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이단계 바이오 수소/메탄 생산공정의 경제성 평가

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Economic Evaluation of Two-step Biohydrogen/biomethane Production Process

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ABSTRACT

본 연구에서는 이 단계 연속 바이오 수소/메탄 생산공정의 경제성을 조사하였다. 경제적 관점에서 다양한 수소 및 메탄 발효용 생물반응기를 비교·평가하였다. 이를 바탕으로 포도당으로부터 일 단계 수소발효를 위해 고온 trickling biofilter 반응기 (TBR, 100 m^3 규모)를, 일 단계 반응의 부산물로 생성된 유기산과 알콜류의 이 단계 메탄전환을 위해 고온 upflow anaerobic sludge 반응기 (UASB; 700 m^3 규모)를 선정하였다. 본 이 단계 공정의 수소생산 비용은 \$ 0.26/Nm^3으로 계산되었고, 이는 고온 TBR 반응기만을 이용한 경우보다 약 30 % 낮았다. 이 단계 공정의 낮은 수소생산 비용은 높은 에너지 회수율과 낮은 슬러지 처리비용에 의한 것이었다. 생물학적 수소 생산공정의 경제성은 탄소원의 종류, 생물반응기의 형태 등 여러 인자에 의해 변경될 수 있으나, 본 연구결과는 향후 연구를 위한 유용한 기준으로 고려될 수 있다.

주요기술용어 : Hydrogen(수소), Biological hydrogen production(생물학적 수소생산), Economic feasibility(경제성), Two-stage fermentation(이단발효), Methane (메탄), Biomass(바이오매스)

1. Introduction

Hydrogen is an efficient energy carrier with a high energy content per unit mass. It is considered to be the cleanest energy carrier because the combustion by-product is only water. It does not produce any green-house gases and neither does it contribute to global warming[1]. Hydrogen is also an important raw material in various chemical industries, and is gaining increasing attention[2].

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Hydrogen can be produced by either chemical or biological methods. The economic feasibility of biological methods mainly depends on the rate and yield of the H_2 production from various substrates. Many researchers have focused on developing efficient processes[3-13]. For example, Kumar and Das[6] immobilized *Enterobacter cloacae* IIT-BT 08 onto lignocellulosic agroresidues and observed a high H_2 production rate of 1,814 mmol H_2/l · d at a short HRT of 1.08 h. Rachman *et al.*[12] reported that self-flocculating *E. aerogenes* mutant AY-2 exhibits a high H_2 production rate as well as a yield of 1,392 mmol H_2/l · d and 1.1 mol H_2/mol glucose at an HRT of 1.49 h. Recently, a thermophilic trickling biofilter reactor (TBR) was studied in detail for fermentative H_2 production[11]. The reactor exhibited a high H_2 production rate of 1,050 mmol H_2/l · d and, more importantly, could be operated stably for an extended period of 234 d.

Fermentative H_2 production from a biomass produces many organic by-products such as alcohols and organic acids[11,14-16]. These by-products can be utilized for producing more energy: either H_2 in a photo-bioreactor or CH_4 in a second anaerobic reactor. Hydrogen production using a photo-bioreactor is attractive since more H_2 can be obtained[17], but the problems are that the H_2 production rate is slow and that not all organic by-products are converted into H_2[18]. In addition, scale-up of the photo-bioreactor is technically difficult and expensive[19]. In comparison, large-scale CH_4 production has been widely practiced in industry for many years[20,21]. Methane fermentation can proceed to completion, whereas most organic compounds generated from dark H_2 fermentation are depleted. Therefore, a two-step biohydrogen/biomethane production process should be considered as one of the ideal energy production processes at present.

