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Technical and Economic Assessment of Producing Hydrogen by Reforming Syngas from the Battelle Indirectly Heated Biomass Gasifier

Margaret K. Mann

National Renewable Energy Laboratory
1617 Cole Boulevard
Golden, CO 80401-3393
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Industrial Technologies Division
National Renewable Energy Laboratory
Hydrogen Program Milestone Completion Report

Program Manager: Catherine Gregoire Padró

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Executive Summary

The technical and economic feasibility of producing hydrogen from biomass by means of indirectly heated gasification and steam reforming was studied. A detailed process model was developed in ASPEN Plus™ to perform material and energy balances. The results of this simulation were used to size and cost major pieces of equipment from which the determination of the necessary selling price of hydrogen was made. A sensitivity analysis was conducted on the process to study hydrogen price as a function of biomass feedstock cost and hydrogen production efficiency.

The gasification system used for this study was the Battelle Columbus Laboratory (BCL) indirectly heated gasifier. The heat necessary for the endothermic gasification reactions is supplied by circulating sand from a char combustor to the gasification vessel. Hydrogen production was accomplished by steam reforming the product synthesis gas (syngas) in a process based on that used for natural gas reforming. Three process configurations were studied. Scheme 1 is the full reforming process, with a primary reformer similar to a process furnace, followed by a high temperature shift reactor and a low temperature shift reactor. Scheme 2 uses only the primary reformer, and Scheme 3 uses the primary reformer and the high temperature shift reactor. A pressure swing adsorption (PSA) system is used in all three schemes to produce a hydrogen product pure enough to be used in fuel cells. Steam is produced through detailed heat integration and is intended to be sold as a by-product.

Three plant sizes, 27 T/day (30 t/day), 272 T/day (300 t/day), and 907 T/day (1000 t/day) were studied. In Scheme 1, the small plant produces approximately 21,594 standard m^3/day (762,580 scfd) hydrogen, which approximates the fuel requirement of 500 vehicles per day with a fuel economy corresponding to 60 miles per gallon of gasoline (Ogden, 1995). The medium-size plant was chosen for study because it is ten times larger. The large plant corresponds to a plant using half the maximum amount of biomass that has historically been considered to be economically and logistically feasible from a dedicated feedstock supply system (DFSS). The two smaller plants would most likely be able to use waste biomass at a cheaper price than that from a DFSS. When examples of costs are given in this report, biomass is assumed to cost $46.30/T and $15/T from DFSS and waste sources, respectively. The cost of biomass from either waste sources or a DFSS will vary depending on the location and crop type, as well as market influences once biomass energy systems are developed. The Department of Energy goal for biomass from a DFSS is $37.50/T ($34/t = $2/MMBtu) to $46.30/T ($42/t = $2.50/MMBtu).

The steam reforming process studied is very similar to that used to reform natural gas. All necessary unit operations are commercially available, and should require no special engineering design. The estimated capital costs of the entire gasification and reforming plant for the most profitable scenario studied (Scheme 1) are $6.1 million for the 27 T per day (30 t/day) plant, $34.5 million for the 272 T/day (300 t/day) plant, and $90.4 million for the 907 T/day (1000 t/day) plant.

The economics of producing hydrogen from this process are moderately favorable for many of the scenarios tested. The most economically feasible design is that tested in Scheme 1. The necessary selling price for hydrogen produced by steam reforming BCL biomass syngas falls within the current market values ($5 - $15/GJ) for many of the cost scenarios studied. However, the results are mostly on the high end of this range for reasonable biomass feedstock costs. Of the three plant sizes studied, the most economic configuration depends upon the availability of waste biomass at a lower price than biomass from a DFSS. If waste biomass can be obtained for the medium size plant, this scale with a Scheme 1 design yields the lowest hydrogen price. If the medium size plant must use biomass from a DFSS, the large plant with a Scheme 1 design is the most economic.

Results show that the small scale plant using any of the process schemes studied does not produce hydrogen cheaper than the medium size plant. However, if the small plant is the only size for which cheaper waste biomass can be obtained, local refueling stations, similar to existing gasoline stations, might be feasible.

The hydrogen production cost from the large plant obtaining biomass from a DFSS at \$46.30/T (\$42/t), is \$6.50/GJ (\$6.90/MMBtu) without taxes. With a 37% tax rate and a 15% after-tax internal rate of return (IRR), the necessary hydrogen selling price is \$13.70/GJ (\$14.30/MMBtu). The hydrogen production cost for the 272 T/day plant is \$4.10/GJ using biomass waste at \$16.50/T (\$15/t). The corresponding hydrogen selling price is \$13.10/GJ (\$13.80/MMBtu). If the feedstock for the medium size plant must be obtained from a DFSS, the production cost and necessary selling price increases to \$7.20/GJ (\$7.60/MMBtu) and \$16.20/GJ (\$17.1/MMBtu), respectively. Hydrogen produced in the small plant using waste biomass will cost \$7.20/GJ (\$7.60/MMBtu), and sell for \$23.20/GJ (\$24.50/MMBtu). A lower specified IRR would decrease the required selling price in each case and the estimates of what biomass from waste and DFSS sources will cost are likely to vary from the examples given here. Hydrogen produced in process Schemes 2 and 3 is more expensive than that produced in Scheme 1 because of the decrease in production.

The discount rate for which the net present value of the project equals zero was calculated for each scenario studied. This rate is set such that the cumulative net earnings from the project exactly balance the initial investment in the process. Using Scheme 1, the discount rate for the large plant using biomass from a DFSS is 9.2%. The rate for the medium plant using biomass from a DFSS is 6.7%; using biomass waste, this rate increases to 10.4%. The rate for the small plant, even with the cheaper biomass waste feedstock, is 4.0%.

The break-even points for the large, medium, and small plants are 13.3 years, 6.3 years, and 7.2 years, respectively. These calculations were made using DFSS biomass for the large plant and waste biomass for the medium and small plants. If the medium plant must use biomass from a DFSS, the break-even point is extended to 9.5 years.

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Abbreviations and Acronyms

BCL - Battelle Columbus Laboratories
CO - Carbon monoxide
CO_2 - Carbon dioxide
DCFROR - Discounted Cash Flow Rate of Return
DFSS - Dedicated Feedstock Supply System
GJ - Gigajoule
H_2 - Hydrogen
IRR - Internal Rate of Return
kPa - Kilopascal
MMBtu - Million British Thermal Units
PSA - Pressure Swing Adsorption
ROI - Return on Investment
SCFD - Standard Cubic Feet per Day
t - ton
T- Metric tonne

Metric Units of Measurement

In accord with recommendations from the Department of Energy, all results from this study are reported in metric units. Occasionally, the English system equivalent is stated in parenthesis. Below are the metric units used in this report with the corresponding conversions to English units.

Mass:	kilogram (kg) = 2.20462 pounds
	metric tonne (T) = 1.10231 ton
Volume:	cubic meter (m^3) = 264.17 gallons
Pressure:	kilopascals (kPa) = 0.145 pounds per square inch
Energy:	gigajoule (GJ) = 0.9488 MMBtu
Temperature:	°C = (°F - 32)/1.8

1.0 Introduction

The technical and economic feasibility of producing hydrogen by reforming syngas from the Battelle Columbus Laboratory (BCL) indirectly heated gasifier was studied. From experimental work conducted at BCL on the gasifier and commercial information on the reforming operation, a process plant was designed using the ASPEN Plus™ simulation software. The material and energy balances obtained were used to size and cost major pieces of equipment, from which a capital cost estimation was made. Using discounted cash flow rate of return (DCFROR) and return on investment (ROI) analyses, the economic position of this biomass-derived process relative to conventional hydrogen production processes was assessed.

Hydrogen has the potential to deliver significant economic and environmental benefits. Hydrogen is a very clean burning fuel; in internal combustion engines, water and a very small amount of NO_x are the only products. When used in a fuel cell to produce electricity, water is the sole product. Hydrogen can be used to produce energy in every application that fossil fuels are currently used. By 2025, the percentage of energy from oil imports could be reduced from the current 50-60% to less than 25%, if hydrogen energy were only to contribute 10% to the overall energy use.

On a life-cycle basis, the emissions associated with hydrogen depend primarily upon the production route used. Renewable resources, such as solar, wind, and biomass are excellent feedstocks for hydrogen because of their inherently clean nature and sustainability.

This study assesses the technical design and economic feasibility of producing hydrogen from gasification, one of the possible biomass-based routes to hydrogen. Biomass is considered to be anything that has participated in the growing cycle recently. Agriculture waste, forest residue, urban wood waste, and trees and grasses grown as energy crops, are commonly the process feedstocks referred to as biomass. Because biomass consumes as much CO_2 in its growing cycle as is produced when it is transformed to energy, the net CO_2 contribution from biomass-derived fuels to the atmosphere is much less than from fossil-derived fuels. Furthermore, producing biomass on a sustainable basis by growing energy crops supports the U.S. agriculture sector and potentially reduces our oil and gas imports.

The gasification system used for this study was the BCL indirectly heated gasifier. The heat necessary for the endothermic gasification reactions is supplied by circulating sand from a char combustor to the gasification vessel. The syngas, containing primarily CO, H_2, CH_4, CO_2, and some higher hydrocarbons, is then steam reformed to produce H_2 and CO_2 in a process based on that used for natural gas reforming. The$_2$H can be purified and sold as an energy carrier to be used in vehicles, power plants, or refinery applications.

2.0 Process Description

Biomass, obtained either from a DFSS, or as agricultural, urban, or industrial waste, is fed to a rotary dryer to reduce the moisture content from approximately 50% to 11%. The biomass is then gasified in the BCL gasifier which is heated indirectly by sand circulating between a char combustor and the gasification vessel. The product syngas is cooled and compressed to the appropriate conditions for reforming. A reactor known as the primary reformer converts the methane and higher hydrocarbons to CO and performs a significant portion of the water-gas shift reaction to convert CO and water to H_2 and CO_2. The remaining CO is consumed via this reaction in the subsequent high temperature and low temperature shift reactors. A pressure swing adsorption (PSA) system is used to separate hydrogen pure enough for use in fuel cell applications from the shift reactors product gas.

2.1 Gasification Using the BCL Indirectly Heated Gasifier

A schematic of the BCL gasifier is shown in Figure 1. This system was simulated using run data from Battelle Columbus Laboratory. A Fortran subroutine controls the simulation of the gasifier and is shown along with the entire run input file in Appendix A. Biomass and char were simulated as non-conventional components; their ultimate analyses are shown in Table 1. The biomass composition used for this study is typical of woody biomass such as hybrid poplar.

