Boundary Analysis for H2 Production by Fermentation

Submitted To:

National Renewable Energy Laboratory

by

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March 12, 2004

Margaret Mann
National Renewable Energy Laboratory
1617 Cole Boulevard, MS 1613
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Dear Maggie:

Enclosed is a conceptual design and order-of-magnitude economic analysis for the production of hydrogen by fermentation of carbohydrates under the following design basis:

Production Rate:	120,480 kg H_2/day
Plant Type:	Battery limits with no capital required for OSBL expansions
Utilities:	Purchased from host facility

Fermentation Yield, moles H_2 per mole Glucose	10
Cost of glucose, \$/lb (dry)	0.05
Nitrogen and other nutrient sources	None
Waste treatment costs	None

The results of the economic analysis are:

Fixed Capital, \$MM (2000)	73.6
Required H_2 Price, \$/kg H2	2.08

where the required H_2 price is defined as the price required for a 10% after-tax real internal rate of return under the assumption of 100% equity financing and 40 years of plant operation.

The results projected by this study should be considered as a “best case” scenario. The assumed technical performance could only be achieved after significant long-term research.

Sensitivity analysis shows major economics drivers include: fermentation yield, cost of glucose, fixed capital, glucose concentration in the fermentation and various financial assumptions.

Sincerely,

Tim Eggeman, Ph.D., P.E.

Introduction

The Hydrogen Program of the U.S. Department of Energy is examining novel methods for hydrogen production. One potential means for hydrogen production is the fermentation of glucose or other carbohydrates:

$$C_6H_{12}O_6 + 6\ H_2O \Rightarrow 12\ H_2 + 6\ CO_2, \qquad \Delta G^o = -16\ \text{kJ/mol}$$
(Glucose)

Theoretically, up to 12 moles of hydrogen can be produced per mole of glucose consumed.

This report presents a boundary analysis for this approach. A simplified overall material and energy balance along with order-of-magnitude economics are presented.

Design Summary

The design basis assumed a hydrogen production rate of 50 MMSCFD at 435 psig. Figure 1 is a block flow diagram. The simplified material balance and utilities summary were computed with the assistance of Aspen Plus 10.2 (files: E0312b).

The base case assumes a yield of 10 moles of hydrogen per mole of glucose, with a glucose concentration of 3 wt% in feed to fermentation. The remaining fraction of the glucose is assumed to either be unconverted or diverted into cell mass production. No cost or value was associated with disposal of the cell mass stream. Also, no costs for nitrogen and other nutrient sources are included in the analysis.

A battery limits addition to an existing biorefinery is envisioned. The host facility will provide glucose, water, utilities and other infrastructure as a variable operating cost for the new facility. The fixed capital investment is limited to the battery limits unit; in other words, no capital for upgrades to the host facility's infrastructure is included in the base case.

Conceptually any carbohydrate source could be used as a feedstock for the fermentation. Corn starch hydrolyzate, juices from sugar cane or sugar beet processing, wood fiber hydrolyzates from pulp mills, or biomass hydrolyzates from agricultural resides, energy crops and other biomass sources are among the many possible sources. Many of these potential feedstocks would actually produce a mixture of sugars upon hydrolysis. To simplify the analysis, I assumed that carbohydrate was present only in the form of glucose. The presence of a mixed carbohydrate feedstock and its impact on the design were not considered in this preliminary analysis.

A few statements are needed to put the scale of this plant into perspective. By today's standards, a 50 MMSCFD hydrogen plant would be a large, but not world-scale, industrial hydrogen production facility. The amount of carbohydrate required to feed this facility roughly corresponds to the total carbohydrate present in 2,000 metric tons per day of corn stover. This feed rate corresponds to the rate used in the 2002 NREL Design Case for bioethanol production (1). The NREL design case assumes corn stover is collected from a 50 mile radius around the plant and is projected to produce 69.3 MMgal of ethanol per year. In terms of starch hydrolyzate from corn, the projected feed rate corresponds to a grind rate of roughly 80,000 bushels per day. This would be considered a medium scale corn dry milling facility or a small scale corn wet milling facility.