There have been a limited number of reports on the economic feasibility of fermentative H_2 production. Recently, de Vrije and Claassen[17] studied the economics of a relatively small-scale plant (425 Nm_3 H_2/h), using lignocellulosic substrate as a raw material. They estimated the production cost to be \$ 0.29/Nm^3 H_2 in a two-stage fermentation that consisted of an extremely thermophilic dark-fermentation as the first stage and a photo-fermentation as the second stage. Tanisho[22] reported on a comparison between fermentative H_2 production and ethanol fermentation, using sugar cane as a raw material. He estimated the H_2 production cost to be \$ 0.48/Nm^3 H_2 and reported that the fermentative H_2 production had a nearly equal feasibility to the fermentative ethanol production. However, none of these studies described the capital investment and plant operation in details.

In the present study, the economics of a two-step, fermentative H_2 and CH_4 production process was investigated. Various processes for H_2 and CH_4 production were reviewed and compared from an economic standpoint. A thermophilic TBR (100 m^3) for fermenting glucose for H_2 production under darkness was chosen as the first stage, and a thermophilic upflow anaerobic sludge blanket (UASB) reactor (700 m^3) for conversion of the first-stage by-products to CH_4 was selected as the second stage (Fig. 1). The capital and operating costs of the two-step process were estimated, and the unit cost of H_2 production (\$/Nm^3) was evaluated. Finally, the cost was compared with that of a single-step thermophilic TBR and with those of other H_2 production processes.

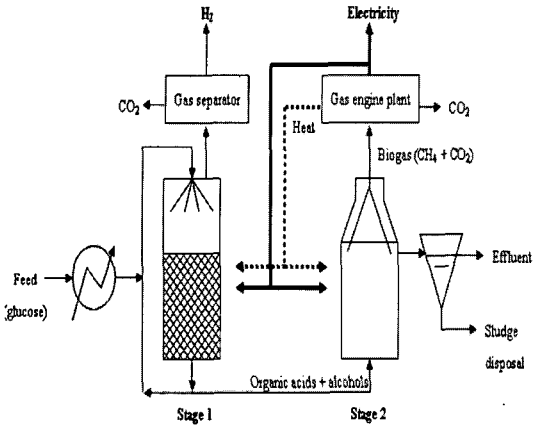

Fig. 1 Schematic diagram of a two-step biohydrogen/biomethane production process

2. Selection of bioreactors for two-step biohydrogen/biomethane production process

For the economical production of H_2 and CH_4, proper bioreactor systems should be selected. Bioreactors can be divided into suspended and immobilized systems. Suspended systems, typically represented by a continuously stirred tank reactor (CSTR), are simple in configuration and easy to operate. However, when operated at a high dilution rate or fed with a toxic substance, the system becomes unstable and often results in the washout of the cells. The immobilized system uses a support matrix onto which biofilms are established. This system yields a higher volumetric production rate than the suspended one. In addition, it is more stable in the event of an abrupt change in operating conditions such as pH, temperature, or organic load, since the cells in the biofilms are generally hardier to an external disturbance than planktonic free cells.

Table 1 summarizes selected H_2-production results from several continuous reactors. Immobilized systems yield a several-times higher H_2 production rate than suspended ones, although the yields for carbon sources are similar. Among the immobilized systems, the results by Rachman *et al.*[12], Chang *et al.*[3], and Kumar and Das[6] on packed bed reactors are interesting since the production rates are very high, in the range of 1,188 to 1,814 mmol H_2/l · d. However, in the cited studies, the operating period was short, less than 40 d, the reactor volume was small, less than 250 ml, and/or pure culture was used[6,12]. This information is not sufficient for evaluating the reactor performance for long-term operation and scale-up. In comparison, a thermophilic TBR by Oh *et al.*[11] and a UASB by Chang and Lin[4] have been stably operated for long periods of more than 230 d. The TBR also exhibited a high H_2 production rate (1,050 mmol H_2/l · d) and yield (1.1 mol H_2/mol glucose). Therefore, the thermophilic TBR was selected as the first stage for the purposes of further analysis in this study.