Table 1: Elemental Analysis of Biomass and Char

	Ultimate Analysis (Weight percent, dry basis)	
	Biomass	Char
Carbon	50.88	65.2
Oxygen	41.90	3.03
Hydrogen	6.04	3.70
Nitrogen	0.17	2.47
Sulfur	0.09	28.65
Chlorine	0	0
Ash	0.92	3.04

Biomass of approximately 50 wt% moisture is dried in a rotary drier using a combination of char combustor flue gas and air. The dried biomass, containing 11 wt% moisture, is fed to the fluidized bed gasifier, with hot sand from the char combustor as the bed material. It operates at nearly atmospheric pressure and 825°C (1517°F). Steam rather than air or oxygen is added to the gasifier to produce a syngas of medium quality: 18.35 MJ/m^3 (493 Btu/scf). After a cyclone separator removes the char, the syngas is expected to be cleaned using the hot-gas clean-up processes currently being developed by the Department of Energy (DOE) and Westinghouse; the current technology uses a water scrubber. Hot-gas clean-up would consist of ceramic candle filters to remove particulates from the syngas prior to downstream operations such as reforming. The resultant syngas composition is shown in Table 2.

Figure 1: The Battelle Indirectly Heated Gasification System

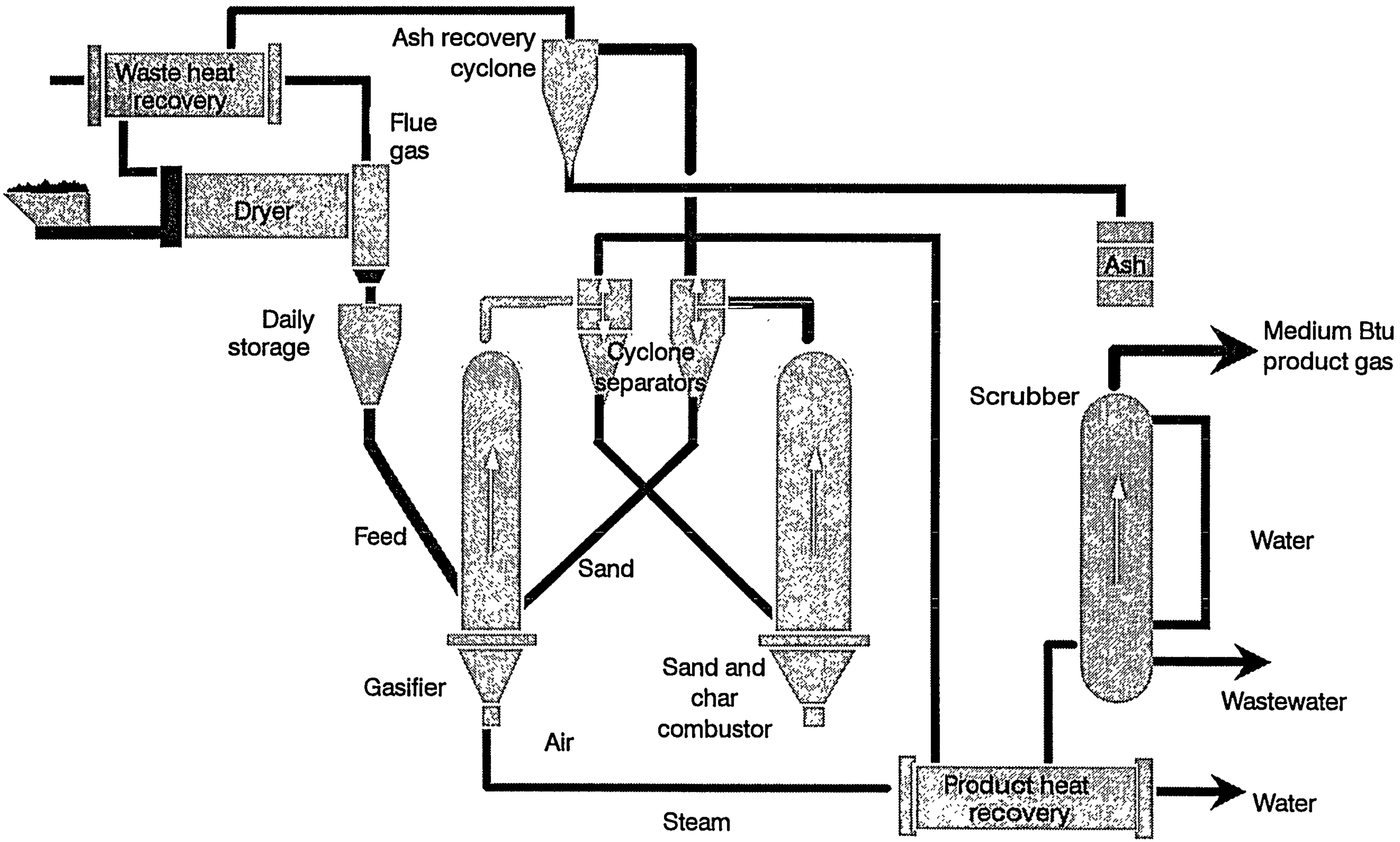


Table 2: Syngas Composition After Cleanup

Component	Volume%
CO	43.17%
H_2	21.22%
CH_4	15.83%
CO_2	13.46%
C_2H_4	4.62%
C_2H_6	0.47%
tar	0.40%
C_2H_2	0.37%
NH_3	0.37%
H_2S	0.08%

After clean-up, the syngas is cooled to 91°C (195°F) so that it can be compressed to the pressure required for the PSA system plus the expected pressure losses in the reactors. During this cooling, the water and higher hydrocarbons (tars) remaining in the syngas will most likely condense and must be removed and pumped before being added again to the compressed syngas. The syngas compressor outlet pressure is 3,654 kPa (530 psi).

2.2 Steam Reforming to Produce Hydrogen

The reforming process, shown in Figure 2, is similar to that used in hydrogen production from natural gas. The major unit operations are a primary reformer to convert methane and the higher hydrocarbons present in the syngas to hydrogen, plus shift reactors to convert CO to hydrogen. The reactions governing the reforming process are shown in the following equations:

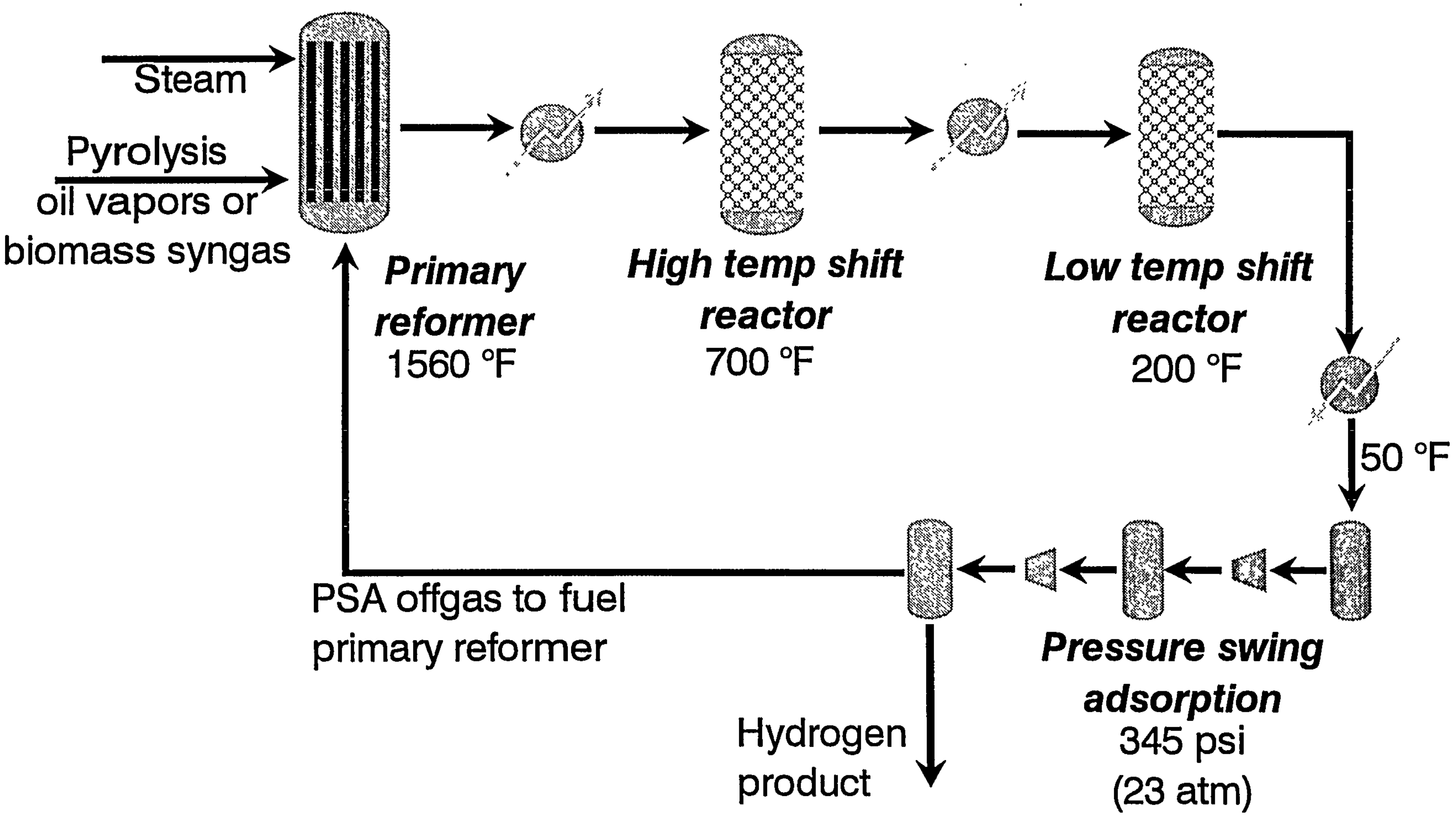
(1) $C_nH_m + nH_2O + \text{heat} \Rightarrow nCO + (m/2 + n)H_2$

(2) $CO + H_2O \Rightarrow CO_2 + H_2 + \text{heat}$

The primary reformer, a reactor similar to a process furnace with catalyst-filled tubes, converts the methane and higher hydrocarbons to CO and H_2 (Reaction 1), and performs a significant portion of the water-gas shift reaction to convert CO and water to H_2 and CO_2 (Reaction 2). The remaining CO is consumed via this reaction in the subsequent high temperature and low temperature shift reactors. A pressure swing adsorption (PSA) system is used to purify the hydrogen.

Reaction (1) typically takes place at temperatures between 800°C and 850°C (1472°F and 1561°F) in the primary reformer. The heat necessary for this endothermic reaction is supplied by combusting the PSA offgas outside of the reactor tubes through which the reactants and products are flowing. These tubes are filled with a commercial nickel-based catalyst. According to results from operating plants, the primary reformer was

Figure 2: Reforming Process



simulated as an equilibrium reactor with an 11 °C approach temperature (Tindall and King, 1991). Reaction (2) is the water-gas shift reaction. According to the thermodynamics of the reforming process, practically all of the tar and C_2H_x species are consumed, 60 mol% of the CH_4 is converted, and there is a 22 mol% net conversion of CO.

