of crucial technological segments

The sensitivity analysis is based on price variation of different resources (input materials), like wooden pellets, electricity etc and hydrogen selling price. These analyses were calculated at hydrogen selling price $C_{H_2} = 5.1 \text{ USD} \cdot \text{kg}^{-1}$ (except the hydrogen price sensitivity analysis). The Figure 22, Figure 23, Figure 24, Figure 25 show the dependency of Payback Time on change of certain prices.

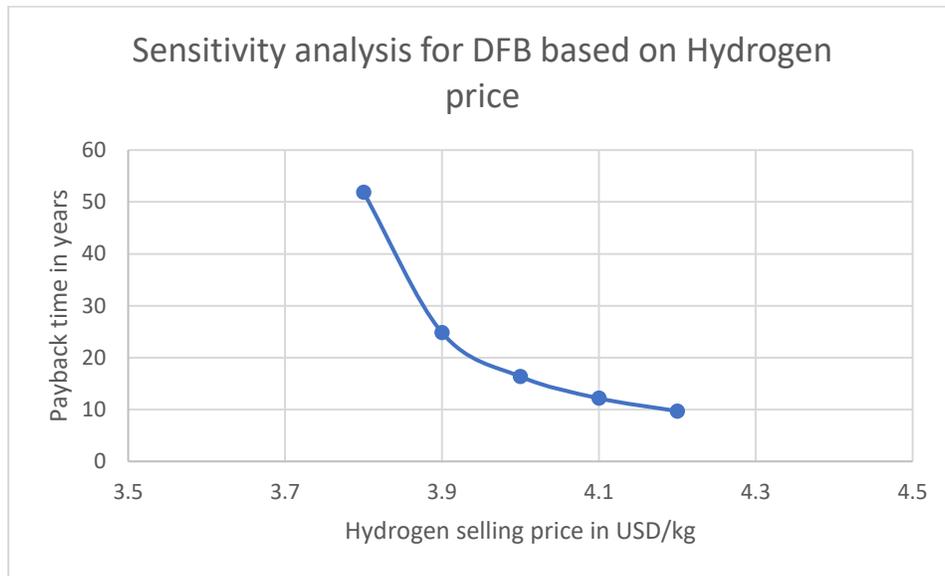


Figure 22 Hydrogen price sensitivity analysis

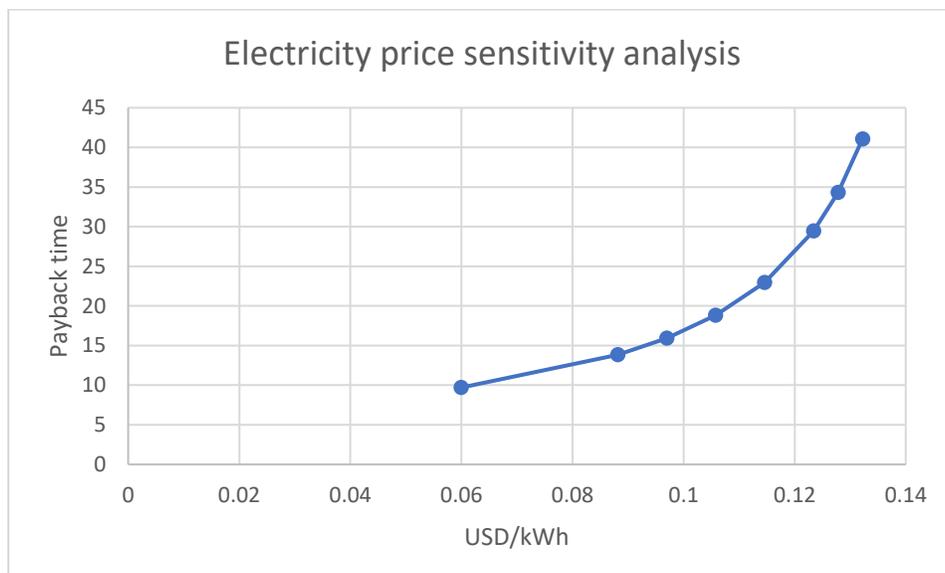


Figure 23 Electricity price sensitivity analysis

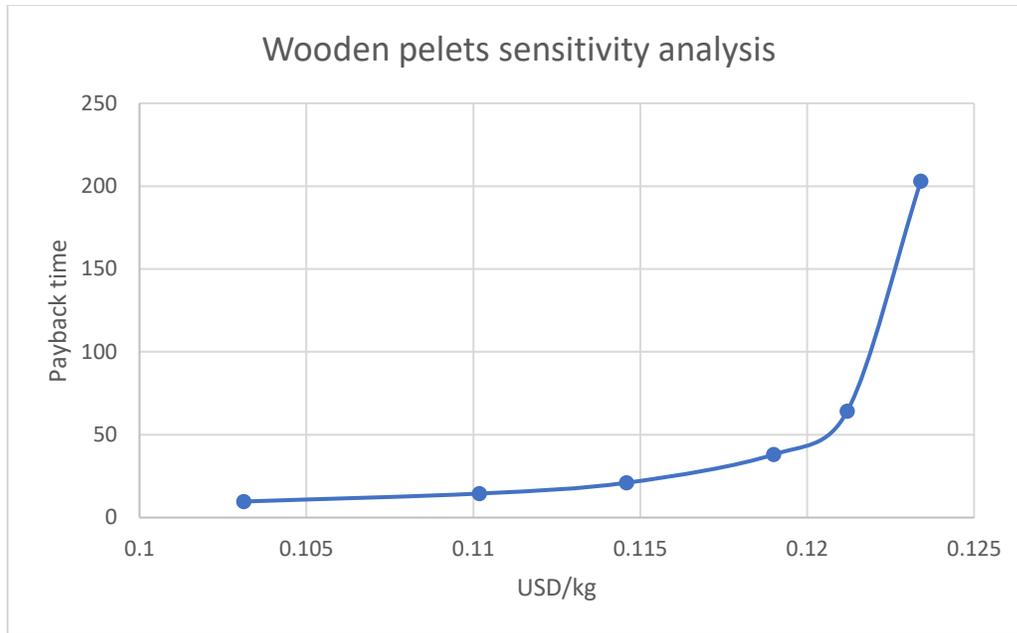


Figure 24 Wooden pellets price sensitivity analysis

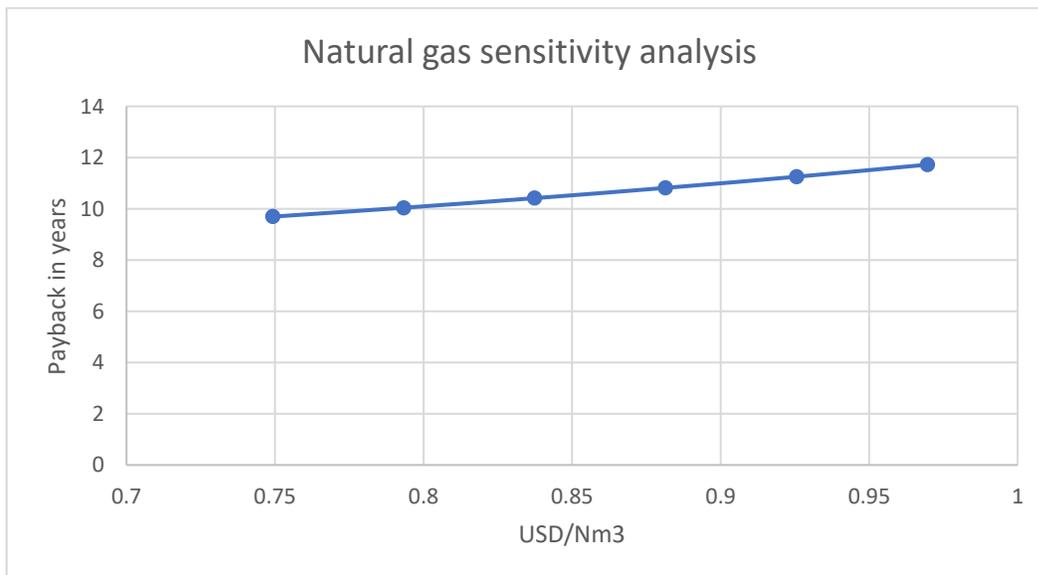


Figure 25 Natural gas price sensitivity analysis

It is clear, that even a slight price change of the wooden pellets has a huge, even disastrous effect on the project Cash Flow. This is seconded by the hydrogen selling price which has a strong effect on the Cash Flow too.

These two factors are the most critical that may ruin the project and are amongst the potential risks.

Economical aspect of risk assessment is one part. The other part is related to technical issues. Malfunction of equipment, force majeure, fire, explosion and other dangers are summarized in Table 29 and are given a certain degree of probability and impact degree (both degrees range from 1 to 5). These risks are additionally visualized in Figure 26.

Risk factor	Probability	Impact degree
Equipment malfunction	3	4
Dust explosion	2	4
Gas explosion	2	4
Fire	3	4
Force majeure	1	5
Environment contamination	2	4

Table 29 Risk assessment

Risk matrix			Impact				