Bioreactors for CH_4 production are summarized in Table 2. The UASB reactor with a conventional or expanded bed generally has been accepted as the most effective, and is the most widely used, in treating high-strength wastewater[21,23,24]. This system employs sludge granules that contain spatially self-organized, microbial consortia of various acidogens and methanogens. For CH_4 production, the granules can utilize not only sugars such as glucose, fructose, sucrose, starch, among others, but also H_2 fermentation by-products such as ethanol, formate, acetate, propionate, succinate and other small organic acids[25]. The UASB process can be operated at either thermophilic (45 - 65°C) or mesophilic (30 - 40°C) temperatures. The thermophilic UASB has a much higher metabolic rate than the mesophilic, allowing a higher COD

Table 1 Comparison of some selected continuous bioreactors for fermentative H_2 production

System	Reactor type (Working vol., l)	Microorganism	Support matrix	Operating period (d)	HRT (h)	Carbon substrate	Volumetric H_2 production rate (mmol H_2/l · d)	H_2 yield (mol H_2/mol substrate)	Ref.
Immobilized	TBR (2)	Mixed culture	Synthetic polymer	234	2	Glucose	1,050	1.1	11
	Packed-bed (3)	Mixed culture	Activated carbon	13	1	Sucrose	1,188 [a]	1.1 [a]	3
	Packed-bed (0.25)	*Enterobacter cloacae* IIT-BT 08	Lignocellulosic agroresidues	> 40	1.08	Glucose	1,814	nr	6
	Packed-bed (0.09)	*E. aerogenes* mutant AY-2	None	18	1.49	Glucose	1,392	1.1	12
	Packed-bed (0.075)	*E. aerogenes* strain HO-39	Porous glass bead	8	1	Glucose	803 [a]	0.73	13
	UASB (3)	Mixed culture	None	230	8	Sucrose	271	1.5	4
Suspended	CSTR (4)	Mixed culture	None	100	13.3	Glucose	184 [a]	1.63	5
	CSTR (2.3)	Mixed culture	None	62	8.5	Glucose	119 [a] (189) [a,b]	0.85 (1.43) [b]	9
	Chemostat-type digester (4) [c]	Mixed culture	None	nr	6	Glucose	711	1.7	7
	Chemostat-type digester (1) [d]	Mixed culture	None	nr	2	Glucose	47 [a]	nr	10
	Anaerobic sequential batch reactor (5)	Mixed culture	None	nr	4	Sucrose	470	1.4	8

nr: Not reported.

[a] Estimated from experimental data of the reference.

[b] N_2 was continuously sparged at a flow rate of 110 ml/min.

[c] Completely mixed with produced biogas.

[d] Completely mixed by a liquid circulation pump.

Table 2 Comparison of some selected continuous bioreactors for fermentative CH_4 production

System	Reactor type	Microorganism	Requirement for support matrix	HRT (h)	Input COD (kg COD/m^3)	Organic loading rate (kg COD/m^3 · d)	COD removal (%)	Ref.
Immobilized	UASB (Mesophilic)	Mixed culture	no	4 - 12	5 - 15	4 - 12	75 - 85	29
	UASB (Thermophilic)	Mixed culture	no	1 - 12	3 - 30	10 - 100	75 - 95	24
	Fixed bed	Mixed culture	yes	24 - 48	10 - 20	1 - 5	75 - 85	29
	Expanded bed	Mixed culture	yes	5 - 10	5 - 10	5 - 10	80 - 85	29
Suspended	CSTR	Mixed culture	no	> 10 d		0.25 - 3		30
	Contact	Mixed culture	no	2 - 10	1.5 - 5	0.5 - 2.5	75 - 90	29

removal rate at a shorter wastewater retention time[26]. For example, Uemura and Harada[27] have reported that, in a lab-scale (5.5 l) thermophilic UASB, the COD (chemical oxygen demand) removal efficiency was 96% at an organic loading rate of 87 kg COD/m^3 · d, using a sucrose-acetate mixture as a substrate. Ohtsuki *et al.*[28] reported that, for beer brewing wastewater, which usually contains many recalcitrant compounds for biological degradation, a large-scale (1,400 l) two-stage UASB system showed a COD removal efficiency of 90 % at 30 kg COD/m^3 · d. Continuous operation of a thermophilic UASB reactor with the effluent of H_2 fermentation broth has yet to be reported, but the literature indicates that this system is the most appropriate for CH_4

production. As a consequence, the thermophilic UASB reactor was selected as the second stage in this study.