Because reaction (2) is exothermic, it is beneficial to convert the remaining CO at a temperature lower than the temperature of the primary reformer. Nearly complete conversion of CO is accomplished in the subsequent high and low temperature shift reactors. The feed to the high temperature shift reactor is cooled to 370°C (698°F) and increases to 435°C (814°F) as the water-gas shift reaction proceeds. The product of this reactor is then cooled to 200°C (392°F) and fed to the low temperature shift reactor that produces a gas at 220°C (430°F) with a dry-basis composition of 61.9% H_2, 34.1% CO_2, 2.9% CH_4, and 1.1% CO.

A steam-to-carbon ratio of three was used for the reforming operations. This is consistent with that used for natural gas reforming. However, higher hydrocarbon feedstocks may require additional steam (Tindall and King, 1991). The higher content of CO in syngas should improve the kinetics of this process over steam reforming natural gas. However, reforming the C_2 and higher compounds could prove more difficult. Actual experimental data will dictate the appropriate steam-to-carbon ratio. The process studied has a great deal of excess heat available from which steam will be produced for export; therefore, a higher reforming steam requirement will not greatly affect the economics of the process.

Before the reformer product stream can be purified in a PSA unit, it must contain at least 70 mol% hydrogen (Anand, 1995). Purifying streams more dilute than this decreases the purity and recovery of the hydrogen. Therefore, part of the PSA product stream is recycled back into the PSA feed. The recovery of hydrogen in the PSA is 85% when purifying a 70 mol% H_2 stream. The incorporation of the recycle loop decreases the overall separation recovery to 77%. The operating pressure of the PSA unit is 2,500 kPa (363 psi).

Three process configurations, or schemes, were studied. Scheme 1 uses all reforming operations typically used in natural gas reforming: the primary reformer, the high temperature shift reactor, and the low temperature shift reactor. Scheme 2 uses only the primary reformer, and Scheme 3 uses the primary reformer and the high temperature shift reactor. All schemes use identical gasification and hydrogen purification processes. Schemes 2 and 3 were studied to assess the profitability of the process if the capital requirements could be lowered at the expense of producing less hydrogen. Because these process configurations are referred to throughout the report as Schemes 1, 2, and 3, Table 3 gives a description of each for easy reference.

Table 3: Summary of Unit Operations Used in Different Process Configurations

Scheme	Reforming operations used
1	Primary reformer, high temperature shift reactor, low temperature shift reactor
2	Primary reformer
3	Primary reformer, high temperature shift reactor

2.3 Steam Generation

In the simulation, gasification and reforming were integrated such that heat available from the reforming operation could generate the steam necessary for gasification as well as a substantial amount of export steam. The process gas was cooled as it moved between the primary reformer, the shift reactors, and the PSA unit, generating steam in each step. Steam was also generated by cooling the primary reformer flue gas. The majority of the steam produced was superheated at 690 kPa (100 psig); the steam produced by cooling the process gas between the high and low temperature shift reactors was at 3,450 kPa (500 psig).

A complete process flow diagram corresponding to the simulation is shown in Figure 3. Appendix C contains stream data for the 907 T/day plant.

3.0 Conversion Efficiency

Two methods were used to estimate the efficiency of producing hydrogen from biomass by this process. The first method looks at the ratio of the amount of hydrogen that was produced to the stoichiometric maximum amount of hydrogen possible according to reactions (1) and (2). The second method calculates the ratio of the energy value of the product hydrogen and export steam to the energy value of the biomass feed plus purchased electricity. The amount of hydrogen produced for each plant size and scheme is shown in Table 4.

Table 4: Hydrogen Produced in Each Scenario Analyzed (standard m^3/day)

	Scheme 1	Scheme 2	Scheme 3
27 T/day biomass	21,600	16,850	20,440
272 T/day biomass	215,940	168,500	204,390
907 T/day biomass	719,800	561,650	681,280

3.1 Stoichiometric Maximum Efficiency Calculation

The "molecular formula" of biomass can be approximated as $CH_{1.47}O_{0.67}$ on a dry basis. Completely steam reforming this biomass yields 2.07 moles of hydrogen per "mole" of biomass as shown by the following stoichiometry:

$$CH_{1.47}O_{0.67} + 1.33H_2O \Rightarrow CO_2 + 2.07H_2$$

This hydrogen yield is equivalent to 2.02 standard m^3 hydrogen/kg biomass (32.38 scf/lb). In Scheme 1, with all reforming operations used, 0.79 standard m^3/kg (12.71 scf/lb) hydrogen is produced. This corresponds to a 39.3% conversion and recovery efficiency.

ASPEN Plus Proce
Producing Hydrogen by Steam

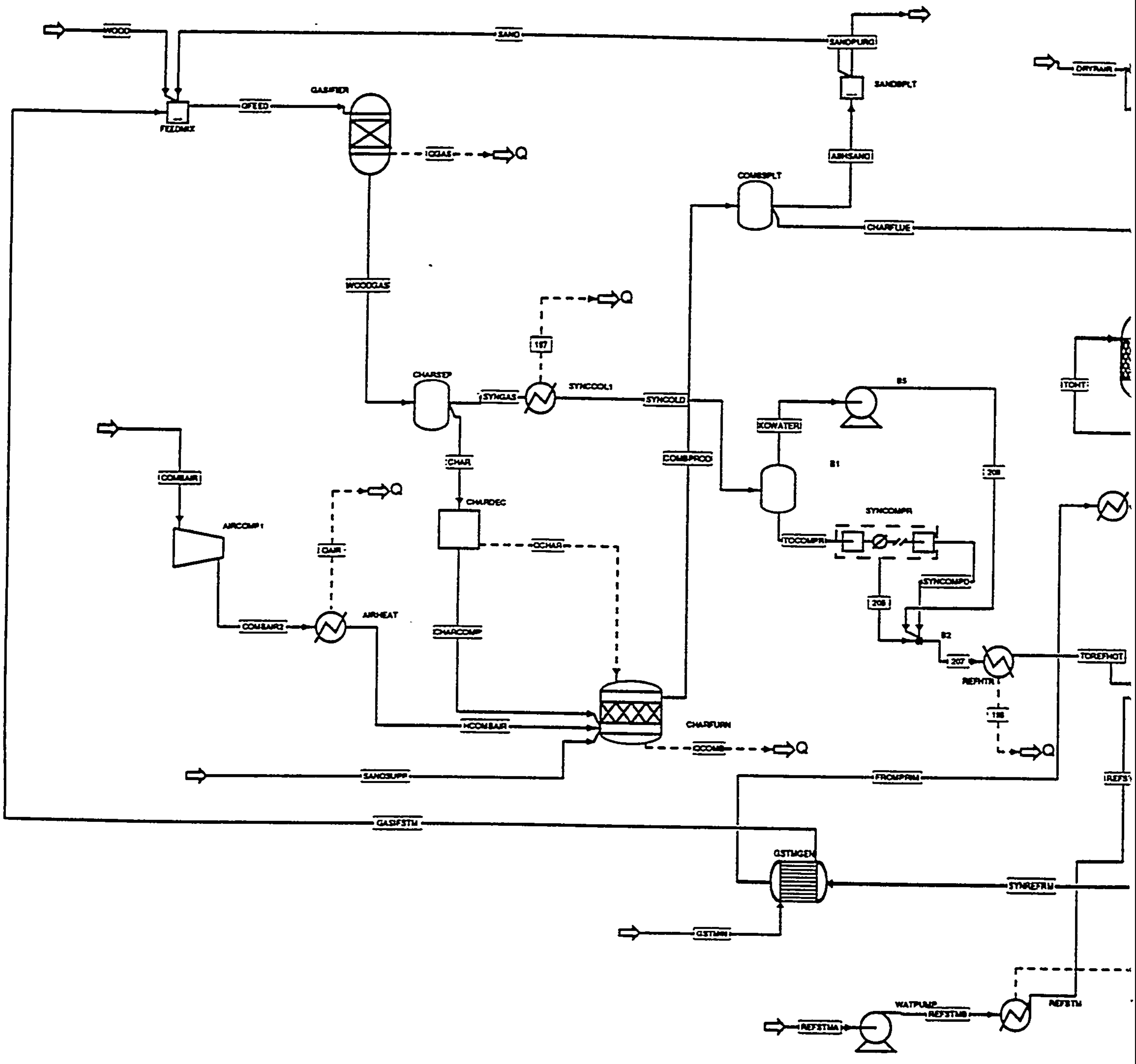
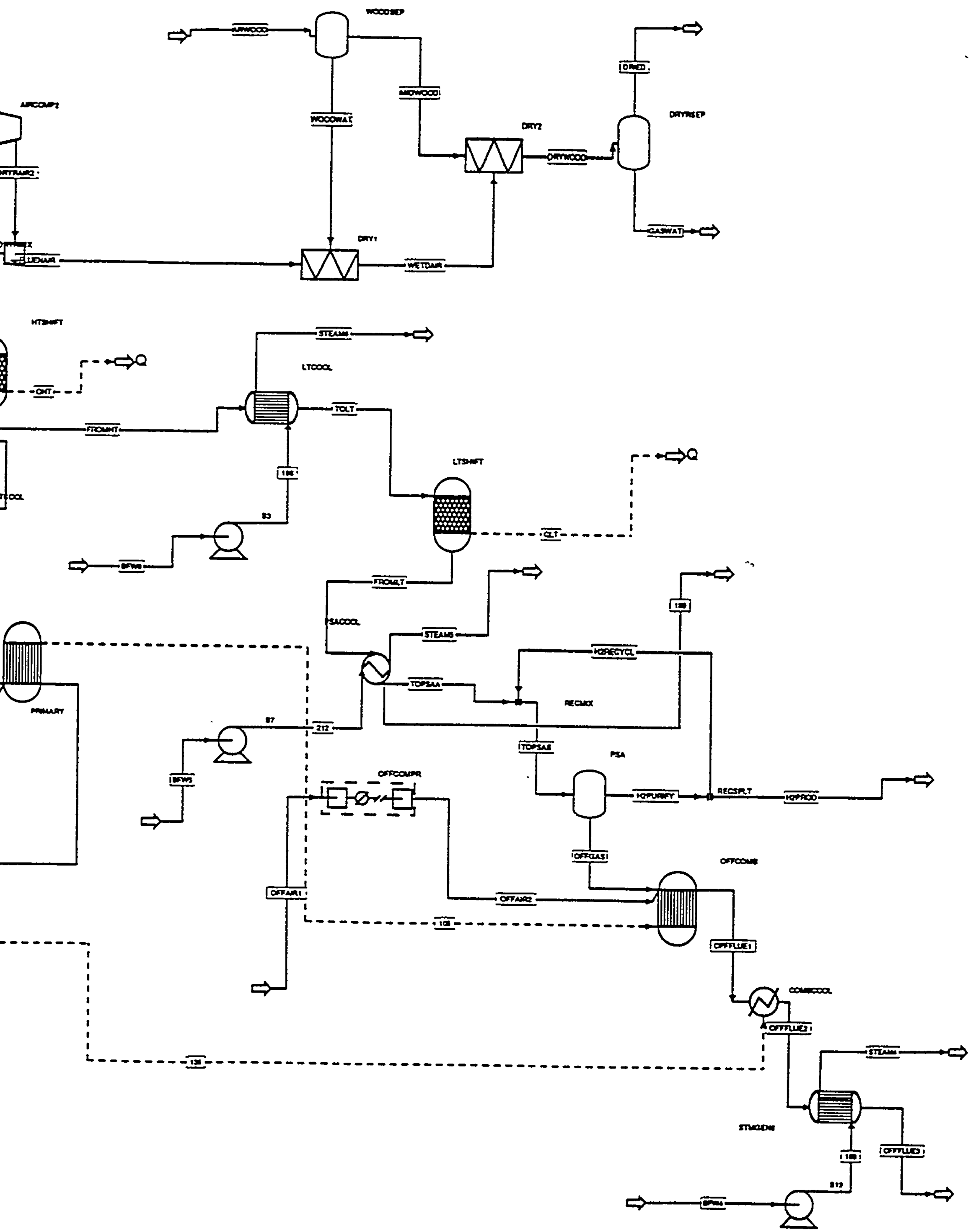