Results

Table 1 summarizes the projected fixed capital investment. Individual cost items were obtained as follows:

1) Fermentation Unit – Taken to be the same as for the fermentation unit in the 2002 NREL Design (1) (i.e. costing by analogy with a similar process)

2) Raw Gas Compressor - Estimated using Questimate, a commercially available software program for generating equipment costs.

3) PSA – Estimated on H_2 production rate and internal cost curves.

4) Cell Recovery Unit – A detailed definition of the unit is not available; a reasonable capital allocation was assumed.

5) Other – The balance of plant was assumed to be 50% of the identified equipment.

The total fixed capital is projected at $73.6MM, or a fixed capital investment of $1.47MM per MMSCFD of H_2 production. For comparison, the standard industry “rule-of-thumb” for fixed capital costs for production of hydrogen by steam methane reforming are $1.00MM per MMSCFD of H_2 in this capacity range, although recent facilities have had slightly lower requirements (2).

Table 2 summarizes projected plant level operating costs. Glucose is the major operating cost driver. The base case assumed that glucose was available at $0.05 per lb (dry). This is approximately the value placed on dilute, unfinished sugar streams in today’s corn processing facilities. The US DOE goal for the cost of sugar in biomass hydrolyzates is $0.07 per lb in Year 2012. Long term goals for US DOE are to reduce cost of biomass derived sugars into the $0.03-$0.05 per lb range.

Steam, electricity, and depreciation are other important operating cost drivers. Steam costs are driven by the assumed need to heat sterilize the fermentation media. Steam usage is sensitive to the glucose concentration in the media. Electricity costs are driven by the power needs of the raw gas compressor. The simulation assumed the inlet suction pressure of the raw gas compressor was 14.7 psia. Pressurized operation of the fermentation vessels may reduce the compression work requirements and compression capital, at the expense of increased capital for fermentation.

Table 3 summarizes projected revenues. Nearly 95% of the revenues are derived from hydrogen. The byproduct fuel gas, derived from the PSA tail gas, was valued on a heating value basis using industrial natural gas as the proxy.

The discounted cash flow calculations use the rational pricing rather market pricing method. For rational pricing, the hydrogen sales price required for a 10% after tax real internal rate of return for the project under the assumptions of 100% equity financing and 40 years of operation is calculated. The calculated required price does not reflect the market value of the hydrogen product.

Figure 2 displays the projected cumulative cash flows. Breakeven occurs between Year 8 and Year 9 of operation. Faster times to breakeven can be had by either increasing the required discount rate (i.e. making the required hydrogen selling price higher) or by increasing the amount of debt financing.

Figure 3 displays contributions to the calculated required price for hydrogen. Glucose feedstock, utilities, and capital are the major drivers.

Sensitivity Analysis

Figure 4 shows the effect of hydrogen yield and the cost of glucose on the feedstock cost component. The base case assumed 10 moles H_2 per mole glucose and \$0.05 per lb (dry) for glucose. The effect on the required hydrogen prices of changing these assumptions can be estimated using Figure 4 to adjust for the feedstock portion of the required hydrogen price. Note that other cost components such as capital and non-feedstock operating costs are ***not*** included in this figure. One opportunity for reduced hydrogen price is finding waste feedstocks or taking advantage of R&D advancements in the biomass feedstock preparation area. If the feedstock could be obtained for \$0.03/lb of glucose, the required hydrogen price would be reduced to \$1.60/kg. At a zero feedstock price (representing a niche-market waste), the required hydrogen price is reduced to \$0.90/lb.

Figure 5 shows the sensitivity of the required hydrogen price with respect to variations in fixed capital. Since the economics are dominated by the cost of carbohydrate, the effect of variations in fixed capital are somewhat muted.

