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Techno-economic Assessment of the Fermentative Hydrogen Production from Sugar Beet

Krzysztof Urbaniec*, Robert Grabarczyk

Warsaw University of Technology, Plock Campus, CERED Centre of Excellence, Jachowicza 2/4, 09-402 Plock, Poland
gstku@mbox.pw.edu.pl

Two-stage fermentative hydrogen production process comprising dark fermentation and photofermentation followed by gas upgrading is studied from the technical and economic points of view. It is assumed that the H_2 plant is connected with an existing sugar factory so that technical sucrose solutions produced there (raw juice, thick juice and molasses) can be used as fermentation feedstocks.
It is shown that the total cost of produced H_2 is highly sensitive to the capital and operating costs of the photofermentation stage. Drawing conclusions from the results of a parametric study, the cost structure is reviewed to identify the directions of improvement in the process and equipment.

1. Introduction

At present, most of hydrogen produced worldwide comes from thermo-chemical conversion of fossil fuels. However, similar to bioethanol or biomethane, hydrogen can be sustainably produced by biological conversion of biomass. Two-stage bacterial fermentation, that is, dark fermentation followed by photofermentation is a promising conversion method which has been extensively studied in EU FP6 research project HYVOLUTION (Claassen et al., 2009; HYVOLUTION, 2011). On the basis of project findings, sugar beet is considered among the best raw materials for fermentative H_2 production (Panagiotopoulos et al., 2010).
Building on a generally positive operating experience of bioethanol plants integrated with existing sugar factories (Keil et al., 2009; British Sugar, 2010), the integration of hydrogen production with beet sugar production can be considered (Markowski et al., 2009). In the present paper, the issues of processing of technical sucrose solutions, that is, raw juice, thick juice and molasses, to H_2 in a plant connected with a beet sugar factory are reviewed, the production costs are estimated and their structure is assessed.

2. Description of the process

A block diagram of the two-stage fermentative hydrogen production from technical sucrose solutions delivered from a sugar factory is shown in Figure 1 (Grabarczyk et al., 2011). The feedstock is first diluted with water to the required concentration of substrate and supplied to the thermophilic fermentation where sucrose is converted to H_2, carbon dioxide and acetic acid. To keep a high H_2 yield, a temperature of about 70 ºC and pressure below 50 kPa are maintained in the bioreactor. The liquid effluent of the thermophilic fermentation is cooled down to 35 ºC, diluted to reduce the concentration of

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acetic acid and sent to the photofermentation. In the photobioreactor acetic acid is reduced by photo-heterotrophic bacteria, under influence of sunlight, to H_2 and CO_2.