Figure 3

s Flow Diagram
Reforming Biomass Syngas



3.2 Energy Conversion Efficiency Calculation

The efficiency of this process on an energy in, energy out basis can be calculated by the following formula:

$$\frac{(H_2)(HHV_{H2}) + (stm_{ex})(\Delta H_{sh})}{(B_f)(HHV_b) + e}$$

where:

H_2 = hydrogen recovered (kg)
HHV_{H2} = higher heating value of hydrogen (GJ/kg)
ΔH_{sh} = difference in enthalpy between incoming water and steam produced (GJ)
stm_{ex} = steam produced to be sold (kg)
B_f = biomass fed to process (kg)
HHV_b = higher heating value of biomass (19.75 MJ/kg = 8,500 Btu/lb)
e = electricity imported for process requirements (GJ equivalent)

The efficiencies calculated for the three schemes are shown in Table 5.

Table 5: Process Conversion Efficiencies

	Stoichiometric Efficiency	Energy Conversion Efficiency
Scheme 1	39.3%	79.0%
Scheme 2	30.6%	69.7%
Scheme 3	37.2%	76.5%

4.0 Economic Analysis

The current market value of hydrogen is between $5 and $15/GJ. By calculating the economics of the process being studied and comparing the results to this current hydrogen market, the potential profitability can be assessed. Possible sources of error in this analysis are in equipment cost estimation, feedstock and product market predictions, and invalid economic assumptions. The total error can be reduced by looking at ranges of profitability, such as the range of hydrogen selling price versus a range of biomass feedstock costs. As more information on the development of biomass-based technologies becomes available, this analysis can be modified to give a more representative process cost.

The economic feasibility of producing hydrogen by steam reforming syngas from the BCL gasifier was studied using the DCFROR method. This method calculates the IRR that will be earned on the initial capital investment over the life of the project. Given this rate and a feedstock cost, the necessary selling price of the product can be calculated. Often, the IRR is specified as the minimum acceptable rate for an investor to finance a project. Therefore, the perceived risk of the project can be incorporated into the IRR. Because the process of producing hydrogen from biomass currently carries higher risks than conventional hydrogen-generating processes, the IRR specified in this study was 15% after tax, while the rate for conventional processes is between 9% and 12%. For a 37% tax rate, a 15% after-tax IRR corresponds to a pre-tax IRR of 20.3%.

As an alternative means of measuring the economic feasibility of this process, an ROI analysis was also performed. The ROI is the sum of the net present value of each project year's revenue, divided by the initial capital investment. The discount rate used to bring all revenues and costs to the value of what money is worth today (or at any defined time) is set so that the ROI equals zero. This practice is also known as setting the net present value equal to zero.

4.1 Economic Assumptions

The economic analysis for this study was based on current dollars and performed using equity financing, assuming that the capital will not be borrowed. The latter assumption is probably valid for the smaller-scale plants, less so for the large plant. The majority of the assumptions used in performing the economic analysis are shown in Table 6. Other assumptions, such as the percentage of the purchased equipment cost spent on piping, can be found in the cost sheets in Appendix B.

Table 6: Economic Assumptions

January, 1995 dollars
Equity financing
20 year plant life
Two year construction period
90% on-line factor
Royalties = 0.5% of sales
Inflation rate = 5%
Tax rate = 37%
Straight-line depreciation for ten years; first and last year at 50% of other years
50% plant capacity first year of production
30% of capital investment spent first year, 70% second year

4.2 By-Product Credit: Steam

A by-product credit was taken for the steam generated in the process. A selling price of \$7.88/1000 kg (\$3.57/1000 lb) was assumed for 3,450 kPa (500 psig) steam. A price of \$5.18/1000 kg (\$2.35/1000 lb) was assumed for 690 kPa (100 psig) steam (Peters and Timmerhaus, 1980). All steam produced contains 17 °C superheat. The amount of steam produced is shown in Table 7. The amount of steam generated for Schemes 2 and 3 is not expected to be significantly different from that generated in Scheme 1 because the same or higher amount of heat will be available. The assumption that the steam will be able to be sold is probably valid for the medium and large plants as they will most likely be located in more industrialized centers to take advantage of other infrastructure. However, it may be difficult to sell the steam produced by the small plant, as this size represents small refueling stations located near the demand for hydrogen.

Table 7: By-Product Steam Generated

Source of Heat (stream name in Figure 3)	Amount Produced, kg steam / kg dry biomass	Pressure, kPa (psig)
Cooling gas between high and low temperature shift (FROMHT)	0.32	3,450 (500)
Compression of syngas (TOCOMPR)	1.26	690 (100)
Compression of air fed to offgas combustor (OFFAIR1)	0.12	690 (100)
Cooling offgas combustor flue gas (OFFFLUE)	0.43	690 (100)
Cooling gas going to PSA (TOPSA)	0.85	690 (100)

4.3 Equipment Sizing and Costing

The material and energy balance results from the ASPEN Plus™ simulation were used to determine the size and corresponding costs of major pieces of equipment for the process. Costs were taken from the ChemCost software package and published literature and brought to January 1995 dollars using equipment cost escalation ratios from Chemical Engineering Magazine (March, 1995). Some costs, especially those related to gasification, were taken from other studies. Detailed cost results can be found in the cost sheets in Appendix B.

4.3.1 Gasification Costs

The cost of the gasification train was estimated in a previous study for DOE's Biomass Power Program, as well as by several consulting firms working for BCL (Breault and Morgan, 1992; Double, 1988; Dravo Engineering Companies, 1987; Weyerhaeuser, 1992). These costs were scaled to the appropriate plant size for this study using a 0.7 scale factor. Unit operations included in these costs were the feed system, dryer, gasifier, char combustor, cyclone separators, hot-gas cleanup system, and necessary pumps and compressors. The gasification steam generator (heat exchanger and flash drum) cost was calculated separately and included with the reformer costs.

4.3.2 Reactor Costs

The reactors sized and priced for this study were the primary reformer, the high temperature shift reactor, and the low temperature shift reactor.

A thermodynamically-controlled reactor block was used to model the primary reformer in ASPEN Plus™. This block predicts the final reaction products based on minimization of Gibbs free energy. An 11 °C temperature approach to equilibrium was used in accordance with results from natural gas reforming operations (Tindall and King 1991). The heat for the endothermic reactions (Reaction 1) taking place in this reactor was supplied by burning the PSA offgas that consists of unrecovered hydrogen, CH_4, CO and inerts. An equilibrium block was also used to model this combustor, and taken together, the two reactor blocks represent the primary reformer. The cost of the primary reformer was based on a furnace reactor, taken from three sources and averaged (see Appendix B cost sheets).

The costs of the high and low temperature shift reactors were based on a space velocity of 4000/hr (Kirk-Othmer, V.13, p 856) and a height-to-diameter ratio of 2. The cost of the reactor as a function of height, diameter, and material of construction was determined using ChemCost. Costs from other sources were similar.

4.3.3 Compressor Costs

The two compressors necessary for this process are the syngas compressor and the air compressor for the offgas combustor. Both are four stage compressors with interstage cooling. As noted in Table 7, steam is generated by cooling the gas being compressed between stages. The costs of these compressors were calculated by ChemCost as eight individual compressors of the required horsepower. See cost sheets in Appendix B for the power requirements and resultant costs. The interstage coolers were calculated separately as heat exchangers and flash drums.

4.3.4 Heat Exchanger Costs

Heat exchangers were modeled as counter-current in the simulation. The minimum approach temperature used was 11°C. The required area for a heat exchanger was calculated from the appropriate heat duty, temperature difference, and heat transfer coefficient. ChemCost was used to derive the corresponding cost.

4.3.5 Pump Costs

The cost of each pump required for the process was calculated by ChemCost using the flowrate and outlet pressure.

4.3.6 PSA Cost

The appropriate PSA design is very specific to the application, therefore, the manufacturers would most likely design and cost it for the potential buyer. Because this study is only to assess feasibility and not to design a planned operation, the capital and operating costs of the PSA unit were taken from the literature (Schendel, et al, 1983) and scaled according to the amount of hydrogen produced. The installed capital of a PSA system was \$7.164/thousand standard m^3/d (\$253/thousand scfd). The operating costs were \$0.184/thousand standard m^3/d (\$6.50/thousand scfd).

4.3.7 Operating Costs

Operating costs for this process include the feedstock, electricity to run the compressors (\$0.05/kWh), water for steam generation and cooling (\$330/m^3), and labor. The revenue from steam produced for export is taken as an operating cost credit. Detailed operating costs for each plant can be found in the cost sheets in Appendix B.

4.4 Economic Analysis Results

The capital and operating costs for each of the scenarios studied are shown in Table 8. These costs were calculated using a feedstock cost of \$16.50/T for the small and medium size plants and \$46.30/T for the large plant. Operating costs would increase significantly if the medium size plant obtained its biomass from a DFSS.

Table 8: Capital (MM$) and Operating Costs (MM$/year)

	Scheme 1			Scheme 2			Scheme 3		
Plant size	sm	med	lg	sm	med	lg	sm	med	lg
Operating Costs	0.30	1.73	14.1	0.28	1.43	13.1	0.31	4.39	14.1
Fixed	0.15	0.27	0.33	0.16	0.27	0.33	0.16	0.27	0.33
Variable	0.16	1.60	5.33	0.13	1.30	4.32	0.16	1.60	5.32
Byproduct credit	-0.16	-1.62	-5.38	-0.16	-1.62	-5.38	-0.16	-1.62	-5.38
Feed	0.15	1.48	13.8	0.15	1.48	13.8	0.15	4.14	13.8
Capital Costs	6.08	34.5	90.4	5.05	29.3	80.0	6.02	34.0	89.1

The results of the DCFROR analysis are shown in Figures 4 through 9. Because the eventual price of biomass needed to supply plants such as those studied here is unknown, these figures give the biomass feedstock cost that can be paid to produce hydrogen within the current market values. Conversely, if the biomass cost can be accurately assessed, the necessary hydrogen selling price from this process can be obtained. Figures 4, 6, and 8 show the pre-tax production cost of hydrogen in each of the three schemes at the different plant sizes. Figures 5, 7, and 9 show the necessary hydrogen selling price given a 37% tax rate and a 15% after-tax IRR. With the market value of hydrogen between $5/GJ and $15/GJ, these figures show that hydrogen can be produced from biomass to compete with current hydrogen production methods on the large and medium scale for all configurations studied. However, the necessary hydrogen selling price is highly dependent upon the biomass feedstock cost, and low-cost biomass will need to be obtained to justify hydrogen from this process, particularly on the small scale. For easy reference, a summary of these results is shown in Table 9. Hydrogen selling prices were calculated using biomass feedstock prices of $16.50/T for the small plant and $46.30/T for the large plant. Because it is uncertain if nearly 300 T/day waste biomass at the lower price can be secured, both $16.50/T and $46.30/T were used to calculate the minimum selling price for the medium size plant. The results shown in this table are representative only of the situations for which biomass can be obtained at the listed prices. The actual biomass price for a given region and situation should be used when assessing the necessary hydrogen selling price from this process. Figures 4 through 9 can be used when assessing specific situations.