Glucose concentration in the fermentation feedstock will affect both the amount of steam required for heat sterilization and the capital cost for the fermentation unit. Calculating the exact projections for the effect of glucose concentration on the required hydrogen price is involved. A rough estimate can be generated by just looking at the impact on steam usage. Lowering the glucose concentration in the feed from 3 wt% to 1.5 wt% doubles the amount of water present in the feed, which in turn doubles the heat load for sterilization, or adds another ~\$0.23 per kg H_2 to the required hydrogen price.

Figure 6 shows the sensitivity of the required hydrogen price with respect to variations in the required internal rate of return. The figure shows that the required internal rate of return has a significant impact on the projected economics.

Conclusions

The cost of carbohydrate and the yield of hydrogen from carbohydrate appear to drive the economics in the "best case" scenario considered in this report. It is questionable whether deployment of resources needed to develop this method of hydrogen production is justified, given the fact that the long term goals of the Hydrogen Program require costs lower than those projected by this analysis. The use of lower-priced feedstocks could reduce the required selling price of hydrogen to as low as $0.90 – 1.60/kg hydrogen. An assessment of the availability of these feedstocks and the real potential for R&D to produce low-cost sugars is needed.

References

(1) Aden, A., Ruth, M., Ibsen, K., Jechura, J., Neeves, K., Sheehan, J., Wallace, B, Montague, L., Slayton, A., Lukas, J., "Lignocellulosic Biomass to Ethanol Process Design and Economics Utilizing Co-Current Dilute Acid Prehydrolysis and Enzymatic Hydrolysis for Corn Stover", NREL/TP-510-32438, June 2002.

(2) Fleshman, J.D. in Chapter 6.2 of Meyers, R.A. (ed.), <u>Handbook of Petroleum Refining Processes</u>, 2nd ed., McGraw-Hill, 1997.

Table 1 - Fixed Capital Summary

		$MM (2000)
Direct		
	Fermentation Unit	10.0
	Raw Gas Compressor	11.3
	PSA	10.9
	Cell Mass Recovery Unit	4.0
	Other	18.1
	Total	54.4
Indirects		
	Engineering & Design	4.4
	Construction Expense	5.4
	Contractor's Fee	2.7
	Total	12.5
Contingency		
	Depreciable Fixed Capital w/o Contingency	66.9
	Contingency	6.7
	Depreciable Fixed Capital	73.6
	Non-Depreciable Fixed Capital	0.0
	Total Fixed Capital	73.6
	Fixed Capital, $MM/MMSCFD of H2	1.47

Table 2 - Operating Cost Summary

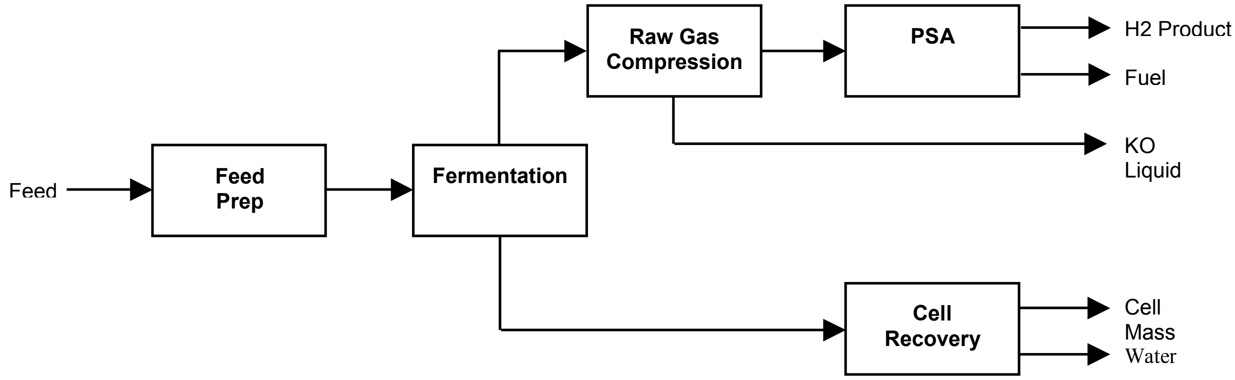
(Plant Level)