			Minor	Moderate	Major	Critical	Fatal
			1	2	3	4	5
Probability	Very high	5					
	High	4					
	Moderate	3				Fire Equipment malfunction	
	Unlikely	2				Environment contamination Dust and gas explosion	
	Rare	1					Force majeure

Figure 26 Risk matrix

Safely enough there are no risks in the red zone, however the risks in the yellow zone are not to forget and have to be mitigated.

Fire and anti-explosion precautions are an absolute must at a plant that is producing explosive gases and is working with dusty material, like sawdust in this case. Organic dust can be very volatile and explosive. Additionally, wood pellets are excellent fire starters meaning additional fire precautions need to be installed at their storage. Force majeure is something one cannot predict but shall always be prepared if especially the plant is built in some hazardous area (frequent earthquakes etc.).

Of course, constant check of equipment is required to ensure flawless operation. Otherwise poor equipment condition can even lead to its destruction.

5.1.8 Discussion

Following table presents a sum-up of economic criteria calculated in previous chapters.

Comparison criteria	Thesis proposal – biomass gasification	Steam reforming [44]	Water electrolysis [38]
CAPEX [$\$ \cdot kg^{-1}$]	3.23	2.72	5.84
OPEX [$\$ \cdot kg^{-1}$]	3.00	3.38	5.6
Electricity [$kWh \cdot kg^{-1}$]	0.05	0.105	0.6
Payback time [years]	5 – 7	4 – 5	8 – 9
Lifetime [years]	Max. 5 for now	10 – 12	8 – 10
TRL [–]	3	9	9

Table 30 Economic criteria

For today, the biomass gasification loses significantly in the field of service life time and in the field of payback time. For comparison reasons, Figure and Figure show payback time of water electrolysis and steam reforming.

