UNIfHY

UNIQUE gasifier for hydrogen production

GA: 299732 (SP1-JTI-FCH.2011.2.3 Biomass-to-hydrogen (BTH) thermal conversion process)

Deliverable 5.3

Techno-economic analysis of UNIFHY hydrogen production system

(Dissemination Level: PU-Public)

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<table>
<thead>
<tr>
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<th>Name</th>
<th>Role</th>
</tr>
</thead>
<tbody>
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</tr>
<tr>
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</tr>
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<td>Technical UNFHY analysis</td>
</tr>
<tr>
<td>USGM</td>
<td>Roberto Tascioni</td>
<td>Conclusion</td>
</tr>
</tbody>
</table>

Issues

<table>
<thead>
<tr>
<th>Issue</th>
<th>Date</th>
<th>Sent by</th>
<th>To</th>
<th>Page</th>
<th>Modifications</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>22/03/2016</td>
<td>Sara Rajabi Hamedani</td>
<td>Enrico Bocci Donatella Barisano Marco Rep</td>
<td>Final Review</td>
<td></td>
</tr>
</tbody>
</table>
INDEX

1 Objectives of the deliverable ........................................................................................................2
2 Technical analysis..........................................................................................................................4
  2.1 System data: gasifier types, biomass, lifetime and availability ............................................. 4
  2.2 Summary of the main components technical data .................................................................. 6
    2.2.1 Gasifier 1000kWth............................................................................................................. 7
    2.2.2 Gasifier 100kWth............................................................................................................... 9
    2.2.3 Filter candles......................................................................................................................10
    2.2.4 PPS.....................................................................................................................................11
3 Economic analysis.......................................................................................................................14
  3.1 General economic consideration on biomass plants ...............................................................14
  3.2 CapEX......................................................................................................................................14
    3.2.1 Gasifier 1000 kWth........................................................................................................... 14
    3.2.2 Gasifier 100 kWth............................................................................................................ 15
    3.2.3 Gasifier plants cost comparison........................................................................................ 17
    3.2.4 Filter candles......................................................................................................................18
    3.2.5 PPS.....................................................................................................................................19
  3.3 Engineering costs & OpEx.......................................................................................................20
4 Comparison of IH and Oxygen UNIfHY plants at 1 MWth size .................................................22
5 UNIfHY under different scenarios .............................................................................................23
6 Conclusion...................................................................................................................................27
Reference.........................................................................................................................................38
1 Objectives of the deliverable

Following the topic and project goals, this deliverable aims to evaluate techno and economic performances of the UNIfHY system regarding the distributed hydrogen generation via the project results and in particular the Deliverables 2.1[1], 2.2[2], 2.3[3], 2.4[4], 2.5[5], 2.6[6], 3.1[7], 3.2[8], 3.3[9], 4.1[10], 4.2[11], 4.3[12], 5.1[13], 5.2[14], 5.4[15], 6.1[16], 6.3[17], 6.4[18].

The “alternative” way (the use of CO₂ capture sorbents in order to increase efficiency and reduce the flow to be purified: D2.7 and D6.2) has not taken into account owing to the only simulative and lab scale activities. Thus the techno-economic analysis refers to a plant configuration that consists of the following main components: an indirectly heated steam or a steam/oxygen gasifier with catalytic filter candles in the freeboard, a ZnO and a WGS reactor, a PSA system. Different plant configurations (as DOE [19] and FCH [20] reports), e.g. cold gas conditioning via biodiesel/FAME scrubber (as Gussing plant), additional methane reformer (as FCH report) or sorbents (inside the indirectly heated gasifier, as in AER-GAS project, or in following separated reactors) or other CO₂ or H₂ removal systems are not taken into consideration. Nevertheless, in the conclusion chapter, the UNIfHY configuration is compared to the different FCH report configuration.

This report quotes the main results of the deliverables in order to assess the technical performance of the two UNIfHY prototypes with the optimisations to be in required for industrial configuration (e.g. the Portable Purification System, realized for a gasifier of about 200 kWₜₜ, is scaled in order to fill the exact flow of the two gasifiers, 100 kWₜₜ and 1000 kWₜₜ). Therefore, the technical performances such as efficiency, reliability, maintenance, etc. at component level (100 kWₜₜ and 1000 kWₜₜ gasifiers, filter candles, PSA, WGS) and at system level (UNIfHY 100 and 1000) were assessed.

The economic performances have been assessed via the analysis of the CAPEX and OPEX for each component of the two different UNIfHY systems: UNIfHY 100 plant (based on indirectly heated steam/air fluidised bed gasification) and UNIfHY 1000 plant (based on steam/oxygen fluidised bed gasification).

In order to compare the two technologies a study at same size (1000 kWth) and same operative parameters (near to the experimental ones: 800 °C gasification temperature, S/B 0.5) has been added.
Techno-economic analysis of UNIFHY hydrogen production system

In addition, an economic and technical evaluation of the best technology at the best operating parameters (i.e. 850 °C gasification temperature, S/B 1.5) for different sizes (100 kW\textsubscript{th}, 1 MW\textsubscript{th}, 10 MW\textsubscript{th}) has been done. The sizes have been chosen following the distributed hydrogen generation sizes, topic and project scope. These varies from a hydrogen production of few kg/h (50-500 H\textsubscript{2}kg/day) to thousands kg/h (1-10 H\textsubscript{2}t/day) thus from about 100 kW\textsubscript{th} thermal input of biomass to about 10 MW\textsubscript{th}. Nevertheless, it has to be mentioned that DOE/NREL consider this thermal BTH (Biomass To Hydrogen) technology only for central production (i.e. for size > 30 t/day, i.e. greater than about 50 MW\textsubscript{th} of biomass input), thus the comparison with NREL data, have to take into account the different size. Furthermore, FCH divides the distributed hydrogen generation into local (0.4-4 t/day) and semi local (4-20 t/day), meanwhile central is for size > 20 t/day (not >30 t/day as NREL/DOE). Taking into account that to really exploit the biomass energy potential, small scale power plants have to be developed to follow the low energy density and perishability of this fuel (not to mention the local and global environmental impacts associated with large biomass plants, the difficulty in Europe to realise large scale biomass power plants and the actual European biomass market mainly composed of a plethora of very small scale few kW\textsubscript{th} biomass boiler and few large plants around 10 MW\textsubscript{th}), this report focus on smaller size (0.05-10 t/day). The CAPEX and OPEX for the other sizes of the chosen gasifier are estimated by a scaling method. The results of this section will be used in the deliverable 7.5 (business exploitation).
2  Technical analysis

2.1  System data: gasifier types, biomass, lifetime and availability

The general assumption according to the concept of UNIFHY is the UNIQUE gasifier (catalytic filter candles in the freeboard) coupled with water gas shift (WGS) and pressure swing adsorption (PSA), considered into two different sizes 100kWth and 1MWth, fitted with bubbling fluidized bed reactors (BFB). Nevertheless, the 100 kWth BFB is an indirectly heated reactor fed by steam in the gasification chamber and by air in the combustion chamber. The 1 MWth BFB has only a chamber which is fed by oxygen and steam (in reality two chambers but they are not separated at the top).

The feedstock for all cases is considered almond shell with a 18 MJ/kg dry LHV (D2.1). Anyway all lignocellulosic biomass has LHV of 18 MJ/kg dry thus they can be considered equivalent, even if bulk density, proximate and ultimate analysis, sulphur, chlorine and minor elements can be different (D2.1). Indeed, the difference in bulk density, proximate and ultimate analysis and trace minor elements not bring relevant change in the global technical and economic analysis.

In general, the first element to consider in assessing viable biomass uses is the energy and economic feedstock production costs. The first one is evaluated by Energy Return On Energy Investment, EROEI, in GJ/GJ while the second one is evaluated by the production cost divided by the useful Heating Value (HV), in €/GJ. The feedstock price is the largest component of the operating costs in a biomass plant and varies from negative price of some waste biomass (e.g. -10 €/t of particular biomass disposal waste) to high price of some dedicated crops (e.g. 500 €/t of particular crops). Fixing an energy yield value of 100 GJ/ha (e.g. yield of 10 t/ha and a calorific value of 10 GJ/t), for a value of 10 GJ/ha for cultivation and harvesting, the energy production cost is 0.1, while a mean economic cost is about €4/GJ. These average optimistic values include, among other items, transport energy and its economic costs of 0.5 MJ/km and 0.02 €/km per ton[21]. Lower yield and lower HV biomass does not have proportionally lower costs; therefore, the energy and economic returns could become negative. For this reason, it is preferable to use low cost residual biomass.

The main residual biomasses are waste, shells, pruning, straw and agro-industrial residues. Among waste we can mention Organic Fraction of Municipal Solid Waste and the manure. Among shells the main used are the shells of pine, hazel, walnuts and almonds. Among pruning the main are the pruning of beech, oak, spruce, poplar, willow, eucalyptus, grape and olives.
main straw used are the straw of wheat, corn, rye, barley, rice. Among agro-industrial residues, we can mention food, textile and wood industrial residues like cane trash, exhausted olives, pomace, etc. In particular, biomass must have availability on a significant scale (t/year). In every energy conversion process, because of energy needs in terms of efficiency and power density, fuels with a high LHV are favourites. This meaning that biomass with lower humidity is preferable. Seasoning can reduce the moisture content or the excess of heat produced by the power plant could be exploited to dry biomass in order to use also biomass with 50% of moisture. Even if, owing to the S/B used in this process (see gasifier section), it can be possible to accept also larger humidity content. The density affects significantly any freight and storage. Furthermore, in fluidized bed gasifier to have a good mixing between fuel and bed material, the biomass density should be comparable with that of the bed. Another important feature that must be considered is the size and shape of the biomass feeding the gasifier. Biomass must be processed to a uniform size or shape to feed into the gasifier at a consistent rate and to ensure homogeneous and efficient gasification. This can lead to significant costs for the shredding: chip size (1-2 cm) is at the moment the right compromise. The chemical composition (C, H, O, N, S, Cl) is another important aspect that must be considered. For lignocellulosic biomass the chemical composition (expressed on a dry and ash free basis) is generally more constant than that of other solid fuels (MSW, coal). Furthermore, more than 80% of the biomass is volatile the remaining 20% is charcoal. Coal is typically only 20% volatile, while the remaining 80% is unreactive coke, which is more difficult to gasify than charcoal. Generally, lignocellulosic biomass has very low Sulphur and Chlorine content compared to coal and MSW. Finally, Ash and TAR contents are one of the main obstacles to economical and viable applications of biomass gasification technologies. Fuels with a high ash content require greater attention because ash brings sintering, agglomeration, deposition, erosion and corrosion problems. Furthermore, they are elutriated by the producer gas, thus more is the ash content and much more problematic will be the gas cleaning procedures. TAR condenses as the temperature decreases, causing clogging and damage to the downstream equipment [22].