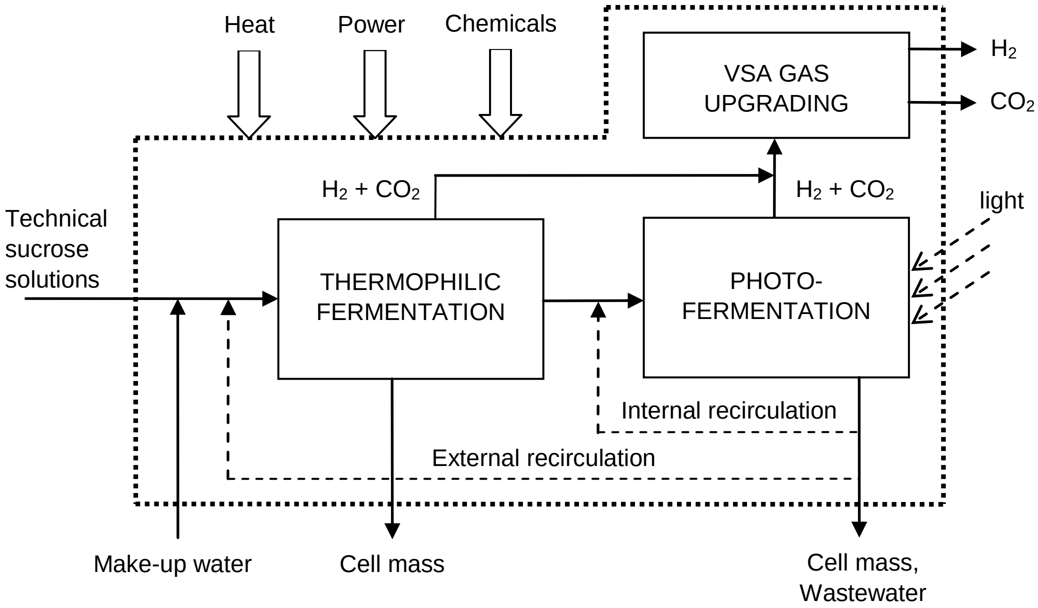

Figure 1: Block diagram of the two-stage fermentative processing of sucrose to hydrogen

As the photofermentation is not operated at night-time, during the rest of 24-h cycle its throughput has to be harmonized with the throughput of continuously operated thermophilic fermentation. When both stages are in operation, a part of the effluent from the photofermentation is recirculated to dilute the fermentation broths (dashed lines in Figure 1). Another part of that effluent stream is collected in a storage tank and can be used for external recirculation during the dark hours; alternatively, the incoming sucrose solution can at night-time be diluted with make-up water (Foglia et al., 2010). Obviously, the effluent from the first stage is collected at night-time in another storage tank for later processing.

In order to reduce the process heat demand, the thermophilic fermentation and photofermentation are thermally integrated (Markowski et al., 2009; Foglia et al., 2010). In a heat exchanger the effluent from thermophilic fermentation is cooled down by the external recirculation stream or by make-up water needed to dilute the sucrose solution.

The gas mixture obtained from bioreactors is supplied to a vacuum swing adsorption unit where separation of CO_2 on molecular sieve takes place.

The optimum pH values of the fermentation broths in thermophilic fermentation and photofermentation are 6.5 and 7.3, respectively and therefore additional chemicals like potassium hydroxide and phosphates are required to set the pH on suitable level.

3. Modelling of the process

A mathematical model was developed and implemented in Microsoft Excel to simulate mass and energy balances of the process thus enabling the calculation of the costs of hydrogen production. The simulation results were used to size process equipment pieces and to estimate their purchase prices.

For the thermophilic bioreactor, pumps, compressors, shell-and-tube heat exchangers, pressure vessels and fan cooler the purchase prices were estimated by:

$$C = C_B \left(\frac{Q}{Q_B} \right)^M \tag{1}$$

where C - purchase price of the equipment piece with nominal capacity Q, C_B - known base purchase price of the equipment piece with nominal capacity Q_B, M - constant depending on equipment type.

For the plate heat exchangers (C_{PHE}), the purchase price was expressed as a function of a heat transfer area (A):

$$C_{PHE} = A \cdot D + E \tag{2}$$

where D, E - constants depending on the type of plate heat exchanger.

The values of base purchase prices and constants D and E were delivered by engineering companies whereas the values of constant M were adopted from Smith (2005).

The photobioreactor was regarded as a set of tubular modules made of plastics, each having an illuminated area of 380 m^2 and a liquid volume of 15 m^3; the purchase price of one module was estimated at 5200 €.

The fixed capital investment was calculated by multiplying the total purchase price of all equipment by Lang factor. The annual capital cost was determined as the fixed capital investment divided by the operating period. For the calculations 15-year operating period and the Lang factor of 3 were assumed. The annual maintenance cost of the plant excluding photofermentation was calculated by multiplying the annual capital cost by a factor of 0.1. Regarding photofermentation, the operating cost comes from the short one-year lifetime of the tubes made of low-density polyethylene that are used in the photobioreactor. This also influences the cost of labor; it was assumed that in addition to 3 skilled workers operating the hydrogen plant, 24 man-hours of unskilled workers are needed to replace the plastic tubes in one photofermentor module. Detailed cost data used in the calculations are collected in Table 1.