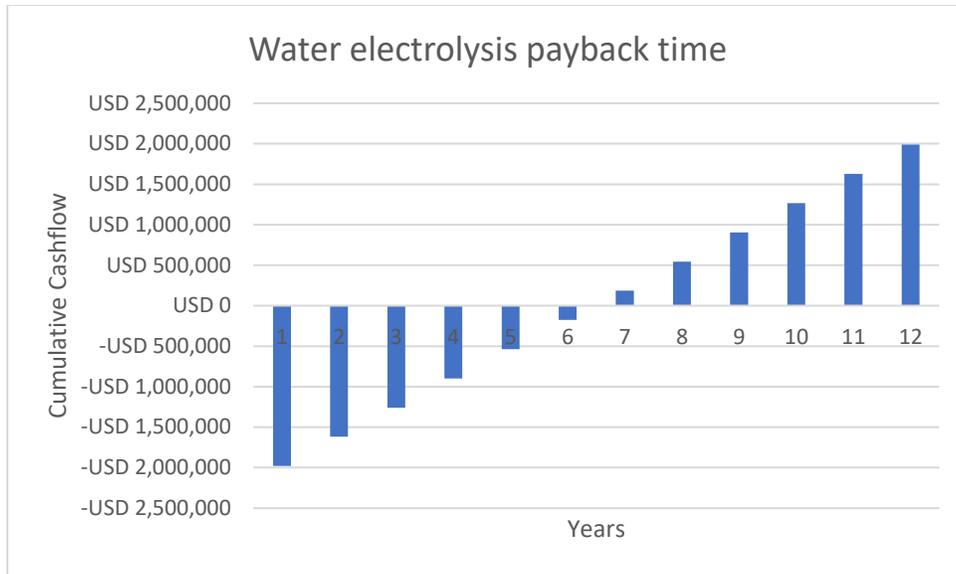


Figure 27 Payback time of water electrolysis

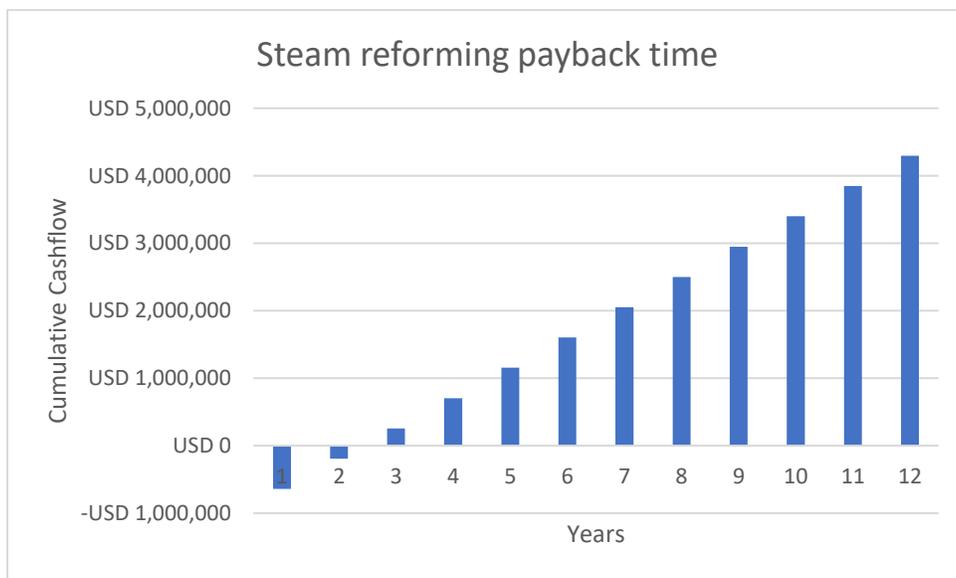


Figure 28 Payback time of steam reforming plant

The main problem here is that there're a lot of research challenges and other obstacles to be resolved before a full-scale plant can be built. All these challenges, advantages and disadvantages are presented in Table 31 and are discussed below the table.

Feature	Thesis proposal	Steam reforming [38]	Water electrolysis [39]
Advantages	Waste reducing “Eco friendly” Very promising Residual gas as side product	Known technology Good economic indexes Good service life	Known technology Cheaper each year
Disadvantages	Explosive substances Researched technology Low service life	Explosive substances Natural gas is not endless Complex separation process	Explosive substances 5 – 10 years till not expensive End quality and purity
Research challenges	Membranes and their modification	Decrease natural gas consumption Improve PSA	Decrease price Increase efficiency and purity

Table 31 Technology comparison

5.1.8.1 Discussion – residual gas

As mentioned in the Chapter 5.1.4.2 a potentially huge advantage of hydrogen production via biomass gasification is the residual gas side product which can be stored, sold and used instead of natural gas supply. If 20% of produced residual gas used as additional fuel in combustion chamber and the rest is stored in sold, the payback time rapidly decreases to 2 years as shown in figure. This modification assumes hydrogen selling price at $C_{H_2} = 3 \text{ USD} \cdot \text{kg}^{-1}$, which is $2 \text{ USD} \cdot \text{kg}^{-1}$ lower than in other cases. And even though the investment pays back in less than 2 years which is shown in the Figure 29.