To sum up, the most suitable biomass for gasification must have availability on significant scale (t/year) and a good physical (low water content and high bulk density) and chemical characteristics (high Caloric Value, high volatile substances, low ash, high Carbon to Nitrogen ratio, low Chlorine and Sulphur content). The focus, as mentioned in D1.2, is on lignocellulosic biomass waste like “shells” (of pine, hazel, walnuts and almonds); “pruning” (of
wood/forestry/agricultural “threes”, thus: beech, oak, spruce, poplar, willow, eucalyptus, grape, olives); “straws” (of wheat, corn, rye, barley, rice); agro-industrial residues (e.g. dry exhausted olive), energy cultivation (e.g. Miscanthus). Among these, shells have the more suitable characteristics (low humidity content not great variable, high density, low ash content, high calorific value). Prunings have a greater variation of the characteristic. Straws/agro-industrial residues not only have a larger characteristic variation, but also a higher ash content that in many case bring to a melting temperature lower than the gasification temperature and thus clog the reactor. Regarding the CHO the lignocellulosic biomass has almost the same wt percentage (respectively 41-51, 5-6, 36-44). N, Cl, S accounts for very low percentages that vary depending on the biomass typologies and cultivation characteristics (soil, fertilizers, etc.). In summary, lignocellulosic waste, as the almond shell used in the UNIfHY plants, with heating value of 18 MJ/kg and a price of 75 €/t (an average lignocellulosic price between the low cost agro-industrial residues and pruning and the higher cost of shells and energy cultivation at the sizes considered, i.e. 0.1-10 MW$_{th}$ input) are considered in this report. Indeed, the technical and economic potentials of biomass are higher than the current world energy consumption, thus, the challenge is in its viable and sustainable use and not in its availability (as long as there is life there will be availability of organic material, used “directly” by living organisms as their own source of energy and materials (food) or used “indirectly” like a source of external energy (biomass) and materials: (clothing, furniture, buildings, chemicals, etc.) [23].

A 20 years lifetime is assumed for the power plant with 7000 h/year of operation, as a global lifetime and operating hours, consistent with these data for gasifier power plants. The efficiency and lifetime of each component is mentioned below. Consistency of these data with the simulation model and the experimental results are discussed.

### 2.2 Summary of the main components technical data

Here below are quoted from the deliverables the main component data.

#### 1- Gasifiers

- The start up time for 100 kW is 5 hours and 24 hours for 1MW$_{th}$ (D4.2 100 kW$_{th}$ gasifier tests, D5.1 1000 kW$_{th}$ gasifier tests).
- 1 MW$_{th}$ operating parameters are biomass feeding rate of 200 kg/h, gasification temperature 800 °C, Steam/Biomass 0.5, Equivalent ratio 0.21 (D5.1, D5.2 1000 kW$_{th}$ UNIfHY tests, D6.4 Global simulation).
Techno-economic analysis of UNIFHY hydrogen production system

- 100 kWth operating parameters are biomass feeding rate of 20 kg/h, gasification temperature 800 °C, Steam/Biomass 0.5 (D4.2, D4.3 100 kWth unifhy tests, D6.4).

2- Catalytic filter candles
- The number of candles integrated in the freeboard of the reactor is 6 and 60 respectively for 100 kWth and 1MWth gasifiers.
- The downstream gas composition at the outlet of gasifier coupled with catalytic candles is: H₂ 45%vol., 26% CO₂, 18% CO, Methane 10% vol, Tar 1g/Nm³ (D2.2, D6.3, D6.4)

3- WGS
According to D2.5 (Synthesis of Cu/Foam catalyst and characterization), D3.2 (WGS design and the construction), D6.1 (Kinetic model of reactions in HT-LT water gas shift reactors), D6.4 (Global simulation) we can define the following assumptions:
- The lifetime of LT (low temperature) WGS catalysts and ZnO guard bed are respectively considered as 8 years and 80 days.
- Operating temperature 300°C
- Residence time 1.2 s
- The CO conversion with Cu foams (5.0 wt%, Cu/45 ppi foam) is considered at 45%.

4- PSA
According to the D2.3 (PSA coupling verification), the D3.1 (PSA design and the construction), the D3.3 (PPU integration), the 5.2, the D6.4 (Global simulation) we use the following data:
- Hydrogen recovery efficiency 70% (50% of H₂ concentration in the inlet gas and a purity 4.0at the outlet).
- Fed gas temperature and residence time in PSA unit are 30°C and 9.8 s, respectively.
In general, the technical assessment is based on the 6.4 data. Here follow a specific components description in order to justify the use of the 6.4 data.

2.2.1 Gasifier 1000kWth
According to Deliverable 5.1, experimental tests campaign with the gasifier in its original configuration (without catalytic filter candles) were performed in the range temperature of 820-900 °C with three different gasification medium: 35%-wt O₂ (850-900 °C), 50%-wt O₂ (850-870...
Deliverable D5.3

Techno-economic analysis of UNIFHY hydrogen production system

°C) and steam/O₂ (820-830 °C) with a steam to biomass ratio (S/B) of 0.4 and an equivalent ratio (ER) of 0.21. By changing from enriched air to steam/oxygen mixture, the data confirmed the expected beneficial effect on the heating value of the product gas. The improvement in the gas quality was indeed not only an effect of the decreased amount of N₂, but also a result of the addition of steam which induced an H₂ enrichment by promoting reaction of gas upgrading, such as water gas shift, char gasification and hydrocarbons reforming. A Cold Gas Efficiency (CGE) increasing from 53% up to 65% was observed when passing from tests carried out with 35 %-wt oxygen enriched air to tests carried out using steam/oxygen mix.

<table>
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<tr>
<th>Test</th>
<th>S/B</th>
<th>ER</th>
<th>Gasification</th>
<th>Temperature [°C]</th>
<th>Pressure [bar_	ext{a}]</th>
<th>Gas yield [kg wet/kg Biomass, dry]</th>
<th>H₂ [%v, Dry]</th>
<th>LHV [MJ/Nm³]</th>
<th>CGE [%]</th>
<th>Tar [g/Nm³ dry]</th>
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<td>850-900</td>
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<td>1.27</td>
<td>7-9</td>
<td>5.9-6.7</td>
<td>53</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
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<td>-</td>
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<td>850-870</td>
<td>1.0-1.1</td>
<td>1.28</td>
<td>11-13</td>
<td>6.3-8.4</td>
<td>-</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>Test 3</td>
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<td>0.21</td>
<td>820-830</td>
<td>1.0-1.1</td>
<td>1.67</td>
<td>30-33</td>
<td>10.9-11.7</td>
<td>65</td>
<td>75</td>
<td>12-18</td>
</tr>
<tr>
<td>Test 4</td>
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<td>0.22</td>
<td>860-880</td>
<td>1.0-1.1</td>
<td>-</td>
<td>25-26</td>
<td>13.1-13.5</td>
<td>75</td>
<td>-</td>
<td>11-14</td>
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<td>Test 5</td>
<td>0.4</td>
<td>0.21</td>
<td>800</td>
<td>1.0-1.1</td>
<td>-</td>
<td>39-42</td>
<td>-</td>
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Table 1. Summary of the main parameters of the tests on UNIFHY 1000 (D5.1 & D5.2)

Since these considerations, it was possible to assume steam/oxygen mixture the preferable gasification medium for UNIFHY 1000 gasifier and perform a second test campaign focused on this medium and on two different gasification temperature ranges, as 820-830 °C and 860-880 °C. It was note that a higher gasification temperature provides a higher CGE (75% versus 65%) and also the results on tar indicated that the content was slightly lower at temperature in the range 860 – 880 °C (14 versus 18 g/Nm³ dry). As far as the gas composition, the increasing of the biomass rate slows down the percentage in hydrogen (25%), but, enhancing the steam to biomass ratio up to 0.5 with an equivalent ratio of 0.21, the results, in a best scenario, could be better with a percentage in hydrogen of 33%. Since these considerations, several gasification experimental
tests with the gasifier in its advanced configuration (with catalytic filter candles and PPS integration) were performed, at a gasification temperature of 800 °C, S/B 0.5-0.6, ER 0.24-0.25 (D5.2). Subsequently, other gasification tests at S/B 0.7 and ER 0.21 were carried out, obtaining a high hydrogen concentration very close to 40%.

For these reasons, the global simulation of Deliverable 6.4 and the economic analysis performed in this deliverable were conducted setting the gasifier temperature at 800 °C, S/B at 0.5 and biomass feeding rate of 200 kg/h. The economic analysis was conducted also by carrying out a sensitivity study on gasification temperature (800-850 °C) and higher steam to biomass ratio (0.5-1.5).

2.2.2 Gasifier 100kWth

According to Deliverables 4.2 and 4.3, test with only steam and steam/air were carried out with the 100 kWth prototype, in order to point out the difference between the composition of the syngas produced by a dual fluidized bed (steam tests) and a normal fluidized bed (steam/air tests). The tests were carried out at a gasification temperature of 720-770 °C and S/B equal to 0.5-0.6, for the former case, and S/B equal to 0.3 and ER equal to 0.4, for the latter case. The results showed a much higher volumetric percentage of H\textsubscript{2} (27% with steam versus 10% with steam/air) and highlighted the effective higher performances of a dual fluidized bed fed with only steam. Moreover, the various gasification tests showed that with an increase of the gasification temperature there is an increase of the produced gas (due to a more efficient gasification). Unfortunately, the syngas composition from the steam tests still contains a percentage of nitrogen, probably coming from leakages between the two chambers or from non-optimized design of the siphons, and a tar content quite high, probably due to the low temperature (770 °C) and the non-optimized design of the freeboard and the gasifier. A new configuration of 100 kWth gasifier was then designed and realized, in order to achieve a higher reaction temperature and an improved structure.