Table 1: Detailed costs used in the calculations

	Price	Unit
Molasses	186	€/t sucrose
Electricity	0.092	€/kWh
Steam, 2 bar	18.64	€/t
Cooling water	0.01	€/t
Potassium hydroxide	140	€/t
Phosphates	500	€/t
Molecular sieve 13X	3140	€/t
Plastic tube	0.11	€/m
Labor skilled/unskilled	15000/7300	€/(y·employee)
Land	4230	€/ha

4. Results and discussion

A parametric study of the costs of hydrogen production from molasses in a plant with gross H_2 output of 60 kg/h (equivalent energy flow 2000 kW) was carried out assuming that photofermentation is operated 10 h per day. Three different process cases named Base, Real and Optimistic were consideed (Table 2). In the Base case, the values of process parameters were assumed in accordance with the state of knowledge at the start of HYVOLUTION R&D project (Claassen et al., 2010). In the Real case, the process parameters were adjusted taking experimental results of HYVOLUTION research on two-stage hydrogen fermentation of molasses into account (HYVOLUTION, 2011). The

Optimistic case corresponds to process parameters that are likely to be attained in the future when the two-stage process has been further developed (Ahrer, 2011).
The results of the parametric study are presented in Table 2. In the Base and Real cases, the cost of H_2 production is very high exceeding 30 €/kg but in the Optimistic case, it is reduced by a factor of about 3.3.

Table 2: Results of the parametric study

Process case	Base	Real	Optimistic
Thermophilic fermentation			
Sucrose concentration [g/L]	10	10	50
H_2 productivity [mmol/(L·h)]	20	16.3	50
Conversion factor [%]	80	70	87
Photofermentation			
Acetic acid concentration [mmol/L]	40	40	100
H_2 productivity [mmol/(L·h)]	0.5	0.5	3
Conversion factor, %	60	45	75
Results			
H_2 production cost [€/kg]	31.88	31.92	9.30
Energy yield of the plant [-]	1.82	1.68	3,28
Molasses demand [kg/h]	652	725	598
Average heat demand [kW]	752	824	434
Average power demand [kW]	347	367	176
External recirculation/make-up water demand [kg/h]	64040	71240	10930
Internal recirculation [kg/h]	349370	475550	140160
Photobioreactor area [ha]	255.8	250.4	43.6

The cost structure is illustrated in Figure 2. In all the cases, the overall production cost is dominated by the capital costs and mainly those induced by a large size of the photobioreactor (see Table 2).

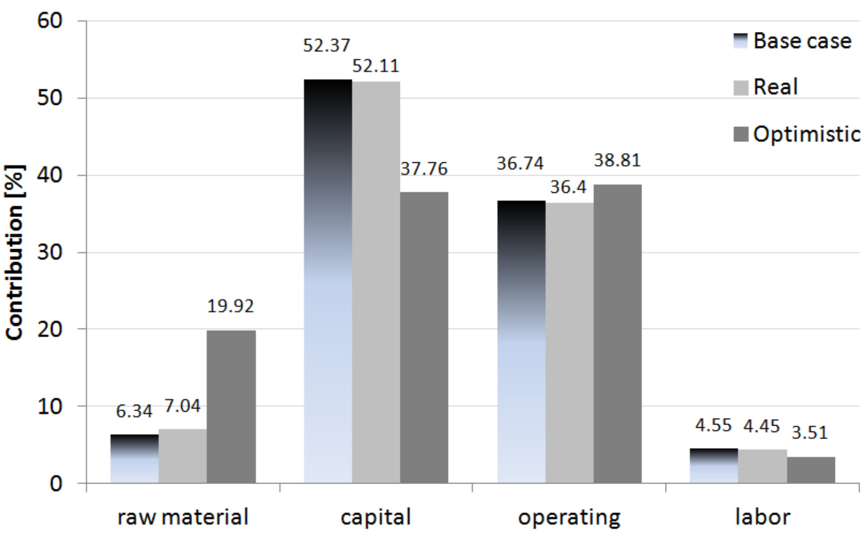

Figure 2: H_2 production cost structure

The capital cost of the photofermentation part of the plant is one order of magnitude higher than the capital costs of thermophilic fermentation and gas upgrading (Figure 3). The operating costs of photofermentation are also high owing to the short lifetime of plastic tubes and labor-intensive maintenance of the photobioreactor (Figure 4).