			$/yr	$/kg H2
Variable				
	Raw Materials			
		Glucose	40,789,314	1.1594
		Subtotal	40,789,314	1.1594
	Utilities			
		Steam	7,993,798	0.2272
		Electricity	7,507,863	0.2134
		Other	799,380	0.0227
		Subtotal	16,301,041	0.4634
	Other		735,916	0.0209
		Subtotal	735,916	0.0209
		Total Variable	57,826,271	1.6437
Fixed				
	Labor		1,248,000	0.0355
	Maintenace Labor		919,894	0.0261
	Maintenance Materials		1,839,789	0.0523
	Property Taxes & Insurance		1,471,831	0.0418
	General and Adminstrative		249,600	0.0071
	Depreciation (10 yr SL)		7,358,155	0.2092
		Total Fixed	13,087,269	0.3720
		Cash Costs	63,555,385	1.8066
		Operating Costs	70,913,541	2.0157

Table 3 - Revenue Summary

Item		$MM (2000)/yr	$/kg H2
Hydrogen		73.08	2.0772
Fuel Gas		4.12	0.1171
	Total	77.20	2.1944

Figure 1 – Design Summary



Stream Summary

American Standard Units

Stream	Feed	H2 Product	Fuel Gas	KO Liquid	Cell Mass	Water
Temperature, F	70	110	110	110.6	110	110
Pressure, psia	14.7	449.7	20.0	14.7	14.7	14.7
Components, lb/hr						
H2	0	11,067	1,953	0	0	0
CO2	0	0	142,080	0	47	0
Water	3,763,850	0	492	13,547	77,760	3,613,880
Glucose	116,408	0	0	0	19,440	0
Total, lb/hr	3,880,258	11,067	144,525	13,547	97,247	3,613,880

SI Units

Stream	Feed	H2 Product	Fuel Gas	KO Liquid	Cell Mass	Water
Temperature, C	21.1	43.3	43.3	43.7	43.3	43.3
Pressure, kPa	101.4	3100.5	137.9	101.4	101.4	101.4
Components, kg/hr						
H2	0	5,020	886	0	0	0
CO2	0	0	64,446	0	21	0
Water	1,707,256	0	223	6,145	35,272	1,639,230
Glucose	52,802	0	0	0	8,818	0
Total, kg/hr	1,760,057	5,020	65,555	6,145	44,111	1,639,230

Utilities Summary

Steam		
Exchanger	Duty, MMBtu/hr	Steam, lb/hr
H-101	185.9	185,899.6
	Total	185,899.6

Electricity		
Item	Power, hp (Shaft)	Power, kW (Motor)
C-400	24,677.1	20,446.4
P-100	184.1	152.6
Other		1,000.0
	Total	21,598.9