The global simulation of Deliverable 6.4 and the economic analysis conducted in this deliverable were performed at a biomass feeding rate of 20 kg/h and sensitivity studies were carried out on gasification temperature (800-850 °C) and steam to biomass ratio (0.5-1.0-1.5).
2.2.3 Filter candles

As regards filter candles, the results of Deliverable 2.2, in which three different types of filters hosted in the freeboard of the reactor vessel were tested, such as non-catalytic candle, catalytic candle and catalytic candle with catalytic foam, showed that the catalytic candles are effectively the best choice for UNIfHY purposes. During tests the temperature was about 800 °C and the filtration velocity was about 90-100 m/h. In particular, the hydrogen content is highly increased by using catalytic candles, being in a range of 49%-57% (without and with foam respectively) in spite of 38% with no candle (Figure 1 (a)). Together to the hydrogen increase, in the case of runs with filter candles with respect to the case without candle, it could be observed a decrease of all remaining gas components (mostly methane). Even as regards the total tar gas content reduction, the catalytic candles provide very low concentration respect to no-catalytic candles, as shown in Figure 1 (b).

![Figure 1](image.png)

**Figure 1.** Gas composition (vol.dryN₂free) and total tar content (101;102,116,117: no catalytic; 103-107: catalytic; 108-113 catalytic foam)

The CFD simulation analysis carried out in the Deliverable 6.3 was validated using CH₄ and tar conversion rate obtained by previous experimental results. It revealed that the maximum conversion rates are reached at a temperature of 850 °C and a filtration velocity of 70 m/h. However, the manufacturer PALL Schumacher suggests the usage of a filtration velocity of a 90 m/h. Since this consideration, the number of catalytic filter candles suitable for UNIfHY system should be thus 60 and 6 respectively for 1 MWₜₐₜ and 100 kWₜₐₜ gasifiers. As regards the operating filter candles temperature, CFD analysis revealed that even for filtration velocity of 90 m/h the conversion rate is maximized at 850 °C. Nevertheless, global experimental test on UNIfHY 1000
Deliverable D5.3

Techno-economic analysis of UNIFHY hydrogen production system

advanced system provided a very low conversion rate of the catalytic filter candles, according to Deliverable 5.2. This was probably because, even if the gasification temperature of the tests was around 800 °C, the temperature of the freeboard was quite lower and very far from the CFD simulated temperature of 800-850 °C. In addition, catalytic candles were subjected to a too low H₂ flow, which didn’t provide the complete activation of the filters catalyst.

For these reasons, in the Deliverable 6.4, dealing with global simulations of the systems, the catalytic filter candles were simulated by using values of conversion rate relative to filtration velocity of 90 m/h and operating temperature of 800-850 °C (taken from D6.3), according to the gasification temperatures.

<table>
<thead>
<tr>
<th>SB=0.5</th>
<th>SB=1.0</th>
<th>SB=1.5</th>
</tr>
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<tbody>
<tr>
<td>XCH₄ [%]</td>
<td>-5.44</td>
<td>-18.5</td>
</tr>
<tr>
<td>XCO [%]</td>
<td>12.26</td>
<td>15.7</td>
</tr>
<tr>
<td>XC₆H₆ [%]</td>
<td>-58.36</td>
<td>-49.3</td>
</tr>
<tr>
<td>XC₇H₈ [%]</td>
<td>-76.173</td>
<td>-72.3</td>
</tr>
<tr>
<td>XC₁₀H₈ [%]</td>
<td>-94.35</td>
<td>-88.7</td>
</tr>
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</table>

Table 2. Conversion rate of catalytic candles

Therefore, the D6.4 data are used in this report. In particular, we use 10.5 kg/h of hydrogen produced for the indirectly heated gasifier and 9.3 kg/h of hydrogen produced for the Steam/oxygen fluidized bed gasifier.

2.2.4 PPS

According to Deliverables 2.5, 3.2 and 3.3, an innovative catalytic system has been specifically developed for Water Gas Shift reaction. To prevent pressure drop along the process because of the gasification is operated at atmospheric pressure and to increase the efficiency of the gas-solid contact, all the catalysts were supported on ceramic foams. Two type of ceramic foams were tested for the two water gas shift reactor, the high temperature reactor (450 °C) and the low temperature reactor (300 °C). After the analysis of the experimental tests carried out on the high temperature water gas shift reactor, it was decided to eliminate this component in the UNIFHY system, since it will not provide satisfactory results, it will considerably increase the pressure drop and in the contrary it will add a useless complication of the entire plant. Low temperature water gas shift reactor was filled with ceria (CeO₂) foam impregnated with copper (Cu). Two
different foam of different porosity, 45 ppi and 30 ppi respectively, were subjected to reactivity tests with a feed gas composition similar to that of the gasifier outlet in the range of 150-300°C and a residence time set to 1.28 s.

The maximum CO conversion (42 %) was obtained with a 5.0 wt.% Cu. Analyzing the two kinds of foam, the 30 ppi ones showed very lower pressure drop respect to the 45 ppi, thus the 30 ppi foams where chosen to prevent exceeding the differential pressure limits.

Due to the presence of hydrogen sulfide in the gasifier outlet gas it was decided that a ZnO guard bed would be included to increase lifetime of the LT-WGS catalyst. The temperature of the WGS reactor should not be higher than 300 °C and the H₂S concentration limit in the syngas flow should not be higher than 0.1 ppm, in order to avoid catalyst damage. The lifetime of LT-WGS catalysts and ZnO guard bed are respectively considered as 8 years and 80 days.

Nevertheless, global experimental test on UNIfHY 1000 advanced system provided a very low conversion rate of the catalytic ceramic foams, according to Deliverable 5.2. This was probably because the temperatures were too low for operation of the WGS and the catalyst was not sufficiently activated (see D5.2).

For these reasons, in the Deliverable 6.4, dealing with global simulations of the systems, the LT-WGS reactor was simulated by using values of conversion rate coming from experimental test on ceramic foams, an operating temperature of 300 °C and a residence time of 1.28 s. These parameters were used also for the subsequent economic analysis.

According to Deliverables 2.3, 3.1 and 3.3, the pressure swing adsorber system (PSA) was designed, constructed and experimentally tested. The main parameters of PSA were a H₂ yield of about 65-70% with a H₂ purity of 4%, an adsorption and feed-gas pressures of 6 bar and a feed-gas temperature of 25-30 °C. The feed-gas flow suitable for the designed PSA is 60 Nm³/h. The PSA performances were also confirmed by the experimental tests carried out in the D2.3. Hence, these parameters were used in the global simulation of the Deliverable 6.4, where a separation
efficiency of 70% was taken into account, besides of a feed-gas pressure of 6 bar and a feed-gas temperature of 30 °C. These parameters were used also for the subsequent economic analysis.
3 Economic analysis

3.1 General economic consideration on biomass plants

Differently from other renewable energy sources, biomass, being classifiable as a fuel, and not an energy resource directly convertible like solar, wind, hydro, and geothermal energy, is subjected to all traditional steps needed to make it available on the energy market (Production, Transport, Conversion, Distribution, End use). Nevertheless, residual biomass, if used in situ, is subjected only at the last three steps and an accurate analysis and design can change the potential negative impacts into positive ones. Indeed, the use of biomass can also remove soil contaminants and reduce pollution, if the biomass power plants meet strict environmental standards (e.g. correct energy use of a biomass waste normally burned in the field). Finally, social implications have to be carefully taken into account. Bio-energy systems require complex organization, many actors, substantial land use and have, generally, negative social acceptance. However, biomass is always available and biomass plants can have positive economic, social and environmental impacts, particularly related to the equitable distribution of biomass and the close connection that it establishes between a community and its territory.

In this chapter, as quoted in the introduction. The analysis of Capital expenditures (CapEx) and Operating expenses (OpEx) for each component of the UNIfHY: 100 kWth plant (based on indirectly heated steam/air fluidized bed gasification), UNIfHY 1000 kWth gasifier plant (based on steam/oxygen fluidized bed gasification), Filter candles, PPS are showed below.

3.2 CapEX: hardware costs

Costs are split in Capital expenditures (CapEx) and Operating expenses (OpEx). CapEx is divided into two sections: Hardware costs and Engineering costs. The hardware cost of the different UNIfHY components are showed in the next paragraphs.

3.2.1 Gasifier 1000 kWth

Hardware costs consist of construction and installation costs of major equipment. The below table indicates this cost for 1MWth gasifier (this cost coming from real cost occurred building the plant in Trisaia, Basilicata, Italy)

<table>
<thead>
<tr>
<th>Major Equipment (€/kWth)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Biomass feeder unit</td>
</tr>
<tr>
<td>Biomass Storage &amp; Feed (Extraction screw, Auger transfer, Bucket elevator, hopper)</td>
</tr>
</tbody>
</table>

March 2016
Techno-economic analysis of UNIFHY hydrogen production system

<p>| | | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Gasifier</td>
<td>Gasifier</td>
<td>188</td>
</tr>
<tr>
<td>Gasification agents</td>
<td>Steam Generator&amp; Feeding system</td>
<td>14</td>
</tr>
<tr>
<td>Cooling Water &amp; Other</td>
<td>Ash Handling Equipment</td>
<td>25</td>
</tr>
<tr>
<td>Utilities</td>
<td>Piping and valves</td>
<td>55</td>
</tr>
<tr>
<td></td>
<td>Auxiliaries (Burner, Blower,</td>
<td>54</td>
</tr>
<tr>
<td></td>
<td>Flare)</td>
<td></td>
</tr>
<tr>
<td></td>
<td>Control</td>
<td>135</td>
</tr>
<tr>
<td>Building &amp; Structures</td>
<td>Concrete Foundation</td>
<td>18</td>
</tr>
<tr>
<td></td>
<td>Plant Structures</td>
<td>59</td>
</tr>
<tr>
<td>Total</td>
<td></td>
<td>(701 €/kWth)</td>
</tr>
</tbody>
</table>

Table 3 Direct CAPEX of UNIFHY 1000

According to Figure 2, the cost of gasifier is the largest one which is followed by cooling water & other utilities, within the same cost range. Moreover, if we consider having an O₂ production plant, 219.86 k€, (this cost is approximated based on CapEx of Oxygen production plant with 300 t/day yield rate [24] and scaling factor (0.76) mentioned in Table 9 for air compressor due to the fact that air compressor has the biggest share of total Oxygen plant cost, scaling factor related that is applied for scaling whole O₂ plant) will be added to the cost. Therefore, the total cost for that configuration will be 921 €/kWth while the O₂ production plant will be the main cost.

Figure 2 Distribution of Hardware costs (UNIFHY 1000)

3.2.2 Gasifier 100 kWth

Owing to the fact that UNIFHY 100 is a pilot and not an industrial plant (e.g. there is not biomass storage & pre-treatment system, all the valves are manual, there is not heat exchanger for air and
steaming heating but electrical heater, etc.) instead of use the real cost occurred building the plant at CIRPS, Rome, Italy, we use scaling factor to calculate this cost.