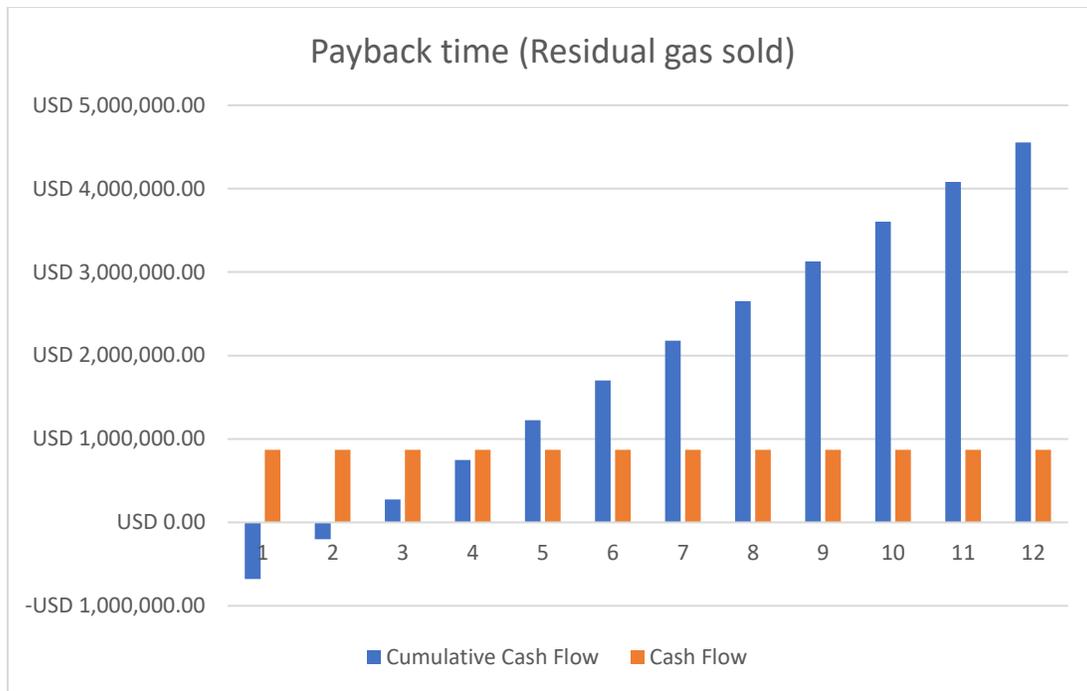


Figure 29 Payback time – Residual gas sold

5.1.8.2 Discussion – Research challenges and ecological impact

Firstly, the main research objects in this case are the membranes. Currently this technology is at its birth, so it is expensive and unresearched, but as said many times before, membranes are a very promising one. One of the challenges is to decrease dependency on gas purity so that membranes modules are near to immune to catalytic poison like sulphur and its oxides. And, needless to say, increase of the gas flux through the membrane is the priority research goal.

Secondly the gasification and membrane separation does not have a serious impact on environment, it is rather way around. These procedures are beneficial due to waste gasification, thus decreasing landfills and producing “eco-friendly fuel of the future”. Such development would open new field of research where other wastes could be researched and found suitable for this application.

5.1.8.3 Discussion – other

There're are some experiments conducted on combining state-of-art technology, like steam reforming, and membrane gas separation. By combining these two applications the humankind can step into a transition time until the gasification and membrane procedures are well researched and slowly but steadily shift towards them away from the old ones.

6 Conclusion

Hydrogen may shape the shape the future and be a solution to many environmental problems the humankind has caused so far. As far as this thesis goes, current prospects are that steam reforming will be slowly replaced with emerging technologies producing “biohydrogen”, amongst which are especially waste gasification and biotechnological production based on fermentation. Water electrolysis is not to forget too but to produce “eco-clean” hydrogen the electrolysis plant would need to be supplied with “green energy” which is not the case nowadays. Additionally, the electrolysis installation and operation costs are uncompetitive even against the biomass gasification and are predicted to fall within the next 7 – 10 years.

In this thesis a model of a technology block producing hydrogen from wooden pellets (originally waste wood) has been described. Preliminary results show that if research is conducted on the mentioned technology (DFB System and membranes) than it is possible to see a pilot of the same concept in the next 5 years. Until that time scientists have to overcome numerous research obstacles. Following list sums up the goals and achieved results:

- Acquiring necessary theoretical background for gasification and membrane
- Search for research applications and pilot plants
- Creation of a model for hydrogen producing plant based on acquired theoretical background
- Assessment of the model based on current economic and technical (TRL level) situation

7 Reference page

- [1] Czech Statistical Office. Czech Statistical Office [online]. [cited 2019-03-23]. Available at: <https://www.czso.cz/csu/czso/home>
- [2] Tab. 12 Produkce podnikových odpadů podle skupin Katalogu odpadů. Czech Statistical Office [online]. [cited 2019-03-23]. Available at: <https://www.czso.cz/documents/10180/61546956/2800201812.pdf/fc0b628a-0c65-44db-9d69-4216ceae0edd?version=1.2>
- [3] FARACA, Giorgia, Alessio BOLDRIN a Thomas ASTRUP. Resource quality of wood waste: The importance of physical and chemical impurities in wood waste for recycling. Waste Management [online]. 2019, **2019**(87), 135-147 [cited 2019-03-23]. DOI: <https://doi.org/10.1016/j.wasman.2019.02.005>. Available at: <https://www.sciencedirect.com/science/article/pii/S0956053X19300765>
- [4] Eurostat. Statistics Explained. Eurostat. Statistics Explained. [online]. [cited 2019-03-23]. Available at: https://ec.europa.eu/eurostat/statistics-explained/index.php/Main_Page
- [5] Forestry in the EU and the world. A statistical portrait. Eurostat. [online]. [cited 2019-03-26]. Available at: <https://ec.europa.eu/eurostat/documents/3217494/5733109/KS-31-11-137-EN.PDF/cbd2d7d5-0cfa-4960-b5d3-02eb065abba5>
- [6] ALAMIA, Alberto, Anton LARSSON, Claes BREITHOLTZ and Henrik THUNMAN. *Resource quality of wood waste: The importance of physical and chemical impurities in wood waste for recycling*. International Journal of Energy Research [online]. 2017, **2017**(41), 2001-2019 [cited 2019-03-23]. DOI: <https://doi.org/10.1002/er.3758>. Available at: <https://onlinelibrary.wiley.com/doi/full/10.1002/er.3758>
- [7] Steam Methane Reforming - Hydrogen Production. Air Liquide - Engineering & Construction [online]. [cited 2019-04-10]. Available at: <https://www.engineering-airliquide.com/steam-methane-reforming-hydrogen-production>
- [8] Using an Adiabatic Pre-reformer. Industrial Efficiency - Technology Database [online]. [cited 2019-04-10]. Available at: <http://ietd.iipnetwork.org/content/using-adiabatic-pre-reformer>