Figure 4: Production Cost of Hydrogen From Steam Reforming Biomass Syngas (Scheme 1), Pre-tax

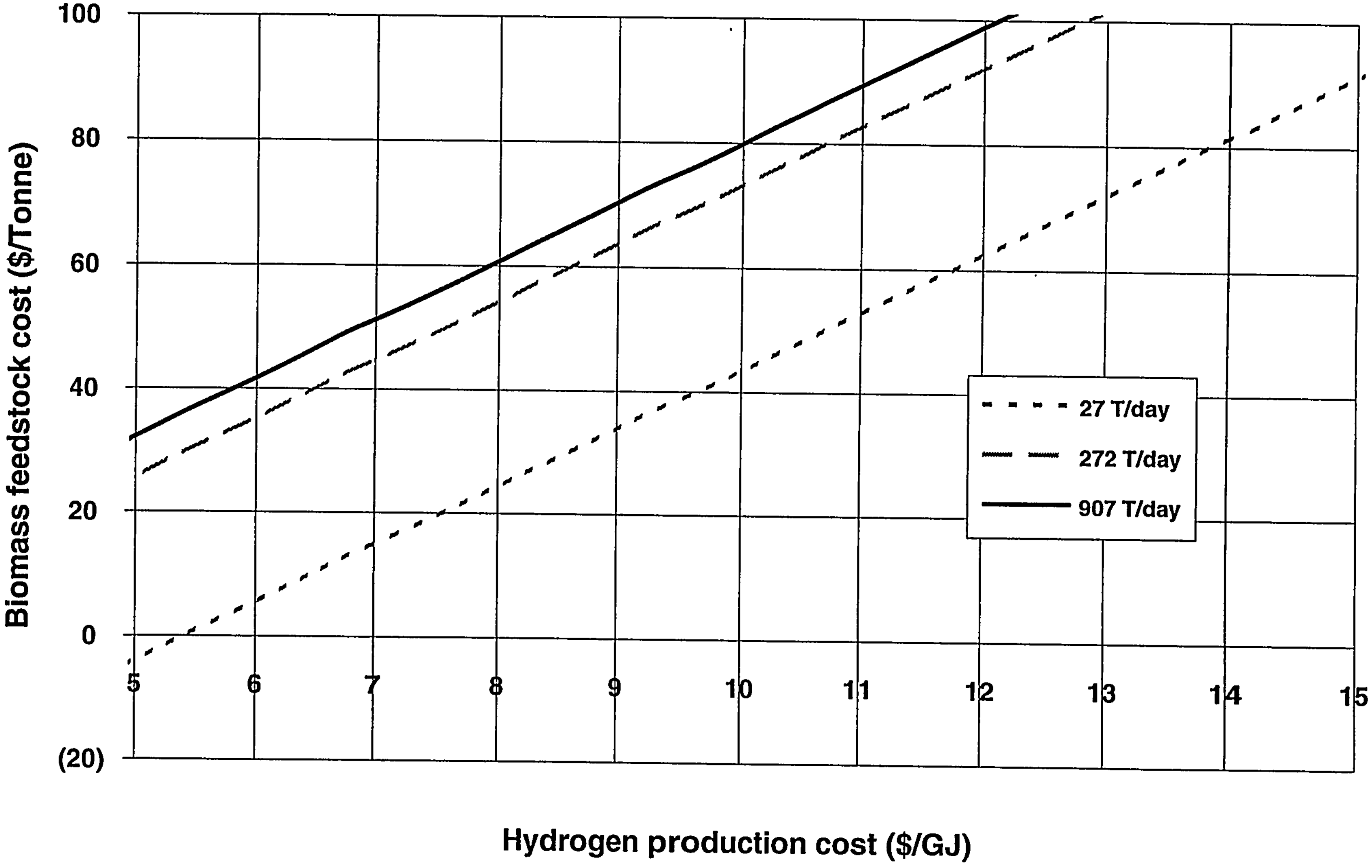


Figure 5: Selling Price of Hydrogen From Steam Reforming Biomass Syngas (Scheme 1), After Tax, 15% IRR

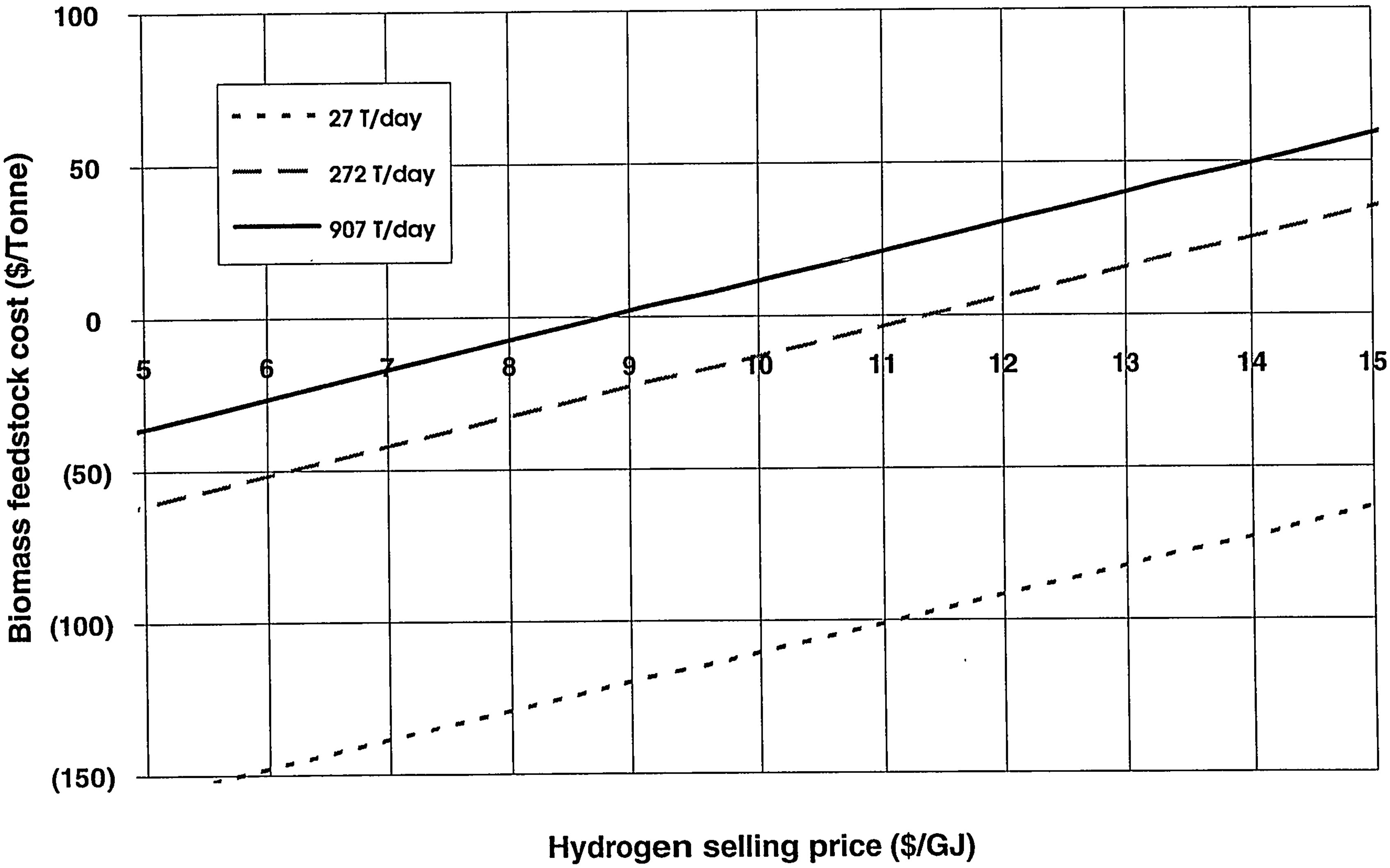


Figure 6: Production Cost of Hydrogen From Steam Reforming Biomass Syngas (Scheme 2), Pre-tax

