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Techno-economic analysis of hydrogen production using biomass gasification -A small scale power plant study

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Abstract

Hydrogen has the potential to be a clean alternative to the fossil fuels currently used. This is especially true if hydrogen is manufactured from renewable resources such as biomass. However, hydrogen from biomass faces techno and economic challenges especially in the small size required for the decentralized hydrogen production. In this purpose, a techno economic analysis was carried out on small scale (100kWth) system. The plant is mainly composed of gasifier (double bubbling fluidized bed reactor) coupled with a Portable Purification Unit (PPS: catalytic filter candles, Water Gas Shift and Pressure Swing Absorption). This work focuses on system costs to identify barriers to the development of this technology. A sensitivity analysis was conducted to study hydrogen production cost as a function of capital cost, operating cost and hydrogen production efficiency. The results showed that although efficiency of the production system is the main factor to fall production cost, it cannot be able to reduce costs to favorable level alone. In other words, PPS cost recognized as the major cost is requisite to go down. Therefore, the 50% reduction of PPS cost and the variation of steam to biomass from 1 to 1.5 allow the special cost to fluctuate between 12.75-9.5 €/kg.

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1. Introduction

Biomass has the potential to accelerate the realization of hydrogen as a major fuel of the future. The main reason lays in the large availability of biomass resources (wood and wood waste, agricultural crops, organic fraction of municipal solid waste, residues from agro-industrial and food processes, aquatic plants such as algae and waterweeds) that can be used for energy production. Hydrogen's share in the energy market is increasing with the implementation of fuel cell systems and the growing demand for zero-emission fuels. Hydrogen production will need to keep pace with this growing market. Hydrogen production from biomass gasification can be a promising technology in terms of environmental impact and economic feasibility. For the environment, hydrogen produced from biomass gasification contributes to an almost zero net CO_2 emissions since the feedstock is a renewable resource which consumes CO_2 during its growth and emits CO_2 when it is gasified.

Gasification is one of the three main thermochemical process solutions to extract energy from biomass. Gasification converts biomass in a combustible gas mixture (called product gas or syngas), mainly made of carbon monoxide, hydrogen and lower content of methane and able to provide a wide range of products, extending from clean fuel gas and thermal and electrical energy to bulk chemicals [1],[2]. Several gasification technologies are available today and fluidization is the most promising among all of them, for a series of reasons, among which excellent gas-solid mixing and large thermal inertia of the bed which gives higher throughput and biomass conversion; possibility to use different fluidizing agents, to inject reagents along the reactor height, to operate with or without a specific catalyst and to be fed with different feedstocks [3],[4]. Moreover, as gasification agents, steam/Oxygen/CO2 gasification are to be preferred respect to the air gasification owing to the higher content of hydrogen in the product gas and to the less inert flow (i.e. Nitrogen) to be treated in the purification units [5].

The key to develop this technology is to overcome the problems associated with technical and economic aspects of the pure H_2 production and, especially for large size plants, environmental acceptance. For this reason, especially the small size, once viable from a technical point of view, requires to be economically achievable. Techno-economic analyses are the only way to make rational selection of appropriate research and development paths in this complex and rich technical area. This work presents outcomes of the system carried out in the UNIFHY project

Hydrogen production costs based on biomass gasification have been estimated by previous studies. Mohamed et al. [6] via a fluidized bed air gasifier of empty fruit bunch, estimated 2.11 USD/kg (considering a biomass cost of 12.32 $/t, a feeding rate of 6 kg/h and a hydrogen flow of 0.311 kg/h, thus an efficiency of 33%, LHV H2 / LHV biomass). Thus, not taking into account H_2 purification and having low biomass cost, Mohamed et al. show that even small size, air gasification and low efficiency give low hydrogen cost (comparable with fossil fuel production cost).

Moneti et al. [7] analysed 1 MWth indirectly heated gasification with catalytic filter candles, WGS (water gas shift) at 400 °C, WGS at 200 °C, and PSA. Effects of steam to biomass ratio (from 0.5 to 2) and temperature (from 750 °C to 850 °C) in hydrogen conversion efficiency has been assessed. The sensitivity analysis showed that, at S/B=2 and 850 ºC, a maximum hydrogen/biomass efficiency of 70% can be reached. Pallozzi et al. [8] analysed a similar 1 MWth input biomass power plant but with only one WGS at 300 °C. The hydrogen production cost, under gasification temperature of 850 °C, S/B=2.0, was estimated as 8.3 €/kg. Thus, when the H_2 purification is taken into account the cost increase, challenging the small size reliability and convenience.