**Scaling factor**

The scaling exponents used in systems analysis work are logarithmically derived from previously obtained vendor supplied cost quotes using Equation 1 [25].

\[
Exp = \frac{\ln \left(\frac{RC_1}{RC_2}\right)}{\ln \left(\frac{RP_1}{RP_2}\right)}
\]

Where:
- Exp – Exponent
- RC – Reference Cost
- RP – Reference Parameter

<table>
<thead>
<tr>
<th>Item</th>
<th>unit</th>
<th>value</th>
<th>Exp</th>
</tr>
</thead>
<tbody>
<tr>
<td>Biomass storage and feed</td>
<td>Biomass feed rate lb/hr</td>
<td>412000-616000</td>
<td>0.66</td>
</tr>
<tr>
<td>Circulating water pumps</td>
<td>water flow rate (gpm)</td>
<td>115000-550000</td>
<td>0.73</td>
</tr>
<tr>
<td>Circ. Water Piping</td>
<td>Circulating Water flow rate,gpm</td>
<td>115000-550000</td>
<td>0.63</td>
</tr>
<tr>
<td>Ash transport&amp;Feed Equipment</td>
<td>Total Ash Flow, lb/h</td>
<td>10-100</td>
<td>0.56</td>
</tr>
<tr>
<td>Gasifier</td>
<td>Total feed flow rate, lb/h</td>
<td>303000</td>
<td>0.69</td>
</tr>
<tr>
<td>Control Board, panels&amp;Rocks</td>
<td>Auxiliary Load, kW</td>
<td>283000-272000</td>
<td>0.13</td>
</tr>
<tr>
<td>Instrument Wiring &amp;Tubing</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Buildings &amp; Structures</td>
<td>N/A</td>
<td>735000-1630000</td>
<td>0.1</td>
</tr>
</tbody>
</table>

**Table 4 Scaling factors employed for estimation** [25]

Equation 2 is used to scale costs [25].

\[
SC = RC * \left(\frac{SP}{RP}\right)^{Exp}
\]

Where:
- Exp – Exponent
- RC – Reference cost
- RP – Reference Parameter
- SC – Scaled cost
- SP – Scaling parameter
Scaling factors in Table 4 were applied for estimation of different sizes. Although the range of value presented to use each exponent is much greater than the amount calculated in our power plant, they were employed as the only reference for these factors. Nevertheless, the cost obtained are in line with the cost of the same components already occurred (e.g. gasifier, piping, auxiliaries, etc. costs are very similar to the real cost occurred).

The below table indicates hardware cost estimated based on scaling factor for 100 kWth gasifier.

<table>
<thead>
<tr>
<th>Major Equipment</th>
<th>UNIFY 100 [k€]</th>
<th>€/kWth</th>
<th>%</th>
</tr>
</thead>
<tbody>
<tr>
<td>Biomass feeder unit Biomass Storage &amp; Pre-treatment</td>
<td>32</td>
<td>320</td>
<td>18</td>
</tr>
<tr>
<td>Gasifier</td>
<td>38</td>
<td>380</td>
<td>21</td>
</tr>
<tr>
<td>Gasification agents Steam Generator &amp; Feeding system</td>
<td>3</td>
<td>30</td>
<td>2</td>
</tr>
<tr>
<td>Cooling Water &amp; Other Utilities Ash Handling Equipment</td>
<td>7</td>
<td>70</td>
<td>4</td>
</tr>
<tr>
<td></td>
<td>Piping and valves</td>
<td>13</td>
<td>130</td>
</tr>
<tr>
<td></td>
<td>Auxiliaries (Burner, Blower, Flare)</td>
<td>10</td>
<td>100</td>
</tr>
<tr>
<td></td>
<td>Flue gas cleaning</td>
<td>5</td>
<td>50</td>
</tr>
<tr>
<td></td>
<td>Control</td>
<td>10</td>
<td>100</td>
</tr>
<tr>
<td>Building &amp; Structures Concrete Foundation</td>
<td>14</td>
<td>140</td>
<td>8</td>
</tr>
<tr>
<td></td>
<td>Plant Structures</td>
<td>47</td>
<td>470</td>
</tr>
<tr>
<td>Total</td>
<td>179</td>
<td>1790</td>
<td></td>
</tr>
</tbody>
</table>

Table 5 Direct CAPEX of UNIFHY 100 (CIRPS)

As inferred from table above, the cost of structures, control unit, biomass storage in smaller size than 1 MWth (reference parameter) rises. This result was predictable based on literature review and on the exponent values of the scaling factors.

The CapEx of indirectly heated gasifier can be different from the one of steam/Oxygen gasifier due to existence of flue gas cleaning filter, more steel used to construct double bed gasifier. After calculation of steel used in construction of two gasifiers, the difference of material used is found negligible. Since extra steel used in double bed indirectly heated gasifier is 7% of steel used in gasifier with Oxygen which leads to only 1% increase in cost, two gasifier construction costs are considered the same.

3.2.3 Gasifier plants cost comparison

In order to recognize if the cost of the 1 MW and 100 kW prototypes are more or less in line with the cost of commercial gasifiers, we quote the costs of two commercial power plants in table below. We choose power plants that we have detailed cost data and that are of smaller size. The
sizing are bigger than our prototypes but this is unavoidable owing that the commercial scale of the biomass gasifier start from 4-5 MW.

<table>
<thead>
<tr>
<th>Major Equipment</th>
<th>4 MW&lt;sub&gt;th&lt;/sub&gt;</th>
<th>8 MW&lt;sub&gt;th&lt;/sub&gt;</th>
<th>%</th>
<th>%</th>
</tr>
</thead>
<tbody>
<tr>
<td>Biomass feeder unit</td>
<td>Biomass Storage &amp; Feed</td>
<td>307</td>
<td>76</td>
<td>21</td>
</tr>
<tr>
<td>Gasifier</td>
<td>Gasifier</td>
<td>418</td>
<td>105</td>
<td>29</td>
</tr>
<tr>
<td>Gasification agents</td>
<td>Steam Generation &amp; Feed</td>
<td>30</td>
<td>8</td>
<td>2</td>
</tr>
<tr>
<td></td>
<td>Water heater</td>
<td>11</td>
<td>3</td>
<td>1</td>
</tr>
<tr>
<td></td>
<td>Oxygen producer and preheater</td>
<td>128</td>
<td>32</td>
<td>9</td>
</tr>
<tr>
<td></td>
<td>Steam and Oxygen mixer</td>
<td>21</td>
<td>5</td>
<td>1</td>
</tr>
<tr>
<td>Cooling Water &amp; Other Utilities</td>
<td>Ash Handling Equipment</td>
<td>N/A</td>
<td>N/A</td>
<td>N/A</td>
</tr>
<tr>
<td></td>
<td>Piping and valves</td>
<td>128</td>
<td>32</td>
<td>9</td>
</tr>
<tr>
<td></td>
<td>Auxiliaries (Burner, Blower, Flare,...)</td>
<td>189</td>
<td>47</td>
<td>13</td>
</tr>
<tr>
<td></td>
<td>Control</td>
<td>N/A</td>
<td>N/A</td>
<td>N/A</td>
</tr>
<tr>
<td>Building &amp; Structures</td>
<td>Concrete Foundation</td>
<td>N/A</td>
<td>N/A</td>
<td>N/A</td>
</tr>
<tr>
<td></td>
<td>Plant Structures</td>
<td>236</td>
<td>59</td>
<td>16</td>
</tr>
<tr>
<td>Total</td>
<td>1468</td>
<td>367</td>
<td>2100</td>
<td>263</td>
</tr>
</tbody>
</table>

Table 6 Direct CAPEX of 4 and 8 MW<sub>th</sub> reference gasifier plants

The data shows that size increasing leads to decrease in cost of major equipment. Comparison of cost estimated for 100 kW<sub>th</sub> with data from literature (8 MW<sub>th</sub>) shows the same trend. Biomass feeder unit and Gasifier costs dip from 320 and 380 to 41 and 108 €/kW<sub>th</sub> of biomass input, respectively. Also, the cost of Cooling Water & Other Utilities falls from 450 to 91 €/kW<sub>th</sub>. Meanwhile the largest decrease in cost belongs to Building & Structures which is reduced from 610 to 23 €/kW<sub>th</sub>.

3.2.4 Filter candles

The number of catalytic filter candles inserted into freeboard of gasifier varies based on different sizes of gasifier. They are considered 6, 60 and 600 in 0.1, 1 and 10 MW<sub>th</sub> power plants, respectively. Regarding that each candle costs 1400 € (Pall Schumascher data), the cost of candles for different sizes are 8.4, 84 and 840 k€ for 0.1, 1 and 10 MW<sub>th</sub> power plants, respectively. The lifetime of these candles are considered the same with whole power plant (20 years). Obviously we have to take into account that this cost is a prototype cost and not an industrial cost, thus there is a huge reduction of cost owing large production of these candles. Moreover, the lifetime (considering the catalyst life) is not 20 years but, e.g. common similar catalyst life time is 7 years. But, owing to the uncertainty of these two opposite factors we do not
take into account (i.e. a common catalyst life time is 7 years but the price can be reduced at 1/3 thus having the same global cost for 20 years).

### 3.2.5 PPS

This cost coming from real cost occurred building the plant in HyGear, Netherlands are showed in the table below.

<table>
<thead>
<tr>
<th>Portable Purification System</th>
<th>Unit (k€)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pressure Swing Adsorption Unit</td>
<td>110</td>
</tr>
<tr>
<td>Low Pressure Compressor &amp; Blower</td>
<td>100</td>
</tr>
<tr>
<td>Rotary and slide Valves, Sensors &amp; Other Controls</td>
<td>80</td>
</tr>
<tr>
<td>Water Gas Shift and ZnO Reactor</td>
<td>50</td>
</tr>
<tr>
<td>Pipeline, Heat Exchanger &amp; Housing</td>
<td>55</td>
</tr>
<tr>
<td>Total</td>
<td>395</td>
</tr>
</tbody>
</table>

**Table 7 Cost of PPS integrated with 200kWth pilot gasifier**

A study, showed in the table below, has been made by HyGear regarding the different cost of the PPS for different annual production unit.