- [9] SEHESTED, Jens. *Four challenges for nickel steam-reforming catalysts*. Catalysis today [online]. 2006, **2006**(1-2), 103-110 [cited 2019-03-20]. DOI: <https://doi.org/10.1016/j.cattod.2005.10.002>. Available at: <https://www.sciencedirect-com.ezproxy.techlib.cz/science/article/pii/S092058610500708X>
- [10] Hydrogen Production: Electrolysis. Energy.gov [online]. [cited 2019-03-15]. Available at: <https://www.energy.gov/eere/fuelcells/hydrogen-production-electrolysis>
- [11] STADLER, Paul. *Cost evaluation of large scale hydrogen production for the aviation industry*. Ecole Polytechnique Federale de Lausanne, 2014. Master Semester Project. Ecole Polytechnique Federale de Lausanne. Supervisors: Priscilla Caliandro, Claudio Leonardi, Dr. Jan Van Herle. Available at: <https://clipair.epfl.ch/wp-content/uploads/2018/08/FullReportMA3.pdf>
- [12] SPATH, Pamela L. and Margaret K. MANN. *Life Cycle Assessment of Hydrogen Production via Natural Gas Steam Reforming* [online]. Bechtel, 2001 [cited 2019-03-15]. Technical Report. National Renewable Energy Laboratory. Available at: <https://www1.eere.energy.gov/hydrogenandfuelcells/pdfs/27637.pdf>.
- [13] BERTUCCIOLI, Luca, Alvin CHAN, David HART, Franz LEHNER, Ben MADDEN and Eleanor STANDEN. Study on development of water electrolysis in the EU. In: *Fuel Cells and Hydrogen* [online]. 2014 [cited 2019-03-15]. Available at: [https://www.fch.europa.eu/sites/default/files/FCHJUElectrolysisStudy_FullReport%20\(ID%20199214\).pdf](https://www.fch.europa.eu/sites/default/files/FCHJUElectrolysisStudy_FullReport%20(ID%20199214).pdf)
- [14] JACOBS, Jeffrey. *Economic Modeling of Cost Effective Hydrogen Production From Water Electrolysis by Utilizing Iceland's Regulating Power Market* [online]. Reykjavík, 2016 [cited 2019-05-23]. Master Thesis. Reykjavík University. Available at: <https://skemman.is/bitstream/1946/23812/1/Jeffrey%20Jacobs.pdf>
- [15] SANTOS, Diogo M.F., Cesar A.C. SEQUEIRA and Jose L. FIGUEIREDO. HYDROGEN PRODUCTION BY ALKALINE WATER ELECTROLYSIS. *Quinn. Nova* [online]. 2013, **2013**(8), 1176-1193 [cited 2019-03-20]. ISSN 0100-4042. Available at: <http://www.scielo.br/pdf/qn/v36n8/v36n8a17.pdf>

- [16] MARSHALL, A., B. BORRESSEN, G. HAGEN, M. TSYPKIN and R. TUNOLD. Hydrogen production by advanced proton exchange membrane (PEM) water electrolyzers— Reduced energy consumption by improved electrocatalysis. *Energy* [online]. 2007, **2007**(32), 431-436 [cited 2019-03-20]. DOI: <https://doi.org/10.1016/j.energy.2006.07.014>. Available at: <https://www.sciencedirect.com/science/article/pii/S036054420600199X>
- [17] MORES, Patricia L., Ana M. ARIAS, Nicolas J. SCENNA, Jose A. CABALLERO, Sergio F. MUSSATI and Miguel C. MUSSATI. Membrane-Based Processes: Optimization of Hydrogen Separation by Minimization of Power, Membrane Area, and Cost. *Processes* [online]. 2018, **2018**, **6**(221) [cited 2019-03-21]. DOI: [doi:10.3390/pr6110221](https://doi.org/10.3390/pr6110221). Available at: <https://www.mdpi.com/2227-9717/6/11/221/pdf>
- [18] LI, Wenping, Zhongwei CAO, Xuefeng ZHU and Weishen YANG. Effects of membrane thickness and structural type on the hydrogen separation performance of oxygen-permeable membrane reactors. *Journal of Membrane Science* [online]. 2019, (573), 370-376 [cited 2019-03-23]. DOI: <https://doi.org/10.1016/j.memsci.2018.12.005>. Available at: <https://www.sciencedirect.com/science/article/pii/S0376738818329120>
- [19] YU, Shuwen, Shichun LI, Yu LIU, Sheng CUI a Xiaodong SHEN. High-performance microporous polymer membranes prepared by interfacial polymerization for gas separation. *Journal of Membrane Science* [online]. 2019, (573), 425-438 [cited 2019-03-23]. DOI: <https://doi.org/10.1016/j.memsci.2018.12.029>. Available at: <https://www.sciencedirect.com/science/article/pii/S0376738818319173>
- [20] ZHAO, Yayun, Dechuan ZHAO, Chunlong KONG, Feng ZHOU, Tengyao JIANG and Liang CHEN. Design of thin and tubular MOFs-polymer mixed matrix membranes for highly selective separation of H₂ and CO₂. *Separation and Purification Technology* [online]. 2019, **2019**, (220), 197-205 [cited 2019-03-24]. DOI: <https://doi.org/10.1016/j.seppur.2019.03.037>. Available at: <https://www.sciencedirect.com/science/article/pii/S138358661832286X>