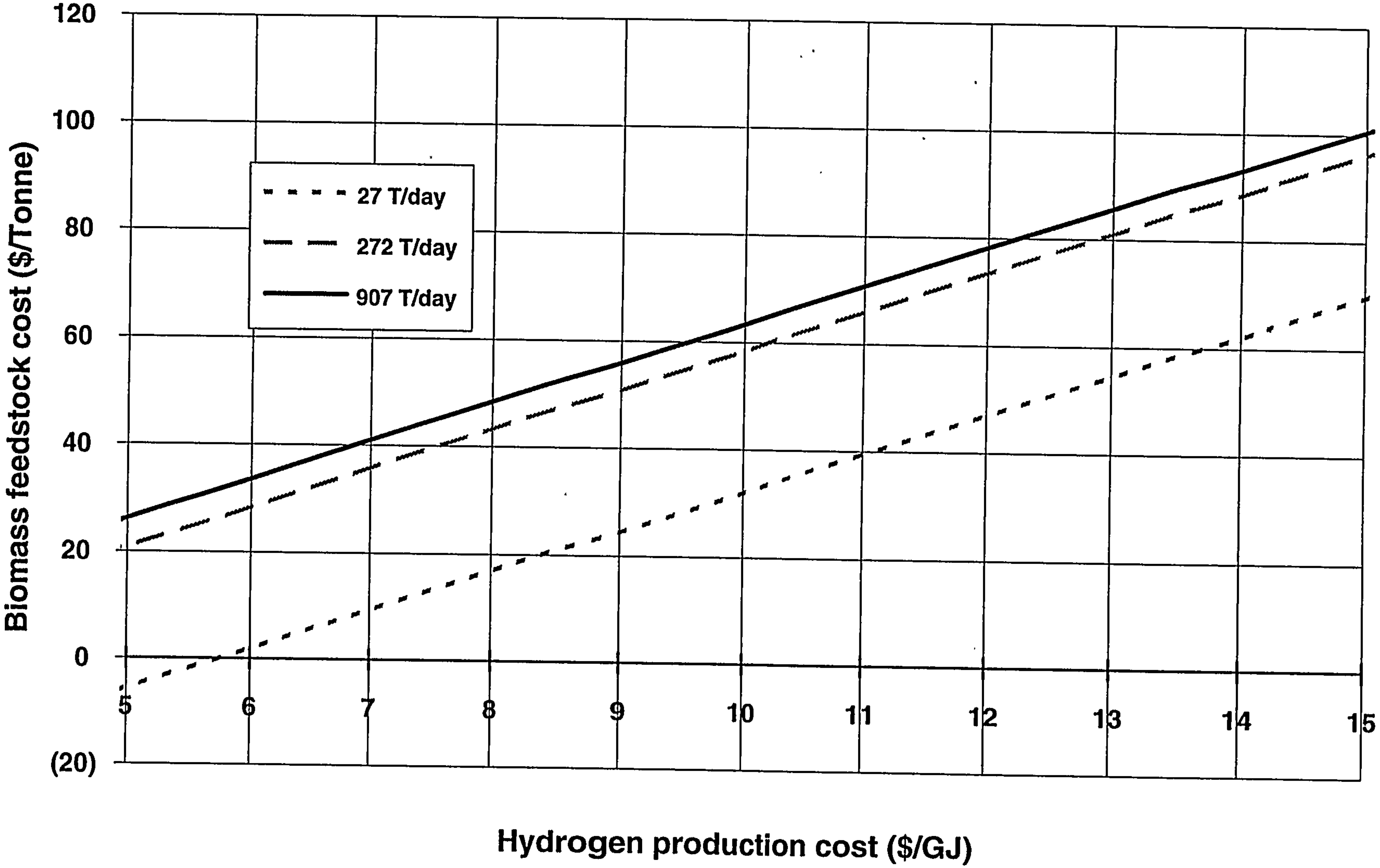


Figure 7: Selling Price of Hydrogen From Steam Reforming Biomass Syngas (Scheme 2), After Tax, 15% IRR

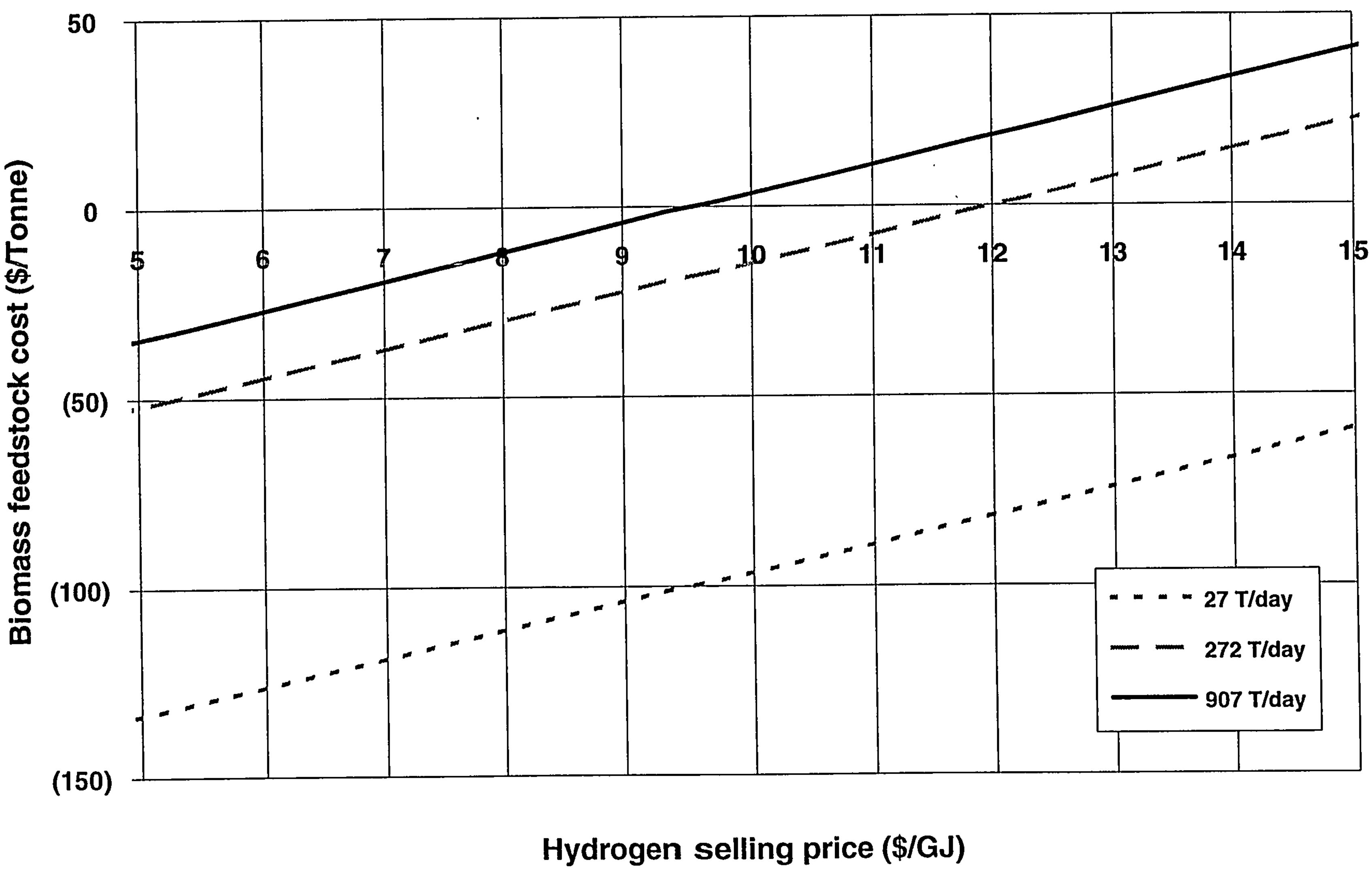


Figure 8: Production Cost of Hydrogen From Steam Reforming Biomass Syngas (Scheme 3), Pre-tax

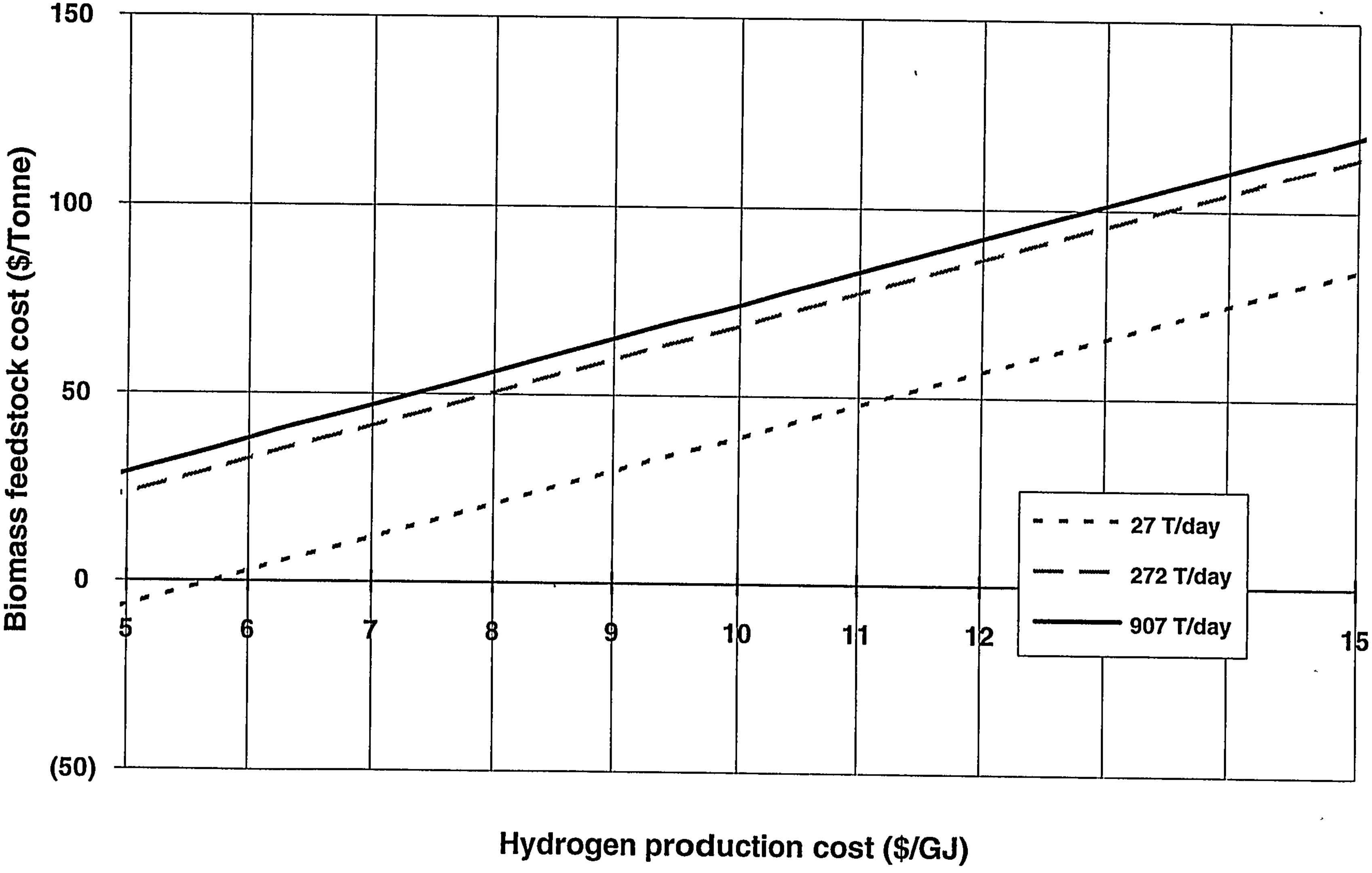


Figure 9: Selling Price of Hydrogen From Steam Reforming Biomass Syngas (Scheme 3), After Tax, 15% IRR