<table>
<thead>
<tr>
<th>The size of power plant</th>
<th>The cost of units (M€)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 MWth</td>
<td>1.975</td>
</tr>
<tr>
<td>5 MWth</td>
<td>3.1</td>
</tr>
</tbody>
</table>

**Table 8 PPS cost based on the number of units**

The production cost of one unit in 1MWth and 5MWth could decrease around 50% and 70%, respectively if 10 units were produced. In addition, the production cost of one PPS unit for 5MWth plant is two times on average higher than the production cost of one PPS unit for 1MWth.

Similar to the 100 kWth plant cost, we use the scaling factor method to calculate the cost of the PPS at the 0.1, 1, 10 MWth input plant size. The table below shows the exponent used.

<table>
<thead>
<tr>
<th>Category</th>
<th>Parameter</th>
<th>Exponent</th>
<th>Range</th>
</tr>
</thead>
<tbody>
<tr>
<td>Shift reactors [25]</td>
<td>WGS catalysts volume ft3</td>
<td>0.12</td>
<td>2,000 – 25,500</td>
</tr>
<tr>
<td>Pressure Swing Adsorption Unit[19]</td>
<td>N/A</td>
<td>0.6</td>
<td>N/A</td>
</tr>
<tr>
<td>Blowback gas systems [25]</td>
<td>Candle filter flow rate acfm</td>
<td>0.3</td>
<td>2,000 – 96,000</td>
</tr>
<tr>
<td>Air compressor [25]</td>
<td>Fuel gas flow, acfm average</td>
<td>0.76</td>
<td>2,000 – 4,000</td>
</tr>
</tbody>
</table>
Table 9 Scaling factors employed to estimate

<table>
<thead>
<tr>
<th>PPS unit sections</th>
<th>Hardware costs</th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>100kW_{th} (k€)</td>
<td>%</td>
<td>€/kW_{th}</td>
<td>1MW_{th} (€/kW_{th})</td>
<td>%</td>
<td>10MW_{th}</td>
</tr>
<tr>
<td>Water Gas Shift and ZnO Reactor</td>
<td>50</td>
<td>18</td>
<td>50</td>
<td>56</td>
<td>5</td>
<td>131</td>
</tr>
<tr>
<td>Low Pressure Compressor &amp; Blower</td>
<td>59</td>
<td>22</td>
<td>590</td>
<td>340</td>
<td>33</td>
<td>1125</td>
</tr>
<tr>
<td>Pressure Swing Adsorber Unit</td>
<td>73</td>
<td>27</td>
<td>730</td>
<td>289</td>
<td>28</td>
<td>876</td>
</tr>
<tr>
<td>Rotary and slide Valves, Sensors &amp; Other Controls</td>
<td>53</td>
<td>19</td>
<td>530</td>
<td>210</td>
<td>20</td>
<td>637</td>
</tr>
<tr>
<td>Pipeline, Heat Exchanger &amp; Housing</td>
<td>36</td>
<td>13</td>
<td>360</td>
<td>144</td>
<td>14</td>
<td>438</td>
</tr>
<tr>
<td>Total</td>
<td>271</td>
<td>2260</td>
<td>1039</td>
<td>3207</td>
<td>322</td>
<td></td>
</tr>
</tbody>
</table>

Table 10 PPS cost estimated based on scaling factors

The comparison between 1000 MW_{th}, 100kWth gasifier and PPS shows that the PPS cost is 32% and 20% higher than gasifier cost. The Results of cost estimation in Table 10 reveals that substantial cost of PPS in 1MW_{th} and 10 MW_{th} power plants belongs to the compressor& blower and PSA (around 35% and 30%).

3.3 Engineering costs & OpEx

Engineering costs

- Engineering and design
- Purchasing and construction

This CapEx can be depreciating within N years, N depending on three main parameters:
- The lifetime duration of the hardware, considering the maintenance quoted in the OpEx costs
- The long term agreement for feedstock procurement
- The long term agreement for green-hydrogen off take

As quoted in the Technical analysis chapter (System data paragraph) the life time is set at 20 years. Indeed, 20 years of depreciation is standard for gasifier plants. That means that in the targeted business models, only locations where feedstock procurement and H₂ off take can be secured for 20 years shall be considered.
The cost of capital is set at 7%. The formula for calculation of annual capital costs is:

\[ \text{k€/year} = \frac{\text{CapEx in k€} \times 7\%}{1 - (1 + 7\%)^{-20}} \]

\( \text{OpEx} \)

Operating expenses for the plant covers the cost for:

- Personnel: operational staff
- Maintenance costs: spare part and sub-contracted maintenance
- Insurance
- Biomass and oxidizer
- Energy

The assumptions adapted to analyse the global costs:

- Engineering and design (13% total installed cost-DOE)
- Purchasing and construction (14% total installed cost-DOE)
- Personnel: 5 equivalent full time (EFT) per year (8 hr/day, 3000€/month)
- Maintenance: 2% total CAPEX (DOE)
- Insurance and taxes: 2% total CAPEX (DOE)
- Biomass: 75 €/ton (see chapter 2, Technical analysis, system data)[28]
- Annual operating hours: 7000 h (as quoted in the Technical analysis)
- VAT free
- Electrical energy consumption price 0.08 €/kWh

Untaxed electricity purchase cost is 0.11 €/kWh; 0.09 €/kWh and 0.08 €/kWh for 0.1, 1 and 10 MWth power plant [29]. Since these costs have a slight fluctuation, we can consider a similar value (0.08) for three different sizes.

The plant supervision in 1 to 10 MWth is granted to 5 personnel, i.e. a work shift made up of 3 people per 8 hr/day each ones, the remainders take into account holiday days, vacations, etc. in order to guarantee the 24/24 h 365/365 day plant supervision.

In 100 kWth plant, the supervision is not considered. This is due to two factors. First similar size power plants do not require full time supervision, because the owner is responsible of the check of the plant and of the easy biomass fueling. Second the control system can be easy automatic meanwhile adding a full time supervision would be not cost affordable at this size.
Taking into account the previously description the technical analysis is first done in order to compare the two technologies at same condition and size, thus following 6.4, at 800 °C gasification temperature and S/B 0.5 (the conditions more near to the experimental ones).

After, is calculated, for the best technology, the hydrogen yields at the best gasification temperature (i.e. 850 °C) and at lower (i.e. 0.5) and higher (i.e. 1.5) S/B in order to compare the less hydrogen efficient but with electrical and thermal energy production configuration (S/B=0.5) (fulfilling the energy plants needs and producing excess of electrical and thermal energy exploiting the residual off gas) with the higher hydrogen efficient but without electrical and thermal energy production configuration (fulfilling the thermal energy plants needs but not the electrical owing to the full use of the off gas for the thermal balance of the system, see D6.4).

4 Comparison of IH and steam/O2 UNIfHY plants at 1 MW$_{th}$ size

As quoted in the objectives of this report, in order to compare the two technologies a study at same size (1000 kW$_{th}$) and same operative parameters (near to the average experimental ones: 800 °C gasification temperature, S/B 0.5, see technical analysis paragraph) has been done. The cost of 1 MWth Indirectly heated are estimated via scaling factors, as done for 100 kWth previously. The cost of 1 MWth steam/oxygen are the cost quoted in the previously paragraph.

<table>
<thead>
<tr>
<th>Hardware costs [k€/year]</th>
<th>1MWth indirectly heated</th>
<th>1MWth with Oxygen production plant</th>
<th>1MWth without Oxygen plant</th>
</tr>
</thead>
<tbody>
<tr>
<td>Gasifier</td>
<td>71</td>
<td>87</td>
<td>66</td>
</tr>
<tr>
<td>PPU</td>
<td>98</td>
<td>98</td>
<td>98</td>
</tr>
<tr>
<td>Engineering Costs [k€/year]</td>
<td>20</td>
<td>22</td>
<td>20</td>
</tr>
<tr>
<td>Purchasing and construction</td>
<td>22</td>
<td>24</td>
<td>21</td>
</tr>
<tr>
<td>Total CAPEX</td>
<td>211</td>
<td>231</td>
<td>205</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>OPEX [k€/year]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Personnel</td>
</tr>
<tr>
<td>Maintenance</td>
</tr>
<tr>
<td>Insurance</td>
</tr>
<tr>
<td>biomass</td>
</tr>
<tr>
<td>Energy</td>
</tr>
<tr>
<td>O$_2$</td>
</tr>
<tr>
<td>O$_2$ tank</td>
</tr>
</tbody>
</table>
Techno-economic analysis of UNIFHY hydrogen production system

<table>
<thead>
<tr>
<th>Total OPEX</th>
<th>394</th>
<th>425</th>
<th>527</th>
</tr>
</thead>
<tbody>
<tr>
<td>Total cost</td>
<td>605</td>
<td>656</td>
<td>732</td>
</tr>
<tr>
<td>Hydrogen production [Ton/year]</td>
<td>73.5</td>
<td>65.1</td>
<td>65.1</td>
</tr>
<tr>
<td>Hydrogen production cost [€/kg]</td>
<td>8.2</td>
<td>10.0</td>
<td>11.2</td>
</tr>
</tbody>
</table>

Table 11 Comparison of three configurations

The total CAPEX is slightly different, the most expensive is 1 MW_th steam/oxygen with Oxygen production plant, mainly because it needs of an air separation unit. The indirectly heated has additional steel (for the combustor reactor) and flue gas line, but this increase the CAPEX in a very little manner (less than 1% compared to the Oxygen without oxygen production). Thus, indirectly heated has a little higher efficiency (higher H2 production yield) and does not need of particular increase of CAPEX and OPEX than the other solutions (the CAPEX are between the other two configurations, the OPEX are the lowest), these factors lead to a low final production cost.

The total electrical consumption is higher for the steam/oxygen gasifier and is mainly due to the higher electrical consumption due to air compressor for the production of enriched air/pure oxygen. The total consumptions of the two plants varied between 62 kW to 78 kW: the mains are due to the PSA intercooler compressor, the air compressor and the syngas blower.

According to the results obtained from the Table 11, H2 production via 1 MW_th indirectly heated can be the most affordable with the specific cost of 8.2 €/kgH2. Therefore, from economic and technical point of view it appears to be the best configuration even if the difference in efficiency and cost are not so relevant.