Figure 2 - Cumulative Cash Flow

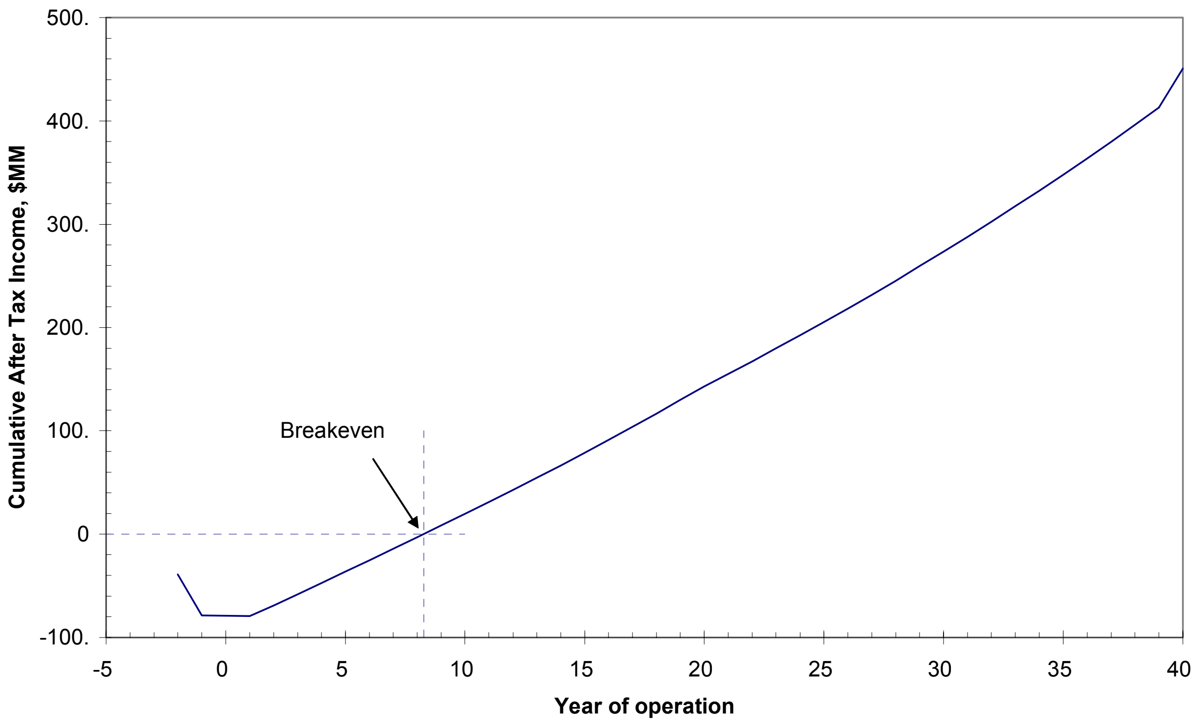


Figure 3 - Category Cost Contributions

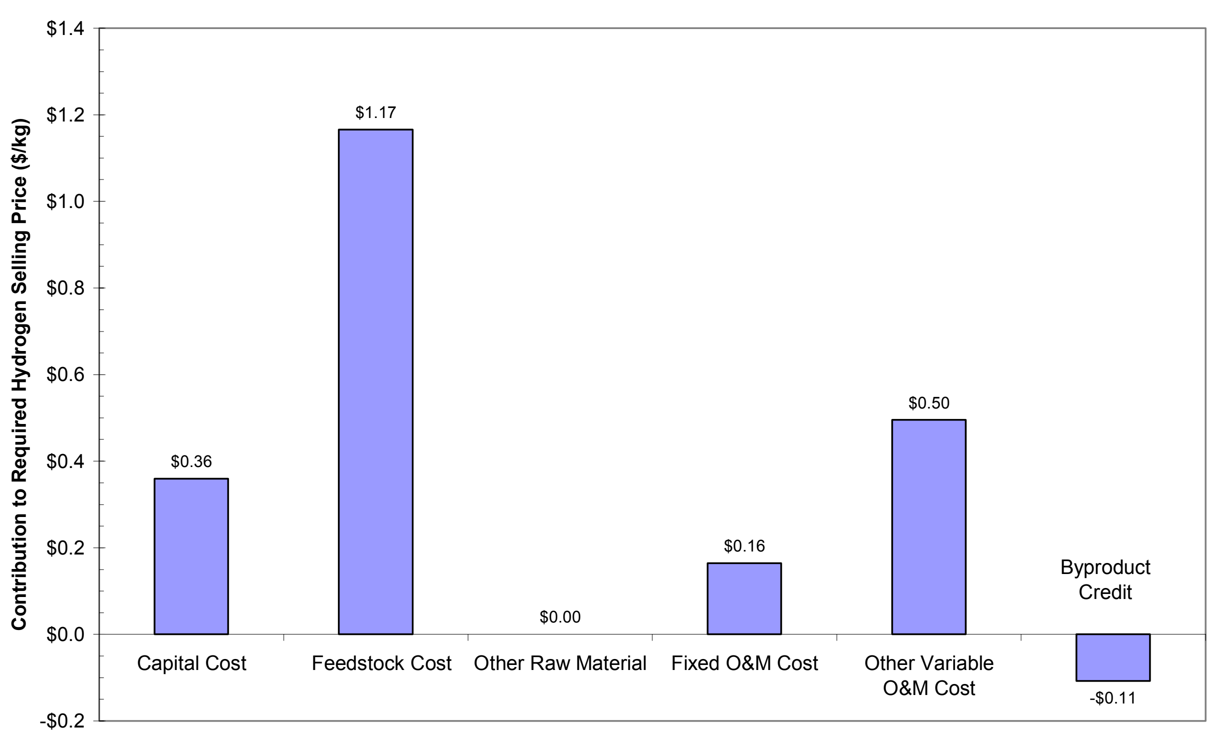


Figure 4 - Sensitivity of Feedstock Cost Component vs. Fermentation Yield and Glucose Cost

(Cost includes affect of H_2 loss in PSA; No capital or non-feedstock operating costs are included)