3. Cost estimation of two-step biohydrogen/biomethane production process

Performance-efficiency and operating- cost estimations of the selected two-step biohydrogen/ biomethane production process are shown in Tables 3 and 4. Glucose was selected as the model substrate in this study. Pure glucose is expensive and not usually used for energy production on a commercial scale. However, most of a biomass is converted into glucose before being utilized in many biological processes and some food wastewater such as those from coffee processing, palm oil mills or cannery distilleries[31] can be used for this purpose. In the case of using wastewater, the cost of pretreating the biomass varies greatly depending on the source of the wastewater and pre-treatment methods, and thus is difficult to estimate. In this study, the cost of glucose and pre-treatment was not taken into account. This is not unreasonable, however, since the cost of wastewater treatment or the gate fee could compensate for the cost of the pre-treatment of a carbonaceous biomass in many circumstances[32].

The basis for the estimation, considering a relatively small-scale system that is to be fed by a locally available biomass, was 100 Nm^3 H_2/h. The TBR was assumed to treat the wastewater at a rate of 1,200 m^3/d for a glucose concentration of 21 kg/m^3 [11]. The operating conditions, H_2 production yield, and glucose removal efficiency of the large-scale TBR were assumed to be the same as those of the bench-scale TBR. Hydrogen from the TBR was assumed to be purified to satisfy the specifications for fuel cell application using pressure swing adsorption. The capital costs of the thermo-bioreactor and H_2 recovery/ purification were estimated based on the analyses by de Vrije and Claassen of a two-step process of hyper-thermophilic dark/photo biohydrogen production[17]. The energy requirement necessary for operating the thermo-bioreactor and H_2 recovery/purification system was also estimated according to de Vrije and Claassen[17], even though it could be fully supplied from the second-stage CH_4 in the present two-stage process. Most H_2 fermentation processes generate microbial sludges, the disposal of which incurs cost. In the present process, however, the biomass generated in the first stage can be continuously treated in the subsequent CH_4 reactor, and thereby its disposal cost is nullified. This might also improve the performance of the UASB reactor by accelerating the acidogensis of carbon sources transferred from the first stage without treatment[33]. The cost for nutrients and other chemicals required for the H_2 fermentation was estimated according to Kaylen *et al.*[34] The labor cost was assumed, according to Luccio *et al.*[35], to be 20 % of the total operating cost.

Maintenance and annualized capital requirements were estimated based on the analyses of a two-stage indirect biophotolysis process by Beneman[36]. Insurance, taxes, and other indirect costs were included in the overall maintenance costs as 6 % of the capital cost. The annualized capital costs were assumed to be 19 % of the actual capital cost, which included actual capital charges (averaging returns on investments and interest payments on debt) and depreciation.

In the CH_4 plant, the CH_4 production yield was assumed as 0.35 Nm^3 CH_4/kg COD[37]. The COD