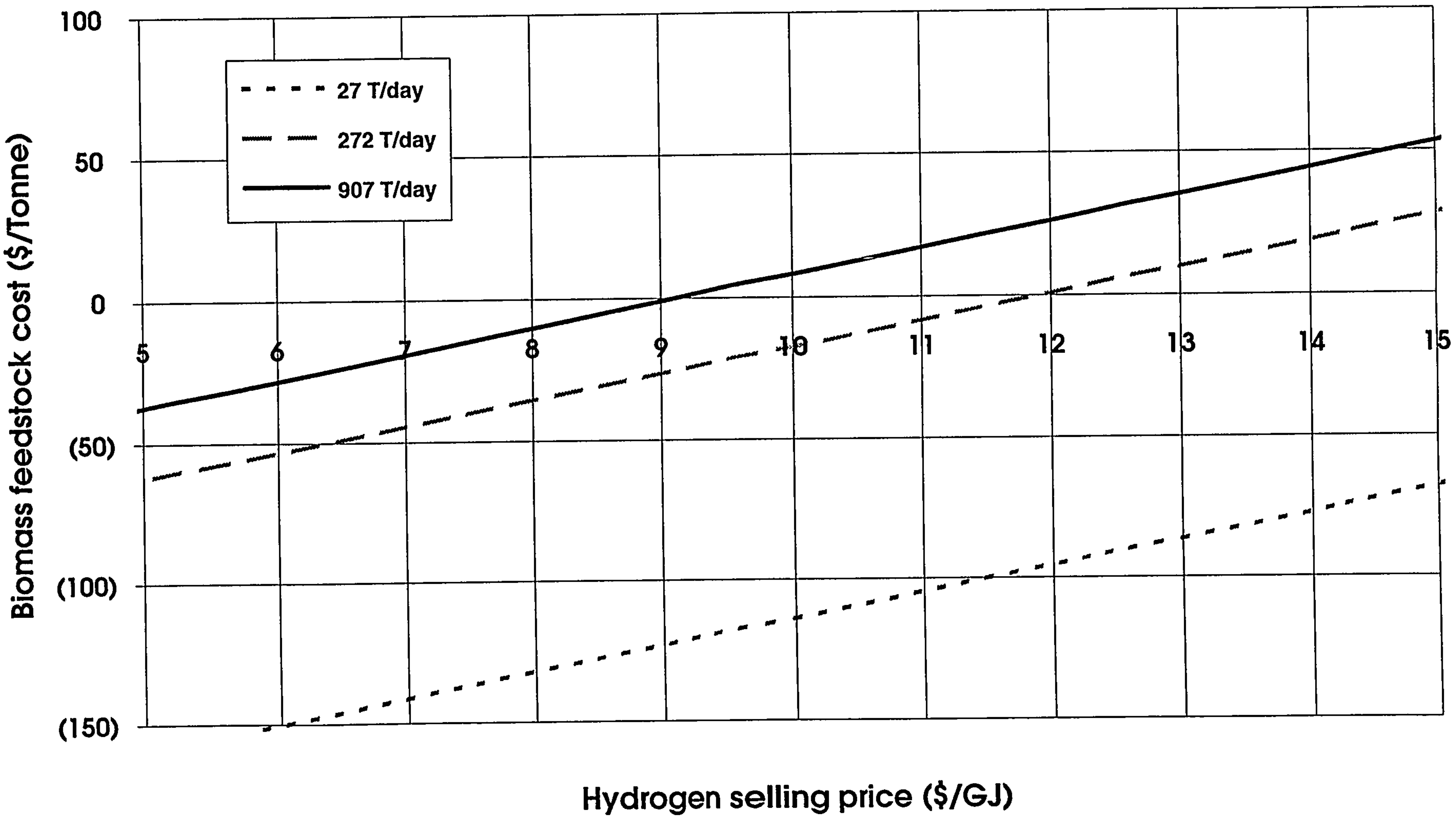


Table 9: Necessary Hydrogen Selling Price for a 15% after-tax IRR After Taxes

	Plant size	Biomass cost ($/T)	Hydrogen selling price ($/GJ)
Scheme 1	small	16.50	23.20
	medium	16.50	13.10
	medium	46.30	16.20
	large	46.30	13.70
Scheme 2	small	16.50	25.10
	medium	16.50	14.20
	medium	46.30	18.20
	large	46.30	15.70
Scheme 3	small	16.50	24.30
	medium	16.50	13.70
	medium	46.30	17.00
	large	46.30	14.20

From Figures 4 through 9, it can be seen that the cost of producing less hydrogen in Schemes 2 and 3 is higher than the savings obtained by eliminating some unit operations in the reforming section of the process. Of the three process configurations studied, the most profitable is Scheme 1. Of the two configurations with reduced reforming operations, Scheme 3, with the primary reformer and high temperature shift reactor, is more economic than Scheme 2 with only the primary reformer. This is because the majority of the water gas shift reaction takes place in the high temperature shift reactor. Scheme 3 is not as economic as Scheme 1 even though capital costs are lower because of the absence of the low temperature shift reactor, the decreased amount of hydrogen that is produced reduces the net income over the life of the plant.

The most economic size for the process studied depends upon the feedstock cost. If the medium size plant can be supplied with waste biomass at a cheaper price (i.e., $16.50/T) than the biomass supplied by a DFSS, the necessary hydrogen selling price from the medium size plant is lower than that from the large plant. However, if the medium and large plants must both use biomass from a DFSS, the larger plant is more economically feasible. The medium size plant is more economic than the small plant if biomass at the same feedstock cost is used in each. Figures 4 through 9 also show that there is a larger economy of scale realized in going from the small to the medium size plant than in going from the medium to the large plant. Figure 5 shows that for positive biomass feedstock costs, the necessary hydrogen selling price would have to be at least $8.70/GJ and $11.20/GJ for the large and medium size plants, respectively. Unless biomass at extremely low costs can be obtained, hydrogen produced in the small indirectly heated gasification and reforming operation is not economically feasible. Figures 7 and 9 give similar results.

Figures 10 through 12 show the cumulative cash flow for the three plant sizes for Scheme 1 over a twenty-year plant life. The corresponding curves for Schemes 2 and 3 are similar. Figures 10 and 11 show the cash flow for the 27 and 272 T/day plant using a biomass feedstock cost of $16.50/T. Figure 12 is the cash flow diagram for the 907 T/day plant using a feedstock cost of $46.30/T. Each diagram is based on a hydrogen selling price of $11/GJ ($12/MMBtu). The break-even point for the medium plant (6.3 years) is sooner than that for the large plant (7.2 years) because of the lower feedstock cost. If the medium plant were also using biomass at $46.30/T, the break-even point would be 9.5 years.

Table 10 gives the discount rate used to set the net present value to zero for each scheme at the three plant sizes. This rate was calculated using a hydrogen selling price of $11/GJ ($12/MMBtu). The biomass costs used were $16.50/T for the small plant and $46.30/T for the large plant; the discount rate for the medium plant was calculated using both biomass costs. The rates obtained in this analysis are low in comparison to other processes which reach commercialization. A better estimate for how much the biomass feed will cost will reduce some of the uncertainty in these calculations, and it may be that the discount rates are higher than reported here.

Table 10: Discount Rate Obtained from Return on Investment Analysis

	Plant size	Biomass cost ($/T)	Discount rate for NPV=0 (%)
Scheme 1	small	16.50	4.0
	medium	16.50	10.4
	medium	46.30	6.7
	large	46.30	9.2
Scheme 2	small	16.50	3.0
	medium	16.50	9.4
	medium	46.30	4.7
	large	46.30	6.6
Scheme 3	small	16.50	3.5
	medium	16.50	9.9
	medium	46.30	6.0
	large	46.30	8.4

Figure 10: Cumulative Cash Flow for a 27 T/day Biomass Gasification and Reforming Facility

Feedstock cost = $16.50/T

Hydrogen selling price = $12/GJ

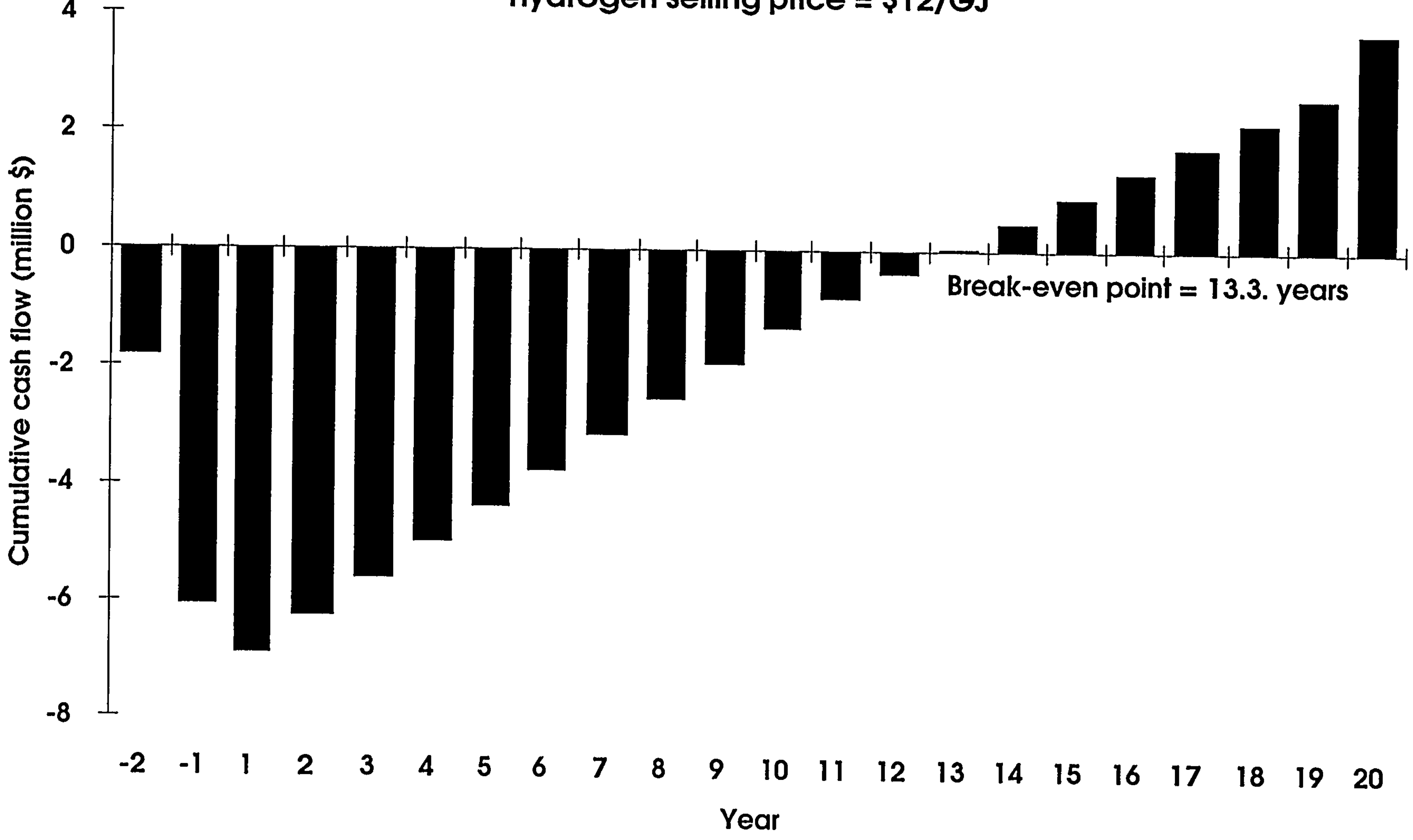


Figure 11: Cumulative Cash Flow for a 272 T/day Biomass Gasification and Reforming Facility