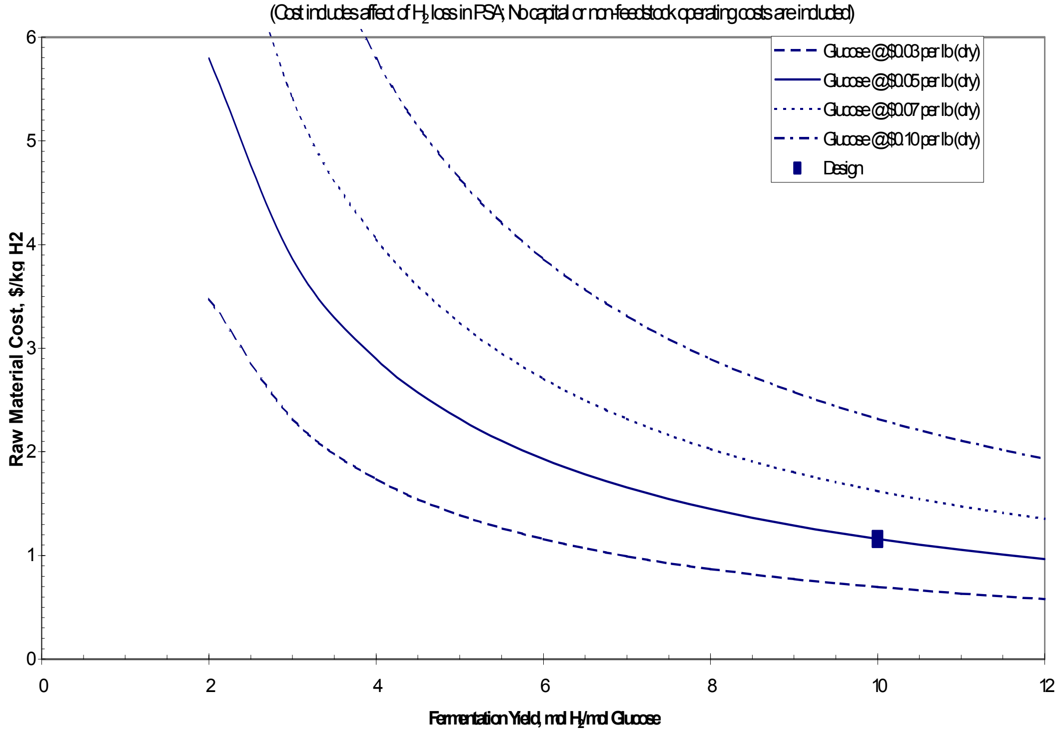


Figure 5 - Sensitivity wrt Fixed Capital