Lv et al. [9] evaluated a downdraft biomass oxygen gasification and CO-shift at atmospheric pressure and determined 1.69 USD/kg H_2 production cost. Biomass cost and feeding rate considered are 39 USD/kg and 266.7 kg/h meanwhile the hydrogen efficiency was 51.5%. The cost sensitivity analysis on this system revealed that electricity cost because of the existence of PSA and catalyst cost due to their short life time (250 h) are the two most important factors to impact hydrogen production cost. Thus, when H_2 purification is taken into account the purification cost becomes the more significant even if here low biomass cost, bigger size and relative high efficiency allow obtaining a low hydrogen production cost.

Inayat et al. [10] designed a heat integrated flowsheet for the production of hydrogen from oil palm empty fruit bunch using a steam gasification in a fluidized bed with in-situ CO_2 capture. At temperature of 1150 K, S/B of 4 and sorbent/biomass ratio of 0.87, H_2 yield of 0.0179 kg/h (purity of 79.91 mol %) and H_2 cost of 1.91 USD/kg has been obtained. They found out considerable saving can be obtained for steam production using heat integration application as there is a large amount of available waste heat from the gas cleaning and cooling units. Here a low hydrogen production cost is obtained but the H_2 efficiency, taking into account the global energy needed, decrease at less than 25% showing that reliability, energy consumption and cost of the CO_2 capture is a challenge yet.

Furthermore, there are many previous studies regarding techno-economical estimation of biomass power plant which mostly focus on large scale gasifiers. DOE (Department of Energy) and NREL (National Renewable Energy Laboratory) have published some reports on cost of H_2 production by large scale biomass power plant [11],[12],[13]. In these reports feedstock and CapEx (capital expenditure) were identified as the most important costs.

Bocci et al. [14] have assessed economic feasibility of a small scale (100 kWth) steam gasification fluidized bed and hot gas conditioning system by NPV (Net Present Value) and PBP (Pay Back Period). In other research by Bocci et al. [5] different biomass feedstock from cost and environmental point of view were examined while they were being applied by the state of the art small scale gasifier under two major gasifiers topologies (fixed and fluidized bed). Finally, high efficiency examples of power production by means of internal combustion engine, micro gas turbine, Solid Oxide Fuel Cell or a mix of them, both as realized plants and process simulated ones, have been then reported. According to results, the combination of fluidized bed, hot gas conditioning and mGT-fuel cells increase the electrical efficiency but the global capital cost rises as well. Villarini et al. [15] carried out a feasibility study on biomass energy exploitation to satisfy farm energy consumption. Olive pruning was recognized as a very appropriate feedstock to provide electric and thermal energy demand of farm in the area studied. Fracaro et al. [16] have evaluated the feasibility of a 100 kWe gasification system (fixed bed) including an engine generator set. The parameters that showed to have a greater impact on the levelized unit cost of electricity delivered were the load factor, the gasifier capital cost, the electric conversion efficiency, the capacity utilization factor and the gasifier useful lifetime. Also, a techno-economic analysis investigates the advantages of small scale plants in the range of 100–600 kWe with a comparison between different design configurations for industrial applications of biomass gasification [17].

The review of literature implies there are not any evaluations which reliably targets pure H_2 efficiency and actual costs of continuous hydrogen production from gasification particularly at small size of power plants. Therefore, this study aims at economic assess an innovative power plant for the small scale industrial application of H_2 production. During the UNIfHY project [18], has been developed a 100 kWth prototype composed by an indirectly heated fluidized bed gasifier with catalytic filter candles inserted in the freeboard and a Portable Purification Unit (PPS) composed of a ZnO, WGS and PSA reactors. The output of energy analysis of the system developed in [19] has been used as input of the present economic analysis. In addition, a cost sensitivity analysis is carried out to recognize the major components influencing the specific cost of hydrogen production.

2. Plant description

Figure 1 shows the plant scheme.