5 UNIfHY under different scenarios

As quoted in the Objectives, an economic and technical evaluation of the best technology (Indirectly Heated) at the best operating parameters (i.e. 850 °C gasification temperature, S/B 1.5) for different sizes (100 kW_th, 1 MW_th, 10 MW_th) has been done. This evaluation considers also the S/B 0.5 because there are two choices: to reach high H2 yields (S/B=1.5) or using it in CHP mode (S/B=0.5). Indeed, one of the most important parameter which influences the

---

1 They are considered always operative in nominal conditions.
Techno-economic analysis of UNIFHY hydrogen production system

technical and economical solutions is the S/B ratio. This leads to different plant solutions, the first one (S/B 1.5) needs a more complicated power plant owing to the more recovering of the internal process energy, the second one requires additional cogenerative internal combustion engine but it is simpler to realise and manage, for this reason the two configurations have been analysed.

As showed in the following table, a higher S/B ratio allows H$_2$ production to increase by 32% keeping a comparable OPEX cost, while the CAPEX cost depends particularly by the oversized steam generator, but this increment accounts for maximum 3% and so it is irrelevant respect to the efficiency increase.

Moreover, the increase in hydrogen production owing to the gasification temperature increase (from 800 to 850 °C) leads to about 8% increase, thus the hydrogen cost decrease from the ones of the previously comparison (at same S/B, i.e. 0.5) and decrease more at S/B 1.5 (from 8.2 to 7.6 and 5.8 €/kg considering the 1 MWth size).

<table>
<thead>
<tr>
<th>Hardware costs [k€/year]</th>
<th>S/B=0.5</th>
<th>S/B=1.5</th>
<th>S/B=0.5</th>
<th>S/B=1.5</th>
<th>S/B=0.5</th>
<th>S/B=1.5</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>100kWth</td>
<td>1MWth</td>
<td>10MWth</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gasifier</td>
<td>25</td>
<td>26</td>
<td>71</td>
<td>75</td>
<td>348</td>
<td>368</td>
</tr>
<tr>
<td>PPU</td>
<td>26</td>
<td>26</td>
<td>98</td>
<td>98</td>
<td>303</td>
<td>303</td>
</tr>
<tr>
<td>Engineering Costs [k€/year]</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Engineering and design</td>
<td>6</td>
<td>6</td>
<td>20</td>
<td>21</td>
<td>78</td>
<td>80</td>
</tr>
<tr>
<td>Purchasing and construction</td>
<td>7</td>
<td>7</td>
<td>22</td>
<td>23</td>
<td>85</td>
<td>87</td>
</tr>
<tr>
<td>Total CAPEX</td>
<td>63</td>
<td>64</td>
<td>211</td>
<td>216</td>
<td>813</td>
<td>838</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>OPEX [k€/year]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Personel</td>
</tr>
<tr>
<td>Maintenance</td>
</tr>
<tr>
<td>Insurance</td>
</tr>
<tr>
<td>biomass</td>
</tr>
<tr>
<td>Electricity</td>
</tr>
<tr>
<td>Total OPEX</td>
</tr>
<tr>
<td>Total cost</td>
</tr>
<tr>
<td>Hydrogen production [Ton/year]</td>
</tr>
</tbody>
</table>
Techno-economic analysis of UNIFHY hydrogen production system

<table>
<thead>
<tr>
<th>Hydrogen production cost [€/kg]</th>
<th>13</th>
<th>9.8</th>
<th>7.6</th>
<th>5.8</th>
<th>3.3</th>
<th>2.6</th>
</tr>
</thead>
</table>

Table 12 Global cost of plant under different S/B ratios and sizes (H₂ production only)

The other scenario which can be adapted under S/B: 0.5 is the state that surplus of offgas is turned into both heat and electricity via ICE. In this case we have a direct reduction of hydrogen production cost owing to the cutting cost of the electricity consumption and an indirect reduction of hydrogen cost owing to the revenue from the surlplus of electrical and thermal energy but we have to take into account the greater CapEx because of the ICE cost.

<table>
<thead>
<tr>
<th>Hydrogen production cost [€/kg] considering cutting cost of the electricity consumption</th>
<th>100 kW&lt;sub&gt;th&lt;/sub&gt;</th>
<th>1 MW&lt;sub&gt;th&lt;/sub&gt;</th>
<th>10 MW&lt;sub&gt;th&lt;/sub&gt;</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>12.4</td>
<td>7.1</td>
<td>2.9</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Thermal energy [kW]</th>
<th>20</th>
<th>200</th>
<th>2000</th>
</tr>
</thead>
<tbody>
<tr>
<td>Electrical energy [kW]</td>
<td>10</td>
<td>100</td>
<td>1000</td>
</tr>
<tr>
<td>Cost revenue&lt;sup&gt;2&lt;/sup&gt;[€/kg]</td>
<td>2</td>
<td>1.5</td>
<td>0.2</td>
</tr>
<tr>
<td>Total estimated cost [€/kg]</td>
<td>10.4</td>
<td>5.6</td>
<td>2.7</td>
</tr>
</tbody>
</table>

Table 13 Hydrogen production cost and extra thermal and electrical energy revenues (CHP mode)

The 1 MW plant is able to produce 100 kW<sub>e</sub> and 200 kW<sub>th</sub> from residual offgas (see D6.4 & D6.5). This energy production is conservative because we take into account only the thermal energy from the off gas ICE (so not from the other thermal output of the plants, e.g. cooling of the syngas before PSA, etc) and because we consider nominal power auxiliaries consumption and the off gas useful (for all the energy pants needs and for the energy production) with a gasifier cold gas efficiency of 74%. The energy produced from the other plant sizes is considered proportionally to the size owing to the fact that the syngas flow is almost proportional and the

<sup>2</sup> ICE cost 1500 €/kWe for 10 kWe, 1000 €/kWe for 100-1000 kWe Electricity selling price 0.05 €/kWh for 700-7000 MWh/a (1-10 MWth), 0.20 €/kWh buying price for 70 MWh (0.1 MWth) [29], thermal energy buying price 0.08 €/kWh for 140 MWh (0.1 MWth), selling price 0.04 €/kWh (1-10 MWth) up to 1400 MWh [30].
efficiency of the ICE and the auxiliaries does not change too much within size. Considering potential of power plant to produce electricity, it will definitely be able to meet its electricity demand and totally cut related cost. In order to better show the impact of the energy produced on the hydrogen cost, we subtract the energy revenue (calculated per kg of hydrogen) to hydrogen production cost. As a result, cost of hydrogen can drop by 18%, 26% and 20% for 10 MW\textsubscript{th}, 1MW\textsubscript{th} and 100 kW\textsubscript{th}, respectively.

The main drop is ascribed in 1 MW\textsubscript{th}, because the most important parameter which influences the final cost comes from the low ICE cost (per unit of power in output), moreover since that beyond 1 MW\textsubscript{th} of size it has been fixed a maximum energy quantity delivered to the users of 1400 MWh (equal to the one produced by the 1 MW\textsubscript{th}). Indeed, is unlike that 10 MW\textsubscript{th} can be able to take advantage from thermal revenue. Thus, even if the ICE price is lower, has not the main cost drop. The 100 kW\textsubscript{th}, even if benefits of both higher electricity and thermal selling cost (buying price), has not the main drop owing to the higher ICE cost. Therefore the 1 MW\textsubscript{th} size is able to reduce more the cost respect to the other size.

Furthermore, also in this case, as common for power plants, increasing the size of power plant (0.1-1 MW to 10 MW), the OPEX overcomes the CAPEX as the more important cost item. This is due to the almost proportional increase of biomass and electricity consumption and the less than proportional increase of CAPEX (i.e. increasing the size the CAPEX, per kW, decrease so the bigger is the most convenient). That is why reduction in electricity cost of 10 MW size power plant leads to deeper fall in direct hydrogen production cost (from 3.3 to 2.9 €/kg so a reduction of 12% versus a reduction of about 5% and 7% for 100 kW\textsubscript{th} and 1MW\textsubscript{th}, respectively).

According to Errore. L'origine riferimento non è stata trovata., benefit gained from selling extra energy leads to additional decrease in cost. Under this condition, the cost of hydrogen will range 2.7-10.4 €/kg while improvement in efficiency gives cost ranges from 2.6 to 9.8 €/kg.

Summarising the results it is possible to say that not only, as was obvious, the 10 MW\textsubscript{th} size is identified as the best from both technical (there is also an increase of components efficiency that has not taken into account) and economic point of view, but, more important, both the S/B 0.5 and 1.5 and also at size of only 10 MW\textsubscript{th} can produces hydrogen at affordable and competitive production cost that can be compared to the fossil fuels selling price even without taxes (different from DOE/NREL studies where this competition starts from 50 MW\textsubscript{th} biomass power plants or from the FCH report, see conclusions, where starts from 33 MW\textsubscript{th} and for data at 2030, meanwhile we use actual cost of components and electrical and thermal energy).
Owing to the few difference in the hydrogen cost, the choice between produce also electric and thermal energy (S/B 0.5) or increase hydrogen yields (S/B 1.5) depends on local condition (e.g. existence of thermal load) and goals (importance of efficiency and environmental impacts, i.e. S/B 1.5 where, as showed, the hydrogen production is more efficient and so the impacts on the use of a clean hydrogen are less than the use of syngas, versus releability and profitability, i.e. S/B 0.5 where, as showed, the plant requires less stringent thermal balance and where the electricity and thermal heat can be exploited more easier respect to the actual hydrogen market, see D7.5).

6 Conclusions

The techno and economic performances of the UNIfHY system via the project results have been evaluated. The final discussion embraces all the assumption foregoing mentioned, for instance no CO₂ capture has been considered, UNIfHY steam/oxygen and indirectly heated configurations and three different sizes of the best configuration have been taken into account. As already mentioned in the objective, a comparison of UNIfHY plant with the one for the same hydrogen production pathway analysed in the recent FCH-JU study [20] is here added in order to evaluate, from a general techno-economic point of view, the best plant configuration.

The FCH JU configuration is based on same UNIfHY gasification technology, the Fluidised Bed (FB). In particular, the Gussing technology is a FICFB gasifier, so indirectly heated, thus equal to the indirectly heated UNIfHY configuration. In detail, Gussing has a Bubbling Fluidised Bed (BFB) gasifier and a Circulating Fluidised Bed (CFB) combustor, meanwhile we have two BFB reactors and the gasifier is inside the combustor (thus not only advection but also conduction, convection and radiation heat transfer). Thus it seems that the FICFB is an improvement (less space and better heat management owing to the BFB instead of CFB that increasing the residence time, reduce the volume and also decrease the plant size suitability) respect to the Battle Columbus gasifier (the first indirectly heated fluidised and the base of the DOE/NREL studies on this pathway) that has two CFB reactors.