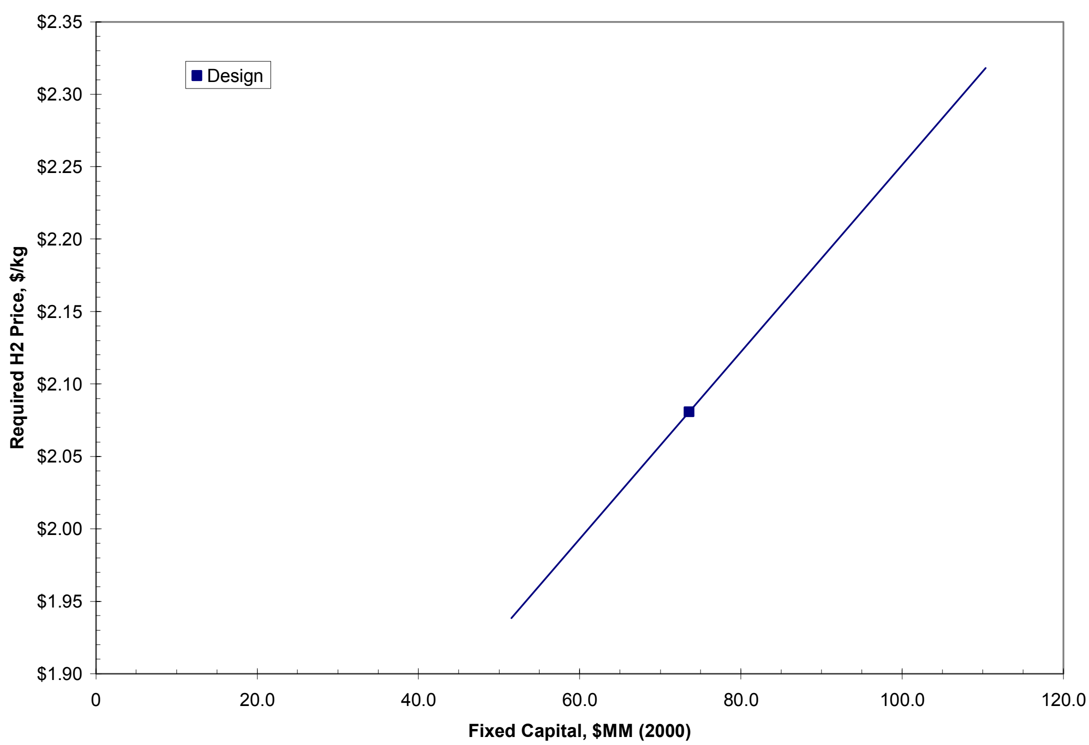
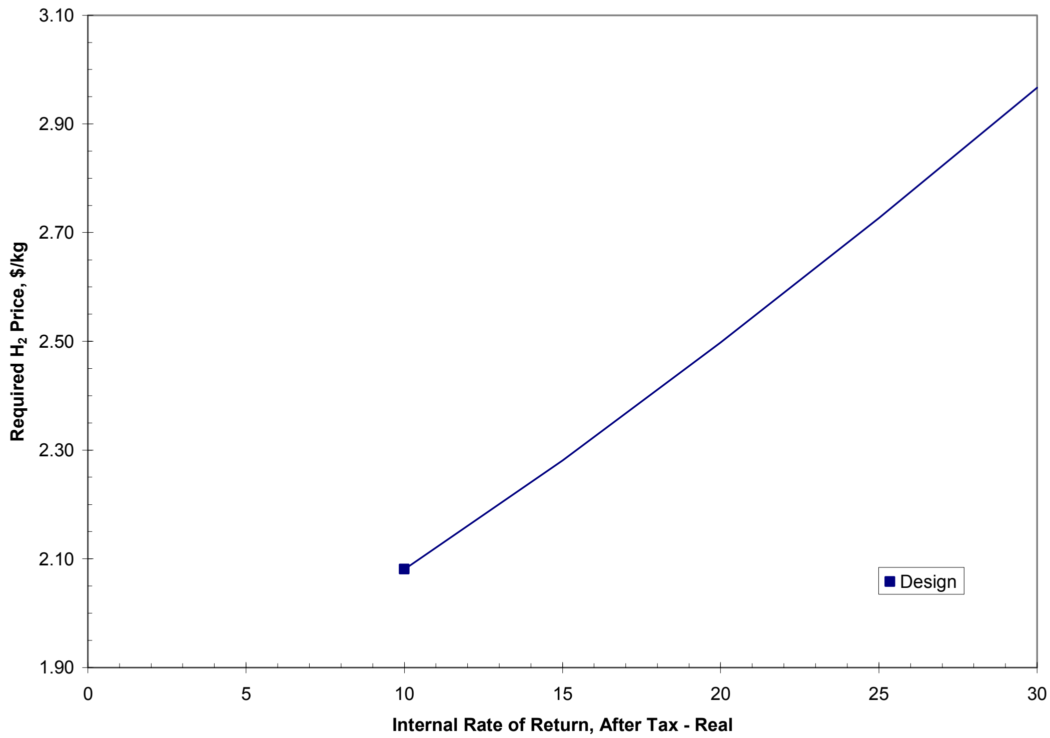