- [21] HASHIM, Siti Salwa, Mahendra Rao SOMALU, Kee Shyuan LOH, Shaomin LIU and Jaka SUNARSO. Perovskite-based proton conducting membranes for hydrogen separation: A review. *International Journal of Hydrogen Energy*[online]. 2018, **43**(32), 15281-15305 [cited 2019-03-24]. DOI: <https://doi.org/10.1016/j.ijhydene.2018.06.045>. Available at: <https://www.sciencedirect.com/science/article/pii/S0360319918318573>
- [22] WANG, Y., X. MA, B.S. GHANEM, F. ALGHUNAIMI, I. PINNAU and Y. HAN. Polymers of intrinsic microporosity for energy-intensive membrane-based gas separations. *Materials Today Nano* [online]. 2018, **3**, 69-95 [cited 2019-03-24]. DOI: <https://doi.org/10.1016/j.mtnano.2018.11.003>. Available at: <https://www.sciencedirect.com/science/article/pii/S2588842018301597>
- [23] DI MARCOBERARDINO, Gioele, Stefano FORESTI, Marco BINOTTI and Giampaolo MANZOLINI. Potentiality of a biogas membrane reformer for decentralized hydrogen production. *Chemical Engineering and Processing - Process Intensification* [online]. 2018, (129), 131-141 [cited 2019-03-25]. DOI: <https://doi.org/10.1016/j.cep.2018.04.023>. Available at: <https://www.sciencedirect.com/science/article/pii/S0255270117307857>
- [24] FASOLIN, Stefano, Simona BARISON, Stefano BOLDRINI, et al. Hydrogen separation by thin vanadium-based multi-layered membranes. *International Journal of Hydrogen Energy* [online]. 2018, **6**(43), 3235-3243 [cited 2019-03-25]. DOI: <https://doi.org/10.1016/j.ijhydene.2017.12.148>. Available at: <https://www.sciencedirect.com/science/article/pii/S0360319917348693>
- [25] BESPALKO, Y., V. SADYKOV, N. EREMEEV, et al. *Synthesis of tungstates/Ni_{0.5}Cu_{0.5}O nanocomposite materials for hydrogen separation cermet membranes* [online]. 1263-1274 [cited 2019-03-25]. DOI: <https://doi.org/10.1016/j.compstruct.2018.06.004>. Available at: <https://www.sciencedirect.com/science/article/pii/S0263822318302447>
- [26] Balachandran, U.; et al. *Dense cermet membranes for hydrogen separation*. [Online] **2014**, 54-49. [cited 2019-03-25]. Available at: <https://www.sciencedirect.com/science/article/pii/S1383586613005777>

- [27] ZEYNALI, R., BASILE, A., et al. Experimental evaluation of graphene oxide/TiO₂-alumina nanocomposite membranes performance for hydrogen separation. *International Journal of Hydrogen Energy* [online]. 2019 [cited 2019-04-10]. Available at: <https://www.sciencedirect.com/science/article/pii/S0360319919308870>
- [28] PETERS, T., et al. Palladium (Pd) Membranes as Key Enabling Technology for Pre-combustion CO₂ Capture and Hydrogen Production. *Energy Procedia* [online]. 2017, vol. 114 [cited 2019-04-10], p. 37–45. Available at: <https://www.sciencedirect.com/science/article/pii/S1876610217313152/>
- [29] YIN, H., YIP, A. C., et al. A Review on the Production and Purification of Biomass-Derived Hydrogen Using Emerging Membrane Technologies. *catalysts* [online]. 2017, vol. 7, no. 10 [cited 2019-04-10]. Available from www.mdpi.com/2073-4344/7/10/297/pdf-vor
- [30] REED, Tom. How Gasification Works. *All Power Labs: Carbon Negative Power & Products* [online]. [cited 2019-04-11]. Available at: <http://www.allpowerlabs.com/gasification-explained>
- [31] MILIONI, Andrea. Gasification Process. *Oil & Gas Portal* [online]. [cited 2019-04-11]. Available at: <http://www.oil-gasportal.com/gasification-process/>
- [32] LOHA, Ch., et al. Advances in mathematical modeling of fluidized bed gasification. *Renewable and Sustainable Energy Reviews* [online]. 2014, vol. 40 [cited 2019-04-12], p. 688–715. Available from <https://www.sciencedirect.com/science/article/pii/S1364032114006595>
- [33] BELGIORNO, V., DE FEO, G., DELLA ROCCA, C., NAPOLI, R. M. A., et al. Energy from Gasification of Solid Wastes. *Waste Management* [online]. 2003 [cited 2019-04-12], p. 1–15. Available at: https://www.researchgate.net/publication/10868430_Energy_from_Gasification_of_Solid_Wastes
- [34] WILK., V., SCHMID, J., HOFBAUER, H., et al. Influence of fuel feeding positions on gasification in dual fluidized bed gasifiers. *Biomass & Bioenergy* [online]. 2013, vol. 54 [cited 2019-04-13], p. 46–58. Available from https://www.bioenergy2020.eu/files/publications/pdf/Feeding_position.pdf

- [35] WHAT A WASTE 2.0: Trends in Solid Waste Management. *The World Bank* [online]. [cited 2019-04-13]. Available at: <http://datatopics.worldbank.org/what-a-waste/trends-in-solid-waste-management.html>
- [36] A.SCHOLES, Colin, Kathryn H.SMITH, Sandra E.KENTISH and Geoff W.STEVENS. CO₂ capture from pre-combustion processes: Strategies for membrane gas separation. *International Journal of Greenhouse Gas Control* [online]. 2010, 26.5.2010, 4(5), 739-755 [cited 2019-04-13]. Available at <https://www.sciencedirect.com/science/article/abs/pii/S1750583610000472>
- [37] PECATE, Sebastien, Sid AHMED KESSAS, Mathieu MORIN and Mehrdji HEMATI. Beech wood gasification in a dense and fast internally circulating fluidized bed. *Fuel* [online]. 2019, 2018, 2018(236), 554-573 [cited 2019-04-13]. Available at: <https://www.sciencedirect.com/science/article/pii/S0016236118315643>
- [38] THOMAS, Denis. COST REDUCTION POTENTIAL FOR ELECTROLYSER TECHNOLOGY. *European powertogas* [online]. 2018 [cit. 2019-04-15]. Available at: <http://europeanpowertogas.com/wp-content/uploads/2018/06/20180619-Hydrogenics-EU-P2G-Platform-for-distribution.pdf>
- [39] SCHMIDT, O., et al. Future cost and performance of water electrolysis: An expert elicitation study. *International Journal of Hydrogen Energy* [online]. 2017, vol. 42, no. 52 [cited 2019-04-16], p. 30470–30492. Available at: <https://www.sciencedirect.com/science/article/pii/S0360319917339435>
- [40] BINDER, M., KRAUSSLER, M., KUBA, M., LUISSER, M., et al. Hydrogen from biomass gasification. *IEA Bioenergy* [online]. 2018, vol. 33 [cited 2019-04-15], p. 1–85. Available at: <http://task33.ieabioenergy.com/download.php?file=files/file/publications/Hydrogen/Wasserstoffstudie-IEA%20final.pdf>
- [41] *VDI heat atlas*. 2nd ed. New York: Springer, 2010. ISBN 978-3-540-77876-9.
- [42] Cost Estimation. *Matches* [online]. [cited 2019-05-10]. Available at: <http://matche.com/equipcost/Exchanger.html>