Table 3 Performance of two-step biohydrogen/biomethane production process

Item	Unit	Value	Basis
Feed			
· Flowrate	m^3/d	1.2×10^3	
· Glucose concentration	kg/m^3	21	
· Ambient temperature	°C	20	
Performance of first-stage TBR[a]			
· Glucose loading rate	$kg/m^3 \cdot d$	250	
· Hydraulic retention time	h	2.0	
· Reactor volume (V1)	m^3	100	
· Reactor temperature	°C	60	
· Glucose removal efficiency	%	70	
· Sludge production (annually)	kg VSS/yr	0[b]	
· H_2 production yield	mol H_2/mol glucose	1.1	
· H_2 production rate (daily)	Nm^3/d	2.4×10^3	
· H_2 production rate (hourly, Q_h)	Nm^3/h	100	
· H_2 production rate (annually, Q_a)	Nm^3/yr	8.8×10^5	
Performance of second-stage UASB			
· Flowrate	m^3/d	1.2×10^3	
· Inlet COD concentration	kg COD/m^3	18[c]	
· Inlet COD loading rate	kg COD/$m^3 \cdot$ d	31	
· Hydraulic retention time	h	14	
· Reactor volume (V2)	m^3	700	
· Reactor temperature	°C	55	
· COD removal efficiency	%	90	
· COD, converted	kg COD/d	1.9104	
· Sludge production (annually)	kg VSS/yr	1.0×10^6	0.1 kg VSS/kg COD[d]
· CH_4 production yield	Nm^3 CH_4/kg COD	0.35[e]	
· CH_4 production rate	Nm^3/d	6.8×10^3	
· CH_4 content	%	60	
· Biogas production rate	Nm^3/d	1.1×10^4	
· Biogas energy production (annually)	GJ/yr	1.0×10^5	1 Nm^3 biogas = 0.026 GJ (lower heating value)[f]
· Energy conversion efficiency of gas engine	%	35	35%, electricity generation; 50%, heat generation; 15%, loss[f]
· Electricity generated by gas engine (annually)	kWh/yr	1.0×10^7	GJ = 277.8 kWh
· Electricity generated by gas engine (hourly, AH)	kWh	1.2×10^3	

[a] Adopted from bench-scale experiments by Oh *et al.*[11]
[b] Under the assumption that the biomass, generated in the first stage, is continuously fed into the UASB reactor.
[c] Composed of biomass, residual glucose, and by-products from the first-stage TBR.
[d] From Oh *et al.*[40]
[e] From Speece[37].
[f] From de Mes *et al.*[20]

Table 4 Cost estimation for two-step biohydrogen/biomethane production process

Item	Unit	Value	Basis
Operating factor			
· Annual operating days	d	365	
Capital and operating costs of first-stage TBR			
Capital cost			
· Thermo-bioreactor with pumps	$	3.8×10^5	$\$1.5\times10^6$ (V1/400 m^3)[a]
· H_2 recovery/purification	$	1.6×10^5	$\$7.0\times10^5$ (Q_h/430 Nm^3 H_2/h)[a]
· 'Saran-Lock' packing	$	2.0×10^4	$\$200/m^3$ [b]
· Total capital cost	$	5.6×10^5	
Annual operating costs			
· Pre-treatment cost	$/yr	-[c]	
· Energy	$/yr	0[d]	
· Sludge disposal	$/yr	0[e]	
· Chemicals	$/yr	4.8×10^4 [f]	
· Labor	$/yr	4.7×10^4	20% of total operating cost (I)[g]
· Overall maintenance	$/yr	3.4×10^4	6% of total capital cost[h]
· Annualized capital costs	$/yr	1.1×10^5	19% of total capital cost[h]
· Total operating costs (I)	$/yr	2.4×10^5	
· Unit cost of H_2 production	$/$Nm^3$ H_2	0.27	I/Q_a
Capital and operating costs of second-stage UASB			
Capital cost			
· Bioreactor unit	$	2.3×10^5	$\$330/m^3$ [i]
· Gas engine plant	$	8.4×10^5	AH×\$700/kWh[j]
· Total capital cost	$	1.1×10^6	
Annual operation costs			
· Energy	$/yr	0[d]	
· Chemicals	$/yr	4.5×10^4 [k]	
· Labor	$/yr	8.8×10^4	20% of total operating cost (III)[g]
· Sludge disposal	$/yr	4.0×10^4	\$40/1,000 kg TSS[l]
· Overall maintenance	$/yr	6.4×10^4	6% of total capital cost[h]
· Annualized capital costs	$/yr	2.0×10^5	19% of total capital cost[h]
· Income from electricity sales (II)	$/yr	-4.5×10^5 [m]	-, Profit
· Total operating costs (III)	$/yr	4.4×10^5	
· Total operating costs (IV)[n]	$/yr	-8.6×10^3	-, Profit; III - II
Operating costs of two-stage TBR/UASB system			
· Total operating costs (V)	$/yr	2.3×10^5	I + IV
· Unit cost of H_2 production	$/$Nm^3$ H_2	0.26	V/Q_a