Figure 6 - Effect of IRR on Required H_2 Price



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Addendum
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July 29, 2004

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Dear Maggie:

This letter is a follow up on several issues raised at the US DOE Workshop on Hydrogen Production via Direct Fermentation held on June 9, 2004, in Baltimore, MD.

1) Operating Capacity Factor: The original analysis assumed an 80% operating capacity factor. I re-ran the economics using a 95% operating capacity factor. The required H_2 selling price drops from $2.077 per kg to $1.955 per kg. This change only makes a small change in the economics, thus the main conclusions of the report are unaffected.

2) Fermentation Assumptions: The report is an order-of-magnitude estimate for the economics of hydrogen production assuming significant improvements in the technology. I assumed that the fermentation was anaerobic, the feed sugar concentration was 30 g/l and the capital cost would be similar to that projected for the fermentation section of bioethanol plants processing a similar amount of feedstock. The current NREL design model for bioethanol production uses a simultaneous saccharification and fermentation design. The rate controlling step is hydrolysis of the complex carbohydrates into simple fermentable sugars, so volumetric productivity is independent of fermentation kinetics. It is too early to speculate on the relative rates of hydrogen production by fermentation vs. saccharification kinetics, but in essence I've assumed the fermentation kinetics will be faster than saccharification.

3) Sanitation Requirements: I envision the sanitation needs for this plant to be similar to those of today's ethanol production facilities. The feeds would be heat sterilized, but the plant would not require extreme levels of sterility. The assumed sugar feed concentration is fairly low for an industrial fermentation. Higher sugar concentrations would reduce steam usage for sterilization, but may lead to issues with feed inhibition.

4) Cell Mass Value: I assigned a zero value to the cell mass. The value of the cell mass depends upon whether it can be sold or has to be treated as a waste. Cell mass from many industrial fermentations today is often sold into animal feed markets. For example, the cell mass from a corn dry mill ends-up in the DDGS product used primarily for ruminant feeds, while the cell mass from citric acid production can sometimes be sold as a poultry feed ingredient. On the other hand, if the cell mass cannot be sold then it has to be treated as a negative value waste. Digestion, compositing, landfill, and incineration

are some possible options for treatment. Assigning a zero value to the cell mass is a reasonable assumption for a preliminary analysis since its value is unknowable at this time.

5) Nutrient Costs: I assumed zero costs for nutrients. It is difficult to estimate costs since nutrient requirements of the organism are unknown and the nutrient content of the feedstock is unknown. As a comparison, current bioethanol models project a nutrient cost of about $0.03 per gallon.

6) Plant Scope: The analysis assumed a battery limits plant that purchased utilities from a host facility. Capital costs for a stand-alone plant could easily be double the costs for a battery limits plant. This, of course, will vary quite a bit depending upon the details of the situation.

Please let me know if there are any other issues that need to be addressed.

Sincerely,

Tim Eggeman

Tim Eggeman