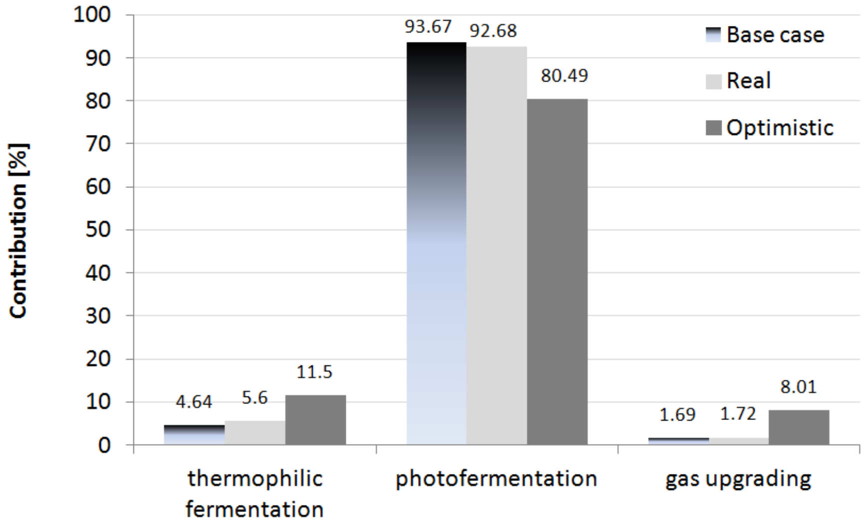

Figure 3: Capital cost structure

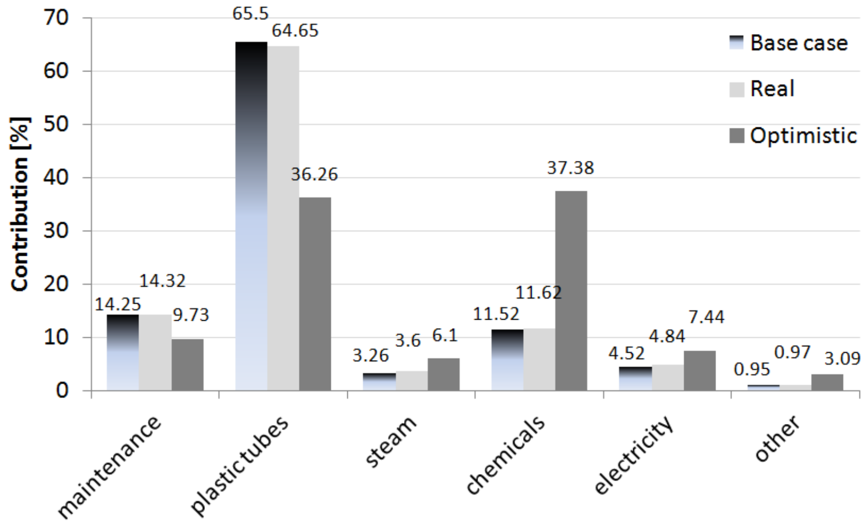

Figure 4: Operating cost structure

5. Concluding remarks

The techno-economic analysis of the two-stage fermentative process of H_2 production from sugar-beet molasses indicates that at the present state of knowledge and technology development, the process is not competitive. The main reasons for that are high capital and operation costs of the photofermentation part of the production plant. However, it is known that photobioreactor design can be substantially improved and other improvements, like a reduction in the demand for process chemicals, are also possible. Consequently, further research can be expected to improve the economic characteristics of fermentative H_2 production.

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