Feedstock cost = $16.50/T

Hydrogen selling price = $12/GJ

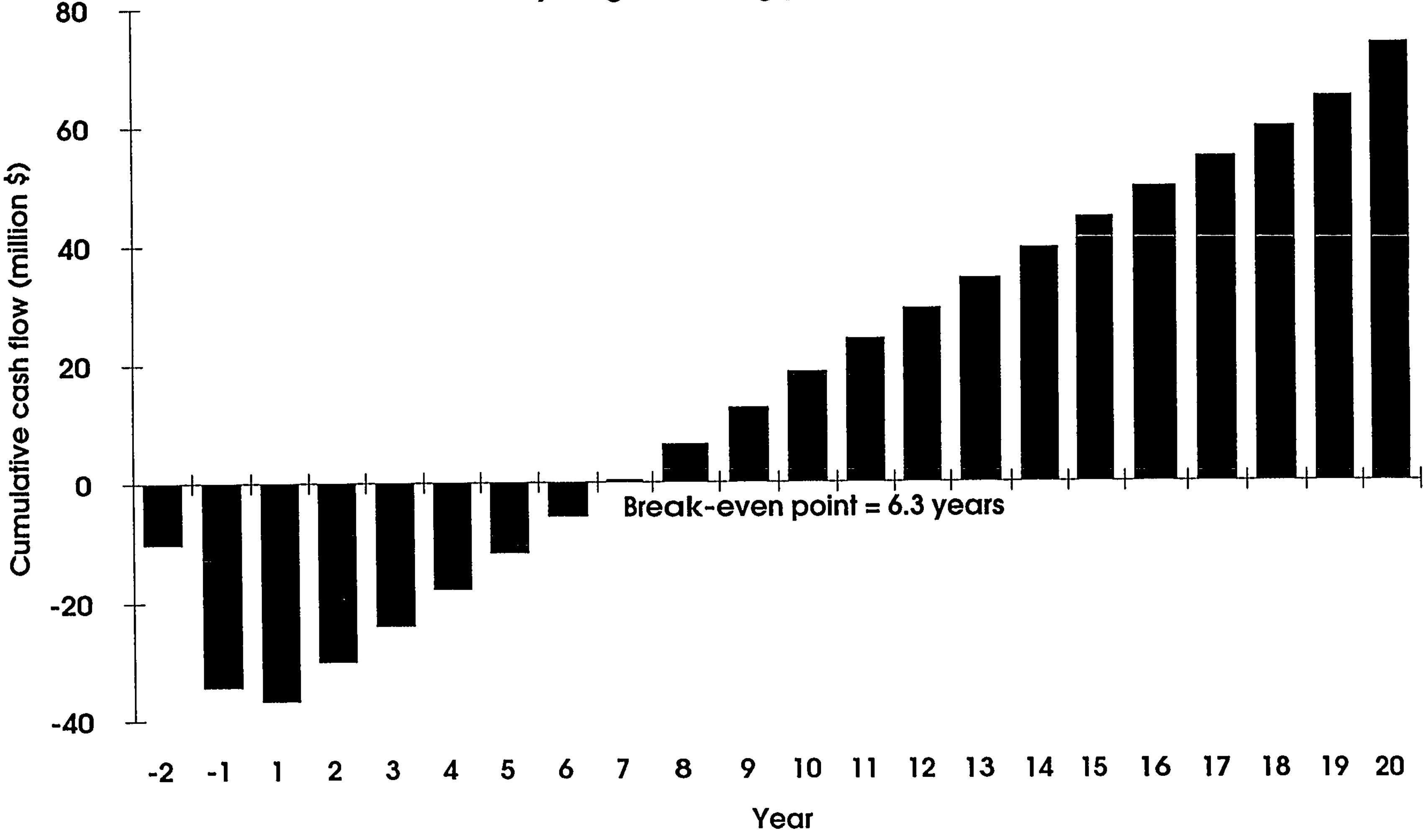
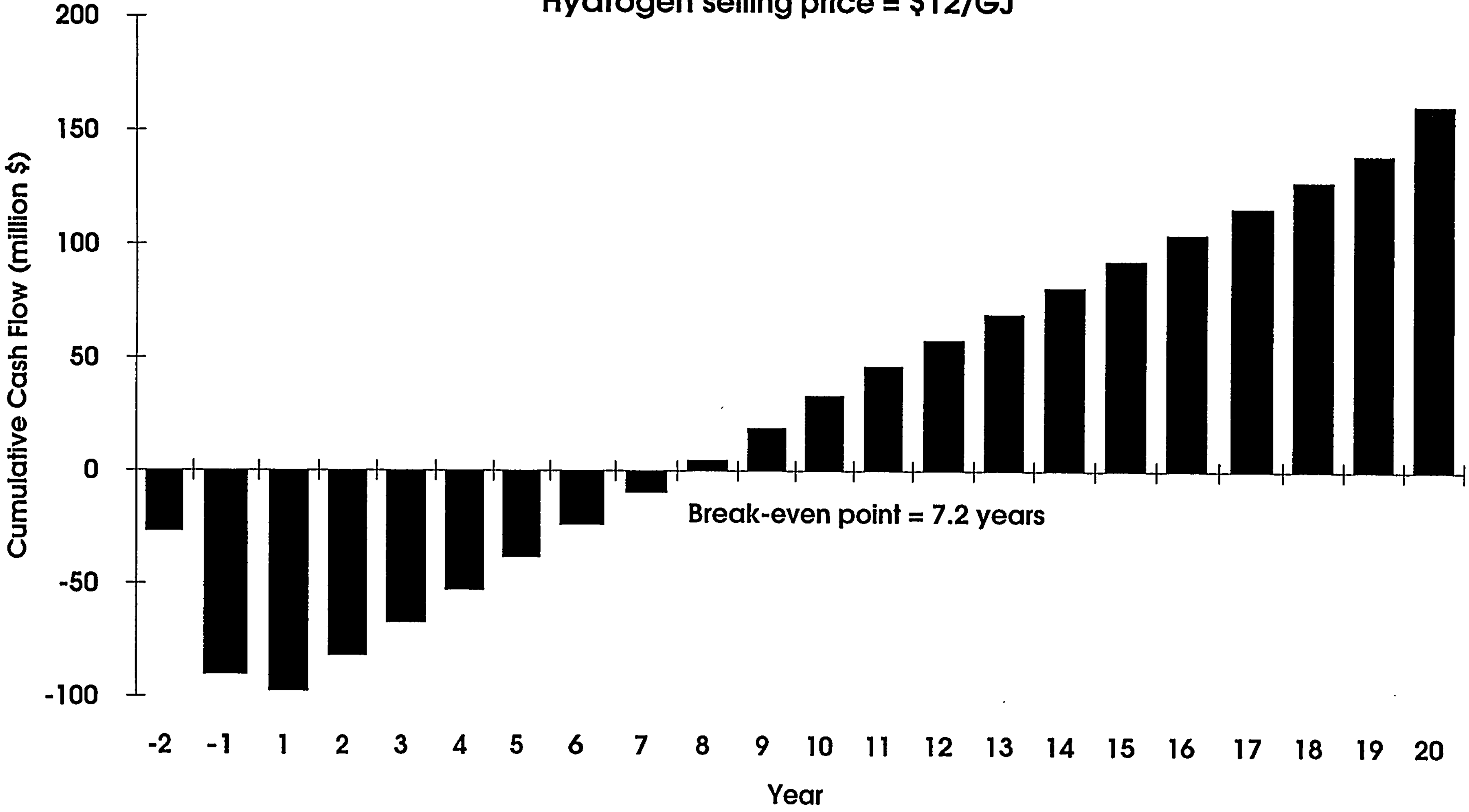


Figure 12: Cumulative Cash Flow for a 907 T/day Biomass Gasification and Reforming Facility


Feedstock cost = $46/T

Hydrogen selling price = $12/GJ

Break-even point = 7.2 years

5.0 Conclusions

Compared to conventional hydrogen producing processes, many of the criteria for successful commercialization of this process are not met. This is because the gasification technology is not fully optimized and tested, and that the reforming process was developed specifically for natural gas and not biomass syngas. Additionally, a higher and more conservative IRR was used for the DCFROR analysis. The necessary selling price of hydrogen from this process falls at the higher end of the current market value range. Also, the discount rates obtained in the ROI analysis are low and the break-even times are fairly long. Therefore, improvements in process design and conversion yields will be necessary for this process to be readily commercialized.

The necessary selling price of hydrogen produced by steam reforming syngas from the BCL gasifier falls within current market values for many of the scenarios studied. The factors that determine which scenario is the most economically feasible are design configuration, plant size, and biomass feedstock costs. The configuration that produces the least expensive hydrogen is that which uses all reforming operations, Scheme 1. Therefore, for the reforming process studied, as much hydrogen as possible should be made at the expense of higher capital equipment charges. Of the three plant sizes studied, the most economic configuration depends upon the availability of waste biomass at a lower price than biomass from a DFSS. If waste biomass can be obtained for the medium size plant, this scale is the most economic. However, if DFSS biomass must be used for both the medium and large plants, the 907 T/day plant produces hydrogen at a cheaper price than the 272 T/day plant. Results show that unless biomass can be obtained at very low prices, producing hydrogen from a very small plant acting as a local refueling station will not be economically competitive.

As the development of biomass-based technologies continues and better predictions for the costs of biomass from energy crop improvements can be made, the examples of costs given in this study can be revisited using the curves of hydrogen price versus feedstock cost. For this process to be economically viable in the marketplace, low biomass costs will probably be necessary. As research continues on processes that use biomass and as uncertainties are addressed, the risk of investing in such projects will decrease. This will reduce the necessary hydrogen selling price and provide a shorter break-even point.

6.0 Future Work

Additional benefits of producing hydrogen on the small scale via reforming syngas from the BCL gasifier should be studied and incorporated into the estimated cost. For example, if hydrogen is produced at the point of its intended use, compression, storage, and transportation costs will be lower than on the larger production scale, or eliminated completely. On the medium and large scales, this cost mitigation is less likely, thus making on-site hydrogen production more attractive than shown in this report.

The ASPEN Plus™ model of the reforming operation will be optimized to increase hydrogen production efficiency and reduce costs. Areas targeted for improvement will be determined from sensitivity analyses within ASPEN Plus™ and the economic spreadsheet model. One option that might help costs is the addition of a steam turbine to produce electricity from the excess heat in the reforming operation. Also, a quench operation will be tested to cool the feed to the PSA unit. Furthermore, new information obtained in the testing and scale-up of the BCL gasifier will be incorporated into this analysis to measure cost improvement.

As biomass-based processes become better developed and the uncertainties associated with the cost of the biomass feedstock decrease, the assumptions made in this analysis will be revisited. Currently, the biomass feedstock cost is a result of the analysis, determined from the cost curves set between the current high and low market values of hydrogen.

A life cycle assessment will be conducted on these processes to determine their environmental impacts in terms of energy consumption and emissions to water and the air. This will include a comparative analysis of conventional hydrogen producing processes.

The economic and environmental effects of reforming a mixture of biomass syngas and natural gas should be studied. Because the stoichiometric maximum amount of hydrogen that can be produced from methane is higher than from syngas, the overall amount produced would be higher. This scenario would allow for higher hydrogen recovery by avoiding the need for the recycle stream used to increase the percentage of hydrogen in the PSA feed, while maintaining or sacrificing only a small portion of the benefits obtained in using a biomass-based process. This may also be a means of transition from the current hydrogen production technologies to a renewable technology.

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