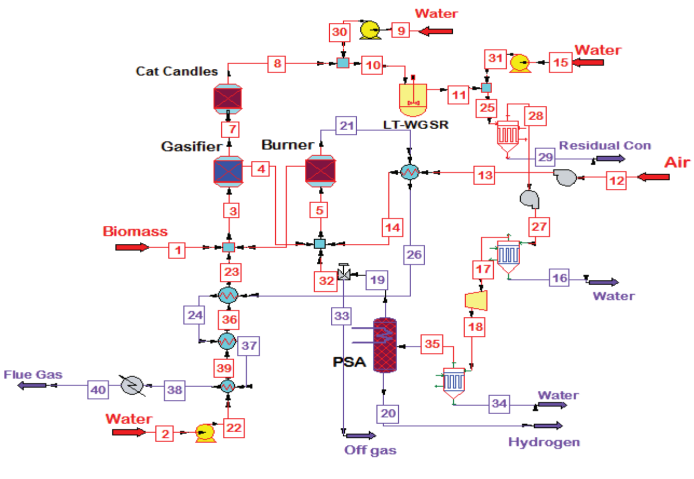

Fig. 1. Flow sheet of the plant evaluated in this study.

As soon as fed into the gasification zone, biomass is gasified with steam. The bed material (olivine), together with some charcoal, circulates to the combustor which is fluidized with hot air and the charcoal is burned to heat up the bed material to a temperature that is higher than the one of the feeding. The hot bed material from the combustor is circulated back to the gasifier supplying the thermal power needed for the gasification reactions. Off gas from PSA is also burned into combustor to provide extra heat demanded by the gasification process, especially at high S/B. Tars are converted by Catalytic filter candles (CFC) that remove particulate in the freeboard of the gasifier.

The composition of gas from WGS is primarily H_2, CO_2, residual steam, traces of CH_4 and CO. Once cooled, compressed and cooled to ambient temperature, the gas is fed to the PSA where pure H_2, is obtained. The off gas is employed in the combustor as was described. The heat of the flue gas from the combustor is used to heat the air, overheating the steam, produce steam and finally to heat the water and released to the environment. The operating conditions of the whole power plant are brought in table 1. Data are simulated based on experimental data at different conditions (in particular gasifier and combustor temperature at experimental conditions reaches 800 ℃ and 850 ℃ respectively, and S/B reaches 0.5, see D4.3 and D6.4, where the models of the gasifier and other components are validated[19,20]). Almond shell as dry (not ash free) with LHV 18 MJ/kg [21] is feedstock used in the process. Indeed, lignocellulosic biomass can be assumed equivalent (same LHV on a dry basis, similar ashes melting point, bulk density, etc.), meanwhile RDF have content of sulphur and chlorine elements ten times higher (about 0.4 versus 0.04 %w dry) thus almond shells have been chosen to be used owing to the lower price respect pellets and greater bulk density versus wood chips.

Table 1. Operating conditions of gasifier under two different S/B

Parameters	S/B (1)	S/B (1.5)
Biomass feeding rate (kg/h)	20	20
Steam feeding rate (kg/h)	20	30
Electricity consumption (kW)		
Start up	0.4	0.4
Process water pump	0.0012	0.002
Deionised water pump 1	0.042	0.052
Deionised water pump 2	0.027	0.027
Air blower	1.4	1.58
Syngas blower	0.62	0.66
Compressor	4.1	4.36
Gasification T(ºC)	850	
Burner operating temperature (ºC)	950	
Olivine sand circulated between combustor andgasifier (kg/h)	1000	
Burner and gasifier operating pressure (bar)	1.1	
PSA inlet pressure (bar)	7	
PSA intercooler compressor efficiency%	62	
PSA intercooler compressor temperature (ºC)	40	
Intercooler compressor stages	2	
WGS inlet temperature (ºC)	300	
Air blower pressure ratio	1.3	
Air blower efficiency	40	
Water pump pressure ratio	3	
Water pump efficiency	80	

The hydrogen chemical efficiency (η) has been calculated by the following equation (based on lower heating value, LHV, or higher heating value, HHV):

$$\eta_{LHV\,(or\ HHV)}[\%] = \frac{\dot{m}_{H2} * LHV_{H2}\,(or\ HHV_{H2})}{\dot{m}_{bio,daf} * LHV_{bio,daf}\,(or\ HHV_{bio,daf})} \times 100 \tag{1}$$

Where $\dot{m}_{H2}$ is the mass flow rate of the produced hydrogen and $\dot{m}_{bio.daf}$ is the mass flow rate of biomass feedstock in dry ash free.