Nomenclature: V1, volume of first-stage TBR; Q_h, H_2 production rate (hourly); and, AH, electricity (hourly) generated by gas engine.
[a] From de Vrije and Claassen[17] based on 1 EURO = \$1.16.
[b] Provided by Dokil Felt Industry, Ltd., Seoul, South Korea.
[c] Assumed zero. See the text.
[d] Supplied by energy (electricity and heat) from gas engine of second-stage CH_4 plant.
[e] Assumed that the biomass, generated in the first-stage, is continuously fed to the UASB bioreactor without treatment.
[f] Estimated based on the unit cost and amounts of nutrients and other chemicals required for production of ethanol from lignocellulosic

feedstocks (Kaylen *et al.*[34]).

[g] From Luccio *et al.*[35]

[h] From Benemann[36].

[i] From van Haandel and Lettinga[25].

[j] From 2002 brochure, Jenbacher AG, Jenbach, Austria (http://www.jenbacher.com).

[k] For pH neutralization only; based on the costanalysis for pH neutralization of anaerobic digester by Trout *et al.*[39]

[l] From Yoon *et al.*[41]

[m] (Electricity for sales) = (annual electricity generated by gas engine) (electricity consumption for the first stage and the second stage). It was assumed that the methanogenic plant consumes 31% of the electricity generated by gas engine (de Mes *et al.*[20]).

[n] Income from electricity sales was considered.

loading rate of 31 kg COD/m^3 · d was assumed to be treated by the second-stage UASB[27,28] because most CODs transferred from the first stage were easily biodegradable. Biogas produced in the UASB can be directly fed to a gas engine and used to generate electricity without further purification. It was assumed that the electricity produced would be partially used for operating the two-stage TBR/UASB plant, and the remainder was supplied to a public grid for sales without loss. The electricity charges for commercial and industrial uses are \$ 0.070/kWh and \$ 0.045/kWh, respectively[38]. In this study, the commercial charge was assumed for sales. The construction cost of a UASB reactor is dependent on volume, depth, type of separator, construction materials, and other factors. The capital cost for the UASB was estimated according to van Haandel and Lettinga[25]. The cost for the gas engine plant was estimated according to Jenbacher AG, Jenbach, Austria (http://www.jenbacher.com). It was assumed that no chemical except for pH neutralization was needed for CH_4 fermentation. The chemical cost (e.g. lime) was estimated based on the cost for pH neutralization of an anaerobic digester, as suggested by Trout *et al*[39]. Other operating costs were estimated similarly to the case of the first-stage H_2 production plant.

The data summarized in Tables 3 and 4 show that the cost by the two-step process was estimated to be \$ 0.26/Nm^3 H_2, which is equivalent to \$ 20/GJ (based on the upper combustion value). If biohydrogen were produced by a single-step thermophilic TBR process without the subsequent CH_4 fermentation, the cost would increase to \$ 0.37/Nm^3 H_2 (data not shown). Without the second-stage CH_4 reactor, savings in capital and operational costs are significant. However, less energy is recovered, and additional costs for treating the sludge and COD from the first-stage are incurred. The difference of \$ 0.11/Nm^3 H_2 indicates that the presence of the second stage is advantageous.