More interesting is the comparison at plant scheme. In order to do that, in the figure below it is reported again the UNIfHY plant scheme.
Techno-economic analysis of UNIFHY hydrogen production system

Figure 3. Flowsheet of the plant evaluated in this study

While hereinafter is shown the FCH plant scheme.
Even if this diagram is not complete like the previous, it provides worthwhile information concerning plant solutions. Both configurations have their own advantages and disadvantages. Regarding the plant just after the gasifier the main differences are listed below.

**UNIfHY**
- **Catalytic hot candles**: provide a dual effect, tars are converted in additional syngas and then particulate is directly removed in the freeboard of the gasifier.
- **ZnO guard bed reactor**: reduces H$_2$S traces in the syngas below the accepted concentration limit for the low-temperature water gas shift reactor (LT-WGS).
- **LT-WGS**: based on ceramic alumina foams and using copper catalysts assure a further hydrogen production limiting pressure drops.

**FCH**
- **Precoat material**: is necessary to avoid condensation of tar compounds directly on the filter bag which could lead to plugging or even to damage of the filter cloth.
- **Filter cloth**: captures the particles and tars come from cyclone, is a well-known gas cleaning technology.
Techno-economic analysis of UNIFHY hydrogen production system

- **Fatty acid methyl esters scrubber**: used as a washing agent for tar removal (well-known gas cleaning technology).
- **Pressurized water scrubbing**: for CO\(_2\) removal before the PSA.
- **Additional methane reformer of the PSA off gas**.

Summarizing the two plant schemes, it is immediately highlighted that the main difference is that UNIfHY uses hot gas cleaning technologies meanwhile FCH uses cold gas cleaning. The cold gas cleaning technologies do not convert tar in further H\(_2\) and requires a cooling step and more space and equipment even if it is less expensive and more reliable. Of course, cloth filters and FAME scrubbers are two well consolidated technologies, as a consequence their applicability has been already proved also at industrial scale, meanwhile UNIfHY hot gas cleaning technologies have showed very good results only at laboratory scale, because, owing to the not attainment of the right conditions during pilot tests, have “failed” their tests at the pilot-industrial scale (see Deliverables of WP2 for laboratory results and D4.2 and D5.2 for pilot tests).

On the other side, UNIfHY lacking of CO\(_2\) removal system means that PSA has to separate a gas with low H\(_2\) content than FCH plant, as result its separating performances decrease, at the same time it must be oversized in order to accept a higher flow rate. Anyway the addition of CO\(_2\) removal system can be integrated in the same way also in the UNIfHY plant scheme. Also here the choice of a presurized water scrubbing for CO\(_2\) removal is a choice of a commercial technology with less efficiency than an advanced technology. E.g. the choice of CO\(_2\) sorbent inside the reactors (using the gasifier and combustor reactors for the sorption and desorption cycle, as in AER-GAS project, i.e. at Gussing) would be better not only owing to the reduction of the CO\(_2\) flow directly from the gasifier instead of before the PSA but overall owing the shift of the thermodynamic equilibrium reactions towards more hydrogen production.

Finally, the presence of a reformer reactor surely increases hydrogen yield but, as UNIfHY simulations show, see e.g. D6.4, this can be “better” placed within the syngas (between gasifier and WGS) not the PSA off gas line. Indeed, as CO\(_2\) removal, it is better to convert the flow before and not after, in order to not oversize and to increase PSA efficiency, moreover a reformer (SMR, Steam Methane Reformer) requires steam and high temperature thus it is less energy expensive use the steam already present in the syngas and the syngas temperature, of course the temperature of the syngas maybe still have to be increased and the reforming would be as the WGS “innovative” working at “atmospheric” and not at high pressure (similar
advantages can be also respect ATR, Auto thermal Reformer, or respect POR, Partial Oxidation reformer).

Anyway, the CO₂ sorbent (that release heat during the sorption process inside the gasifier and requires heat for desorption process in the combustor, thus improving the heat transfer but challenging the heat balance) and the reformer have to be carefully checked within the energy plant balance (e.g. at S/B 1.5 all the syngas is used for the internal energy balance, thus it has to be checked the more or less advantage of increasing hydrogen yield via increase S/B and/or via CO₂ sorbent and SMR).

Here below are quoted the FCH plant data compared with the UNIfHY plant data.

<table>
<thead>
<tr>
<th>Unit</th>
<th>3 MW_{H2,LHV}</th>
<th>9 MW_{H2,LHV}</th>
<th>33 MW_{H2,LHV}</th>
</tr>
</thead>
<tbody>
<tr>
<td>Applicability</td>
<td>local</td>
<td>semi-central</td>
<td>central</td>
</tr>
<tr>
<td>Production capacity t_{H2/day}</td>
<td>0.2-4</td>
<td>4-20</td>
<td>&gt;20</td>
</tr>
<tr>
<td>Wood chips</td>
<td>MW</td>
<td>6.59</td>
<td>17.5</td>
</tr>
<tr>
<td>Electricity consumption MW</td>
<td>0.12</td>
<td>0.31</td>
<td>5.6</td>
</tr>
<tr>
<td>Output</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Hydrogen</td>
<td>MW</td>
<td>3.00</td>
<td>9.00</td>
</tr>
<tr>
<td>H₂ Chemical Efficiency %</td>
<td>46%</td>
<td>51%</td>
<td>67%</td>
</tr>
<tr>
<td>Economic data</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>CAPEX M€</td>
<td>14.6</td>
<td>32</td>
<td>100</td>
</tr>
<tr>
<td>O&amp;M, repair, insurance, overhead % of CAPEX/yr</td>
<td>6.75% of CAPEX/yr</td>
<td>6.75% of CAPEX/yr</td>
<td>3% of CAPEX/yr</td>
</tr>
<tr>
<td>Lifetime year</td>
<td>20</td>
<td>20</td>
<td>20</td>
</tr>
<tr>
<td>Equivalent full load period h/yr</td>
<td>7,500</td>
<td>7,500</td>
<td>7,500</td>
</tr>
<tr>
<td>H₂ generation cost €/kg_{H2}</td>
<td>5.8</td>
<td>4.1</td>
<td>3.3</td>
</tr>
<tr>
<td>GHG g_{CO2/kWh_{H2}}</td>
<td>96</td>
<td>96</td>
<td>112</td>
</tr>
</tbody>
</table>

Table 14 results obtained based on FCH report [20]

<table>
<thead>
<tr>
<th>Unit</th>
<th>0.05 MW_{H2,LHV}</th>
<th>0.5 MW_{H2,LHV}</th>
<th>5 MW_{H2,LHV}</th>
</tr>
</thead>
<tbody>
<tr>
<td>Applicability</td>
<td>local</td>
<td>local</td>
<td>semi-central</td>
</tr>
<tr>
<td>Production capacity t_{H2/day}</td>
<td>0.036</td>
<td>0.36</td>
<td>3.6</td>
</tr>
<tr>
<td>Biomass</td>
<td>MW</td>
<td>0.1</td>
<td>1</td>
</tr>
<tr>
<td>Electricity consumption MW</td>
<td>0.007</td>
<td>0.066</td>
<td>0.661</td>
</tr>
<tr>
<td>Output</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Hydrogen</td>
<td>MW</td>
<td>0.05</td>
<td>0.5</td>
</tr>
<tr>
<td>H₂ Chemical Efficiency %</td>
<td>50</td>
<td>50</td>
<td>50</td>
</tr>
<tr>
<td>Economic data</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>CAPEX M€</td>
<td>0.8</td>
<td>2</td>
<td>7</td>
</tr>
<tr>
<td>O&amp;M, repair, insurance, overhead % of CAPEX/yr</td>
<td>2% of CAPEX/yr (DOE)</td>
<td>2% of CAPEX/yr (DOE)</td>
<td>2% of CAPEX/yr (DOE)</td>
</tr>
</tbody>
</table>

March 2016
Techno-economic analysis of UNIFHY hydrogen production system

<table>
<thead>
<tr>
<th>Lifetime</th>
<th>Equivalent full load period</th>
<th>H2 generation cost</th>
<th>GHG</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>year</td>
<td>20</td>
<td>20</td>
</tr>
<tr>
<td></td>
<td>h/yr</td>
<td>7000</td>
<td>7000</td>
</tr>
<tr>
<td></td>
<td>€/kgH2</td>
<td>9.8</td>
<td>5.8</td>
</tr>
<tr>
<td></td>
<td>g CO2/kWhH2</td>
<td>48.24</td>
<td>48.24</td>
</tr>
</tbody>
</table>

Table 15 Techno-economical results of UNIFHY Indirectly heated technology 1.5 S/B

The FCH selected three process scales: 3, 9, 33 MWH2, LHV aimed to supply a single refueling station or multiple of them respectively. Indeed, standard refueling station is expected to be about 0.2-4 tH2/day, production plant between 0.5-5 MWH2, LHV. But today there is lacking of deep penetration into the mobility market of FC-vehicle thus the actual HRS range is smaller, 0.05-0.4 tH2/day, and the hydrogen market at small scale is smaller (see D7.5, e.g. welding, etc). Therefore, it seems more appropriate a size of 0.05-0.5 MWH2, LHV, as local, not to mention that, as described before, a smaller plant is better able to exploit all the biomass availability throughout the territory. The choice to not overcome 5 MWH2, LHV size depends on many factors. First the hydrogen production cost already become competitive at this size, second it is more difficult find sites that have availability of local biomass at larger size, third the environmental balance and social acceptance of larger power plant is more challenge. For all these reasons we consider only 5 MWH2, LHV as maximum size, even if, of course, when there is biomass availability and the environmental impacts can be kept low, larger power plants (up to 100 MWH2, LHV, thus reaching the greatest world biomass size pants of more than 200 MWth input, i.e. about 400,000 annual tons of biomass, more than 50 trucks of 20 tons per day!) will have economic greater performance.