- [43] Cost Estimation. *The University of Oklahoma* [online]. [cit. 2019-05-10]. Available at: <http://www.ou.edu/class/che-design/design%201-2013/econ-2.pdf>
- [44] MAYERS, Duane B., Gregory D. ARIFF, Brian D. JAMES, John S. LETTOW, C.E. (Sandy) THOMAS a Reed C. KUHN. Cost and Performance Comparison Of Stationary Hydrogen Fueling Appliances. *National Renewable Energy Laboratory* [online]. [cited 2019-05-10]. Available at: <http://www.ou.edu/class/che-design/design%201-2013/econ-2.pdf>

8 List of symbols and abbreviations

Symbol	Unit	Description
P	[Barrer]	Gas permeability
D	$[cm^2 \cdot s^{-1}]$	Diffusion coefficient
S	$[cm_{Standard}^3 \cdot cm^{-3} \cdot cmHg^{-1}]$	Solubility coefficient
$\alpha_{A,B}$	[-]	Membrane selectivity
$P_{A,B}$	[Barrer]	Individual gas permeability
DFB	[-]	Dual Fluidized Bed
Q_{TOT}	[J]	Total required heat energy
Q_W	[J]	Heat energy to heat water to 100 °C
Q_L	[J]	Latent heat
Q_S	[J]	Heat to superheat steam
c_p	$[J \cdot kg^{-1} \cdot K^{-1}]$	Water heat capacity
ΔT_B	[K]	Temperature to boiling point
ΔT_{SH}	[K]	Superheat temperature
a	$[J \cdot kg^{-1} \cdot K^{-1}]$	Heat to superheat steam
\dot{Q}	[W]	Heat duty
\dot{m}	$[kg \cdot s^{-1}]$	Mass flux
c^V	[%]	Volumetric concentration
\dot{V}	$[Nm^3 \cdot s^{-1}]$	Volumetric flux
LVH_{NG}	$[MJ \cdot Nm^{-3}]$	Natural gas Low Heating Value
$\dot{V}_{TO\ BURN}$	$[Nm^3 \cdot s^{-1}]$	Vol. flux to burn
LVH_{SYNGAS}	$[MJ \cdot Nm^{-3}]$	Syngas low heating value
\dot{Q}	[MW]	Required heat duty
$c_{CH_4}^V$	[%]	Methane volumetric concentration

c_{CO}^V	[%]	Carbon monoxide volumetric concentration
$\dot{V}_{CO\ SYNGAS}$	$[Nm^3 \cdot s^{-1}]$	Carbon monoxide volumetric flux in syngas
$r_{CO\ SYNGAS}$	[%]	Ratio
$\dot{V}_{SYNGAS\ STORAGE}$	$[Nm^3 \cdot s^{-1}]$	Stored syngas volumetric flux
P	$[W]$	Compressor power duty
P_{ad}	$[W]$	Compressor adiabatic duty
η_{ad}	[%]	Adiabatic efficiency
κ	[–]	Adiabatic index
p_1	$[kPa]$	Inlet pressure
p_2	$[kPa]$	Outlet pressure
T_1	$[K]$	Inlet temperature
p_{STP}	$[kPa]$	Standard pressure
T_{STP}	$[K]$	Standard temperature
\dot{V}_1	$[m^3 \cdot s^{-1}]$	Syngas volumetric flux
\dot{V}_{AIR}	$[m^3 \cdot s^{-1}]$	Air volumetric flux
$\dot{V}_{AIR\ FEED}$	$[Nm^3 \cdot s^{-1}]$	Air volumetric flux
P_{BLOWER}	$[W]$	Blower power duty
OpEx	$[USD \cdot kg_{Produced\ Hydrogen}^{-1}]$	Annual operation expenses
k	$[W \cdot m^{-2} \cdot K^{-1}]$	Heat transfer coefficient
S	$[m^2]$	Effective heat transfer area
ΔT	$[K]$	Heat exchanger temperature gradient
CCF	$[USD \cdot year^{-1}]$	Annual cumulative cashflow
INV	$[USD]$	Project investment
t	$[year]$	Years
C_{H_2}	$[USD \cdot kg_{Hydrogen}^{-1}]$	Hydrogen price
Tax	[%]	State tax
CapEx	$[USD \cdot kg_{Produced\ Hydrogen}^{-1}]$	Capital expenses
TRL	[–]	Technology readiness level

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Annex 1 – EXCEL Model: Main sheets