Yield of hydrogen under S/B 1 and 1.5 was obtained 1.38 and 1.52 kg/h, respectively. Therefore, according to biomass feeding rate in table 1, hydrogen efficiency is in the range of 46 - 50%. This value is in line with the recent FCH-JU study[22]. Indeed, 50% is a maximum for this configuration, i.e. it requires that all the off-gas from PSA is sent to the combustor in order to support the gasification reactions and to heat the steam (also the flue gas from the combustor is almost fully used to the internal plant heat requirements), as showed in the Figure 1. Obviously if other improvements are taken into account (e.g. CO_2 sorbents inside the gasifier to shift the thermodynamic equilibrium and decrease the gas flow to be treated or additional methane reformer reactor or CO2 capture system) greater hydrogen chemical efficiency can be reached but the plant complexity will increase.

3. Economic analysis hypothesis

Costs are split in Capital expenditures (CapEx) and Operating expenses (OpEx). CapEx is divided into two sections: Hardware costs and Engineering costs. The engineering costs includes engineering and design(13% total installed cost [12]) and purchasing & construction (14% total installed cost [12]).

3.1. CapEx

The CapEx can be depreciating within N years, N depending on three main parameters, namely the lifetime duration of the hardware, considering the maintenance quoted in the OpEx costs; the long term agreement for feedstock procurement and the long term agreement for green-hydrogen off take. As a first assessment, we consider 20 years of depreciation. That means that in the targeted business models, only locations where feedstock procurement and H_2 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:

$$\text{k€/year} = CapEx\ in\ k€ * 0.07/(1-(1+0.07)^\wedge(-20)) \quad (2)$$

3.2. OpEx

Operating expenses for the plant covers the cost of Maintenance costs,2 % of total CAPEX [12], Insurance and taxes,2% of total CAPEX [12], Biomass, 40-75 €/ton [23] and Electrical energy, 0.08 €/kWh [24]. Annual working hours was considered 7000h VAT free.

PPS component is fully automated. The control unit cost of this component is included [24]. Being the 100 kWth gasifier actually operated manually, the control cost is estimated by using Equation 3 and based on the control unit cost of 1 MWth gasifier (UNIfHY 1000) which used as reference cost. Although 1MWth gasifier is not fully automated, due to the fact that the control cost of 1 MWth gasifier is an actual prototype cost and that the scaling factor is low, the cost obtained can be reasonable for an automated system.

$$SC = RC * \left(\frac{SP}{RP}\right)^{EXP} \quad \ldots\ldots\ldots\ldots \quad (3)$$

Where SC is Scaled cost , Exp is Exponent: 0.13 [25], RC: Reference cost: 134.95 k€ [24] , RP: Reference Parameter: size of reference gasifier (1000kWth), , SP: Scaling parameter: size of considered gasifier (100kWth)

4. Results and discussions

Hardware cost of power plant including gasifier system and portable purification system (PPS) cost is estimated 370.96 and 270.63 k€, respectively. The cost analysis shows that the costs of control unit and gasifier with filter candles inserted in the free board have the biggest proportion of gasifier CapEx while in PPS, PSA reactors represents the highest cost. According to current conditions of plant, the annual total cost of 100 kWth gasifier integrated with PPS was calculated by using equation 2. The most major costs are gasifier and PPS. This result also was obtained for

1 MWth BFB gasifier in the Unifhy project while personnel cost PPS and gasifier were identified as costly ingredients for that size of gasifier. In our case, the small scale, personnel cost are not affordable and a full automation of the system has to be implemented. Thus, no personnel cost has been considered. In OpEx analysis, the maintenance and insurance costs have the large proportion of cost since they are influenced by the high capital cost.

Owing to the fact that at S/B: 1.5 only hydrogen yield changes and all cost (except electricity cost mentioned in table 1 and 3) are constant, it is avoided to present a table of cost at S/B: 1.5).

Table 2 Total cost of the plant

Total cost (k€/year)	
Hardware cost	
Gasifier system	35
PPS	25.54
Engineering cost	
Engineering and design	7.87
Purchasing and construction	8.47
Total CAPEX	*76.91*
Maintenance	16.29
Insurance	16.29
Biomass	10.5
Energy	3.7
Total OPEX	46.79
Total cost	*123.7*
Hydrogen production (Ton/year)	9.7
Hydrogen production cost (€/kg)	12.75