4. Comparison of various H_2 production process

In Table 5, some data from the literature are summarized and compared with the present data according to H_2 production cost and net CO_2 emission. It was not easy to compare the current result with the others, since the economics are dependent on not only many direct factors (H_2 conversion efficiency, the H_2 production scale, the configuration and long- term stability of a system, and the utility of related facilities) but also indirect factors (labor cost, annualization of capital costs, overhead, licenses, civil works, etc.) that are site- and project-specific. Nonetheless, we noticed that the H_2 production cost by the present two-step or single-step process is comparable to that reported for most other H_2 production technologies except for the biophotolysis of water. In the latter case, the conversion efficiency of sunlight was assumed to be 10 %. This is far

Table 5 Comparison of H_2 production costs and net CO_2 emissions in small-scale production plants with capacity in the range of 20 - 1,500 Nm^3 H_2/h[a]

Production	Assumed conversion efficiency	Production cost ($/$Nm^3$ H_2)	Net CO_2-emission (kg/Nm^3 H_2)	Ref.
Two-step biohydrogen/biomethane production process from glucose	1.1 mol H_2/mol glucose	0.26	0	Present study
Single-step biohydrogen production process from glucose	1.1 mol H_2/mol glucose	0.37	0	Present study
Two-step dark/photo biohydrogen production process from biomass[b]	3.2 mol H_2/mol glucose	0.29[f]	0	17
Single-step biohydrogen production process from sugar cane		0.48[d]	0	22
Biophotolysis of water	10% [e]	0.13[f]	0	36
Steam-reforming of natural gas		0.37[f]	0.8	17
		0.35-0.38	nr	38
Partial oxidation of oil		0.42	nr	43
Steam-reforming of biomethane		0.37[f]	0	17
Gasification of biomass		0.17	0	44
Pyrolysis of biomass		0.17	0	44
Electrolysis with conventional electricity		0.27[d]	1.8	17
Electrolysis with CO_2-lean electricity		0.31-0.42[e,g]	0	17
Electrolysis with electricity from wind turbines		0.29[f]	0	17
Electrolysis with electricity from photovoltaic cells		3.42[f]	0	17

nr: not reported.

[a] For large-scale H2 production, >40,000 Nm^3 H_2/h for biophotolysis of water and >30,000 Nm^3 H_2/h for gasification of biomass.

[b] Based on zero substrate cost (acquisition of biomass and hydrolysis) and without personnel costs.

[c] Assumed 1 EURO = $1.16.

[d] Assumed $1 = 108¥ and 1 Nm^3 H_2 = 12.77×10^{-3} GJ (high heating value).

[e] Assumed that solar conversion efficiency of CO_2 fixation into storage products (glycogen or starch) is 10% and that the storage products are stoichiometrically converted to H_2 gas (e.g. 12 H_2/mole of glucose).

[f] Assumed 1 Nm^3 H_2 = 12.77×10^{-3} GJ (high heating value).

[g] CO_2 is sequestered, rendering the process CO_2-neutral.

more optimistic, considering that the current value in most experiments falls below 1 %[42]. For the small-scale H_2 production of 100 Nm^3 H_2/h, the thermophilic TBR process combined with the auxiliary UASB reactor can be considered to be a potentially economical process in its present state.

The emission of CO_2 in H_2 production processes is an important concern, since it is responsible for global warming[2]. As distinguished from the physicochemical methods in which natural gas or conventional electricity is used, biological methods yield no net CO_2 emission. Biological H_2 production also proceeds along with the production of CO_2; however, this CO_2 has been previously immobilized from atmospheric CO_2 into the biomass[17]. As a consequence, the biological process is considered to be CO_2-neutral.

The present analyses are based on the many data from the literature as well as our own assumptions. For more reliable analyses, extensive experiments on the proposed two-step biohydrogen/biomethane production process are required. First of all, a scale-up of the thermophilic TBR process should be conducted. Second, many assumptions on the performance of the combined H_2 and CH_4 production process should be verified. Especially, the rate and yield of biogas production along with the operational stability of the two-stage system should be confirmed. Third, depending on local legislation concerning standards of effluent quality, post-treatment might be required for removal of organics and other nutrients including nitrogen and phosphorus. In this case, the cost for such post-treatment facilities should be included in the economic analysis of the two-step process.

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