The electric power consumed by auxiliaries of UNIfHY has been assumed constant in function of plant size (see chapter 4 and D6.4, indeed the electrical consumption varies but mainly within different configurations respect to different size of same configuration; thus it is possible to consider for 0.1 and 10 MW the same percentage of power consumption respect to power input of 1 MW). Thus the auxiliary electrical power is always equal to 7% of the power input (as similar to DOE study for size > 35 tH2 per day), while FCH study keeps it constant only for 3 and 9 MW (2%), whereas for 33 MW is equal to 11%. The FCH study does not quote any explanations about these values, anyway not only our value is obtained from the real value of the 1 MWth plant and it is in line with the DOE value but overall this value is always below 11 % so its variation (from 2 to 11 %) does not change the global analysis. Regarding the electricity price
Techno-economic analysis of UNIFHY hydrogen production system

FCH use taxed value (in particular 0.166 €/kWh for the two smaller gasification plants and 0.139 €/kWh for the 33 MWH₂, this is also questionable taking into account that the data are estimated for 2030 and compared with the electrolysis hydrogen production cost study where the price of electricity is estimated at 0.06 €/kWh) meanwhile we use untaxed value (fixed at 0.08 €/kWh), thus the difference between the electricity consumption is reduced from an economic point of view (because even if, for similar size the FCH study uses 2% instead of our 7%, so a weight 3 times more in our study, they use a price two times higher, thus the different is reduced within only 70% of more economic expense in our case).

Regarding the third table line, i.e. hydrogen yield and so hydrogen chemical efficiency (MWH₂/MWbiomass) FC reports an increment with increasing plant size: 46%-51%-67%; meanwhile we keep it constant at 50%. FCH study does not quote any explanations about these values that anyway are in line with our values of 5 MW sizes (5 MW are between 3 and 9 MW for which the efficiency is between 46 and 51%). Indeed, the efficiency increases with size, but this has to be carefully checked.

The UNIfHY’s CAPEX fluctuates more than FCH study (from 16 to 1.5 M€, i.e. from 8 to 1 M€/MWH₂ respect from 100 to 15 M€, i.e. from 2 to 2.2 M€/MWH₂), in fact, even if the scaling factor for cost evaluation is lower [2,3], the little sizes of plants taken into account cause a more variability of this data. The UNIfHY’s CAPEX are less than half of the FCH CAPEX for similar size (i.e. 7 M€ for 5 MWH₂ UNIfHY meanwhile, for FCH, are 15 and 32 M€ for 3 and 9 MW). This can be explained by the presence of more components in the FCH plant (additional RME scrubber, Pressurized water scrubbing and reformer). This can also explain the higher FCH OPEX (O&M at 7% and 3% of CAPEX versus the constant 2% of UNIfHY, as DOE/NREL study). Of course, as efficiency, we are aware that the CAPEX and the OPEX change within the size (even if the OPEX decrease less than CAPEX increasing the size), but, in order to put data different from scaling factors of real data of one size and literature data, a detailed analysis based on real data of prototype/commercial plant at each size have to be done.

One of the most important data is the Hydrogen cost, it depends from all the techno-economic parameters previously described. By comparing with FCH report, UNIfHY1000 reaches already the target foreseen by FCH in 2030 (at 5 MWH₂ size) owing to lower CAPEX (per unit of Hydrogen), inferior biomass cost (75 €/t against 90 €/t) and less O&M cost even if it has higher electricity consumption and lower equivalent full load period.
Concerning the GHG emissions, the FCH plants always totalize a value two times higher than UNIfHY. It is referred to the use of two different kinds of biomass. Since woodchip is a product processed, it requires more energy during production and consequently leads to more GHG released than the feedstock like almond shell needs. In fact, Almond shell is a by-product of agro-industrial sector. Therefore, a big proportion of energy is consumed to produce main product (kernel) for food goals. Unlike FCH, UNIfHY considers the energy requirements and GHG emissions resulting from the construction and decommissioning of manufacturing plants. However, it is just 3% of total CO₂ emissions thus its influence is limited. As a result, as described in D 6.5, kind of biomass employed is the main factor that strongly influences total emissions released during biomass hydrogen production.

Regarding the other UNIfHY technology, the steam/Oxygen Gasifier, it has not be disregarded. Indeed, even if it is not less competitive as Indirectly Heated, not only the difference, in efficiency and cost, is low, but overall if a local O₂ production is available, for example combining it with WE, this can be more competitive than IH. Especially at small scale, because oxygen is a by-product and in small-medium scale production plant is not convenient compress it in tanks, hence a decentralized biomass gasification might use this “by-product” as feedstock instead of use an air separation unit, this leads to drop in Hydrogen production cost (Table 11). The integration of both the systems provides further benefits also in operative hours, because electrolysis is more reliable than a single FICFB, and it is able to supply H₂ during the gasifier’s maintenance.

On varying of S/B ratio, the IH gasifier provides different H₂ yields, in particular S/B=0.5 is convenient when you want to keep the plant independent by electric network supply, moreover it works such as CHP plant, nonetheless the H₂ production drops unavoidably, but the revenues coming from heat and electricity selling help the H₂ production cost to be maintained at a low value, just above of S/B=1.5. Unlike the previous one, S/B ratio equal to 1.5 is the right choice when is necessary to fulfill a high H₂ demand. The different S/B has to be choised also in function of the additional components (to increase the efficiency, see above in the plant configuration comparison) evaluating the advantages of increasing efficiency via additional CO₂ removal and reformer respect the disadvantages of increasing CAPEX and OPEX and the disadvantages of use the off gas for these processes instead of use it in order to increase the S/B.

As a preliminary consideration it is possible to foreseen that the best configuration (in term of efficiency and less hydrogen production cost) would be with hot gas conditioning (once the lab
scale results will be confirmed at pilot/industrial scale), CO₂ removal (as sorbent in the gasifier, if this technology will be demonstrated) and reformer (in the syngas line, if ad hoc reformer, as the WGS in UNIfHY will be developed) and so S/B less than 1.5 (in order to maintain the thermal balance and so do not require additional fuel/energy) but simulative, experimental and pilot/demonstration activities, that where not foreseen in this project limited at the configuration without CO₂ removal and reformer, have to be performed.

Finally, the hydrogen production cost must take into account also distribution, compression and the structure of refueling stations, but this topic is mastered in D7.5. Indeed, even if the cost decreases sharply with increasing in plant size, in order to evaluate the real competitiveness these other costs have to be taken into account.

Nowaday, has been demonstrated the IH gasification is able to reach with extraordinarily reliably and regularly 7,000-8,000 operating hours per annum⁸, UNIfHY owns the same potentiality and it can reach the same aim as shown in Table 15 (indeed we use 7,000 equivalent annual hours instead of 7,500 used in the FCH study, mainly because 7,000 is the average equivalent annual hours of the last 5 years of the Gussing plant, overall plant, i.e. with ICE not only gasifier, thus 7,000 is more appropriate for a 2016 evaluation, meanwhile 7,500 is more appropriate for a 2030 study, as the FCH).

In conclusion, after this deep analysis we provide actual and foreseen techno-economic evaluation aimed to clarify the current technology status and future possible improvements.

**Configurations:**

- **Hot gas conditioning** (by means of ceramic candles) implemented in UNIfHY gives major benefits converting in hydrogen tars and then LT-WGS provides further Hydrogen conversion exploiting the remainder thermal flow content, this is the most promising solution if the lab scale results will be replicated at pilot/industrial scale.

- The insertion of **CO₂ removal system** is particularly important, it increases hydrogen yield in the gasifier and the efficiency of the following components and at the same time reduces the following pipeline and components dimensions and costs, if sorbents inside the gasifier are used. FCH study uses a pressurized water scrubber, that have the advantages that it is a well-known equipment but it need of an additional component with further cost and reduce the

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³http://www.gussingrenewable.com
benefits only at the PSA level. Therefore, with a medium long-term vision, a CO₂ capture inside the reactor, by means of sorbents, will be in future the best way to obtain a significant rise in global plant performance. In fact, they shift the reaction inside the gasifier enabling the hydrogen yield to increase and improving the heat balance between gasifier and combustor reactors (less circulation of inert material), in addition, thanks to the sorbents the whole process could be integrated into the reactor (as catalytic candles versus scrubber) promoting a plant simplification.

- Similar consideration can be done for the additional methane reforming.
- In general biomass power plant demonstrates an interesting flexibility to fulfil user requirements, the simulations and the economic assessments carried out on UNIfHY ensure a competitive H2 production cost also in cogeneration mode using an ICE. Moreover, the 1 MWₘₜ plant in CHP mode gives a lower H2 cost than the same plant with H₂ production only (owing to the exploitation of all the excess of thermal energy not only electricity and the reduced hardware cost).

**Efficiency:**

After several lab scale, simulation and pilot scale activities we can affirm that UNIfHY configuration is able to reach 50% of hydrogen chemical efficiency based on the biomass input (without additional energy input). Obviously, only a larger demonstration activites (in UNIfHY not foreseen) may highlight the shortcomings which prevent the efficiency maintaining stable during all the plant operativity (this, toghether with the replications of the lab scale hot gas conditioning and WGS performance, has been an issue during the 1 MW plant tests). The 2011 BTH topic target was 70% for all plant sizes, but the 2015 FCH study report for this technology 50% of efficiency. Indeed, 70% efficiency can be a target of very advanced BTH plants integrating all the possible efficiency improvements at the most advanced solution (hot gas conditioning, CO₂ sorption/desorption, innovative reformer, etc) and at larger size.

**Size:**

Even though DOE and FCH indicate the centralized production as the best choice for biomass gasification because it leads to high competitiveness with other technologies. In our opinion this choice contrasts with a real exploitation (provided it is used waste biomass), because this feedstock, widespread in the territory, is not convenient to harvest it in big production plant (because of low energy density and perishability). For this reason UNIfHY

March 2016
is conceived for decentralized production, 0.1 MW\textsubscript{th} to 1 MW\textsubscript{th} is a local production without any distribution cost in order to get the maximum competitiveness, while 10 MW\textsubscript{th} is a blended size (semi-local production), it can refill a bug utility and/or different utilities via distribution.

**Operation hours – durability - availability:**

The 2011 topic target was durability > 10 years (80,000 h) with availability > 95% (i.e. 7,600 equivalent annual hours, i.e. at nominal power). The UNIfHY lab, simulative and pilot activities showed that the plant can have 10-20 years with maintenance items (filter candles, catalysts, valve seals) thus 7,000 hours seems more appropriate, also taking into account Gussing availability. Further research and demonstration activities have to be done in order to confirm these data. Durability and availability targets, as efficiency target, have to be different for centralised and distributed generation and for different configurations (i.e. detailed for each componts).

**Cost:**

The final H2 production cost is heavily dependent by plant size, obviously the larger sizes are the best solution, provided that the needed biomass quantity is available and its use is sustainable, see above. The small size plants even if they have greater and so not competitive production costs (greater than the fossil fuel selling price without taxes) can have affordable H2 cost, comparable with current production technologies, as described in D7.5, taking into account the distribution costs
7 References

Techno-economic analysis of UNIFHY hydrogen production system