5. Sensitivity analysis

In order to determine cost sensitivity to H_2 efficiency, the steam to biomass ratio (S/B) between 1 and 1.5 was varied under operating conditions presented in Table1. Meanwhile, the PPS as second high cost and prototype component have been altered to analyze total cost sensitivity. The aim of PPS cost reduction as an achievable target is evaluation of the influence of PPS cost on hydrogen production cost. Since PPS cost has been estimated for Prototype it can decrease per increase in the number of units. Manufacturing 5-10 more unit of PPS results in 44-50% decrease in cost for each unit [19]. On the other hand, standardization of components can reduce this cost more intensively. That is why fall in cost of PPS as a feasible and accessible way to drop in total cost is considered in this article. Three scenarios can be defined namely; base scenario 'worst' S/B: 1 and no PPS cost decrease (table 2), scenario A 'middle' describes plant costs at S/B:1 while PPS cost has a 50% decrease and scenario B 'best' which relates costs according to S/B:1.5 and a 50% decrease in PPS cost. Tables 3-4 indicate the effects of alleviation in costs. The cost sensitivity is the change in total cost that comes from decreasing one unit of input cost (CapEx and OpEx).

Table 3. Cost changes under scenario A, B

	S/B=1	S/B=1.5
Hardware cost (k€/year)		
The percentage of cost decrease	50% PPS	50% PPS
Gasifier	35	35
PPS	12.77	12.77
Engineering Costs (k€/year)		
Engineering	6.21	6.21
Construction	6.7	6.7
Total CapEx	*60.7*	*60.07*
Maintenance	12.85	12.85
Insurance	12.85	12.85

Biomass	10.5	10.5
Energy	3.7	4
Total OpEx	*39.9*	*40.2*
Total cost	*100.6*	*100.9*
Hydrogen production (ton/year)	9.7	10.6
Hydrogen production cost (€/kg)	10.37	9.5

According to table 2 and table 3 costs, cost sensitivity has been calculated and results have been provided in table 4. The purpose of this analysis is to determine the level of sensitivity of total cost to the effective input cost.

$$\text{Cost sensitivity: } \frac{\Delta y}{\Delta x} \bullet \frac{x}{y} \tag{4}$$

Where, Y: H_2 specific cost and X: input cost (PPS cost for scenario A and H_2 efficiency for scenario B).

Sensitivity cost shows that changes in total cost are highly influenced by changes in PPS cost, due to the fact that regarding the considered assumptions in part 3, PPS cost is included in hardware cost which impacts directly on engineering costs and OpEx.

Table 4. Costs and cost sensitivity under Scenario A and B

	PPS Cost (k€)	% on total cost	Total cost (k€)	Specific cost of H2 (€/kg)
Basic cost	25.54	21%	123.7	12.75
Scenario A	12.77	13%	100.6	10.37
Scenario B	12.77	13%	100.9	9.5
Cost sensitivity	Scenario A	0.36	Scenario B	0.9

Cost sensitivity based on H_2 efficiency shows that cost and performance change nearly at the same rate. Therefore, it is practical to decline about 1% of H_2 cost per 1% more efficiency, while in scenario A where PPS cost halves, 18% of specific cost could be cut which leads to a 0.36% decrease in production cost per 1% fall in PPS cost. Therefore, as a result technical efficiency of plant has the most influence on the cost. After 50% reduction in PPS cost, specific cost can reach 9.5 €/kg. A value below 10 €/kg is competitive considering with respect to the actual cost of the hydrogen in the market, especially considering that, owing to the small size, i.e. hydrogen distributed production, there is no distribution cost. In order to be competitive in the refueling station fuel market the cost have to be less than 5 €/kg. This can be obtained in large centralized plant [22] or via a more important capex and biomass reduction cost together with a more important efficiency increase (e.g. via CO_2 capture as indicated in the technical analysis) in this small size plants.

6. Conclusion

This analysis was conducted to study hydrogen production cost as a function of hydrogen production efficiency and portable purification unit (PPS) cost. The results showed that system efficiency increase cannot be able to reduce costs to favorable level alone. In other words, PPS cost recognized as the major cost is requisite to go down. Therefore, the 50% reduction of PPS cost and the variation of steam to biomass from 1 to 1.5 allow the cost to fluctuate between 12.75-9.5 €/kg. Different feedstock, technologies and configurations have to be further analysed respect the variation of efficiency, cost and plant reliability. E.g. hot gas cleaning technologies, CO_2 sorbent inside the reactors (using the gasifier and combustor reactors for sorption and desorption cycles), reformer reactor between gasifier and WGS, etc. they are more efficient and requires less space but are more expensive and less reliable than others as cold gas cleaning, CO_2 removal via presurized water scrubbing and reformer in the PSA off gas line, etc. Thus, they 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 CO_2 sorbent and SMR avoiding that excessive S/B or other thermal needs leads to increase in energy demand which is followed by decrease in thermal efficiency.

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