Annex 2 – PFD Scheme

Steam calc

SYNGAS		CP at 850 [kj/(kg*k)]	CP at 400 [kj/(kg*k)]	Heat capacity % [kj/(kg*k)]	Rho at 850	Rho at 400	Total rho	Mass flux [kg/h]	Q [kW]
CO		0.13	1.22	1.1	0.1508	0.566395	0.856	0.092456	
CO2		0.289	1.27	1.102	0.342754	0.5	0.8144	0.189931	
CH4		0.064	4.475	3.432	0.253024	0.314	0.41	0.023168	
H2		0.494	15.14248	14.702295	7.371659425	0.242	0.3454	0.145088	
H2O	-								
Ethane		0.023	3.846	3.126	0.080178	0.3616	0.5167	0.0101	
Total mean CP					8.198415425		0.460743	598.9655425	508.6484 0.649957
FLUEGAS		CP at 1000 [kj/(kg*k)]	CP at 100 [kj/(kg*k)]	Heat capacity % [kj/(kg*k)]	Rho at 1000	Rho at 100	Total rho	Mass flux [kg/h]	
O2		0.0435	1.115	0.934	0.04456575	0.3848	1.069	0.03162	
N2		0.569	1.204	1.042	0.638987	0.317	0.9027	0.347005	
CO		0.0002	1.22	1.045	0.0002265	0.314	0.3433	6.57E-05	
CO2		0.1153	1.27	0.918	0.1261382	0.5	0.3063	0.046483	
CH4	-								
H2	-								
H2O		0.272	2.425	1.89	0.58684	0.202	0.59	0.107712	
CxHy	-								
Tar	-								
Total mean CP					1.39675745		0.532886	1260.807625	273.9399 0.350043
Water									
dT		102							
cp water		4200 j/(kg*K)							
latent heat water		2300000 j/(kg)							
to heat steam		2000 J(kg*K)							
dt to heat steam		223 K							
Total steam		1000 kg/h							
Total heat required		881777.778 W							
		881.778 kW							
Total heat covered		782.588 kW							
Ratio of recovered heat									
against heat for steam									
from outside source		0.888							
Steam produced in process		887.512 kg/h							
Steam produced in process		0.247 kg/s							
Steam from outside		112.488 kg/h							
Steam from outside		0.031246679 kg/s							

Compressor calc

<i>Kappa</i>	1.29144	
<i>CO</i>	1.32	
<i>CO2</i>	1.2	
<i>CH4</i>	1.15	
<i>H2</i>	1.36	
<i>H2O</i>	1.25	
<i>Ethane</i>	1.2	
<i>p1</i>	120	kPa
<i>p2</i>	670	kPa
<i>R</i>	8.314	
<i>t_1</i>	217.2905866	
<i>T_1</i>	490.4405866	
<i>Vst_1</i>	0.540283425	m3/s
<i>Vst_2</i>	0.142652695	m3/s
<i>T_2</i>	723	K
<i>M_smesi</i>	19.058	
<i>R</i>	436.2472453	
<i>Prikon</i>	194.6155087	kW

Air blower

<i>p1</i>	101.325	kPa
<i>p2</i>	120	kPa
<i>kappa air</i>	1.4	
<i>T_air</i>	293.15	
<i>Vst</i>	0.329524419	m3/s
<i>Prikon</i>	8.266799366	kW

Total electricity consumption

<i>Syngas compressor</i>	194.6155087
<i>Blower</i>	8.266799366
<i>Additional equipment</i>	20.2882308
<i>Total</i>	223.1705388
	225

<i>Production</i>	1.3MW Hydrogen	<i>Unit</i>
Operating time	7500	hours/year
Hourly production	53.43104	kg/h
Yearly production	400732.8	kg/year
<i>Raw materials</i>		
Olivine	89.7	kg/h
Natural gas	16.71641791	Nm3/h
Wooden pellets	1000	kg/h
<i>Energies</i>		
kWh per biomass feed electricity	0.225	kWh compressor/kg biomass feed
<i>Cost per kg of H2</i>		
Natural gas	USD 93,933.15	USD/year
Olivine	USD 121,095.00	USD/year
Wooden pellets	USD 773,468.49	USD/year
Steam	USD 29,745.63	USD/year
Water	USD 19,969.02	USD/year
Electricity	USD 101,145.88	USD/year
<i>Personal costs</i>		
Operators	3	
Wage per operator	21419.12737	USD/year
Personal costs	2831.969242	USD/year
<i>Total</i>		
Total	USD 1,142,189.14	USD/year
Selling price to NPV = 0	USD 2.85	USD/kg
Investment	USD 1,548,187.57	USD

<i>Production</i>	1.3MW Hydrogen	<i>Unit</i>
Operating time	7500	hours/year
Hourly production H2	53.43104	kg/h
Hourly production Biogas	526.24	Nm3/h
Yearly production H2	400732.8	kg/year
Yearly production Biogas	3946800	Nm3/h
Raw materials		
<i>Raw materials</i>		<i>Unit</i>
Olivine	89.7	kg/h
Natural gas	16.71641791	Nm3/h
Wooden pellets	1000	kg/h
Energies		
<i>Energies</i>		<i>Unit</i>
kWh per biomass feed electricity	0.225	kWh compressor/kg biomass feed
Cost per kg of H2		
<i>Cost per kg of H2</i>		
Natural gas	USD 93,933.15	USD/year
Olivine	USD 121,095.00	USD/year
Wooden pellets	USD 773,468.49	USD/year
Steam	USD 29,745.63	USD/year
Water	USD 19,969.02	USD/year
Electricity	USD 101,145.88	USD/year
Personal costs		
<i>Personal costs</i>		<i>Unit</i>
Operators	3	
Wage per operator	USD 21,419.13	USD/year
Personal costs	USD 2,831.97	USD/year
Total		
<i>Total</i>		<i>Unit</i>
Total	USD 1,142,189.14	USD/year
Selling price to NPV = 0 Hydrogen	USD 2.85	USD/kg
Biogas selling price	USD 0.22	Kč/kg
Investment		
Investment	USD 1,548,187.57	Kč
REV	USD 2,011,911.48	Kč
Tax	19.00%	%

