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Office of Energy Efficiency & Renewable Energy

Analyzing the Levelized Cost of Centralized and Distributed Hydrogen Production Using the H2A Production Model, Version 2

Technical Report
NREL/TP-560-46267
September 2009

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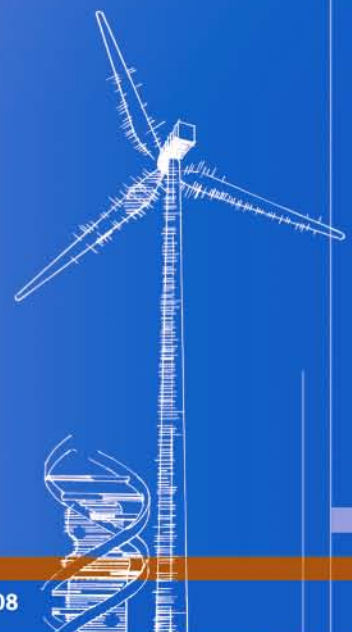

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Executive Summary

The U.S. Department of Energy's (DOE) Hydrogen, Fuel Cells, and Infrastructure Technologies Program is exploring the necessary hydrogen production, delivery, storage, and fuel cell technologies needed to commercialize fuel cell vehicles and hydrogen fuel infrastructure. A key component of the DOE Hydrogen Program's R&D effort is investigation of the potential costs of these hydrogen technologies, including the cost of hydrogen production.

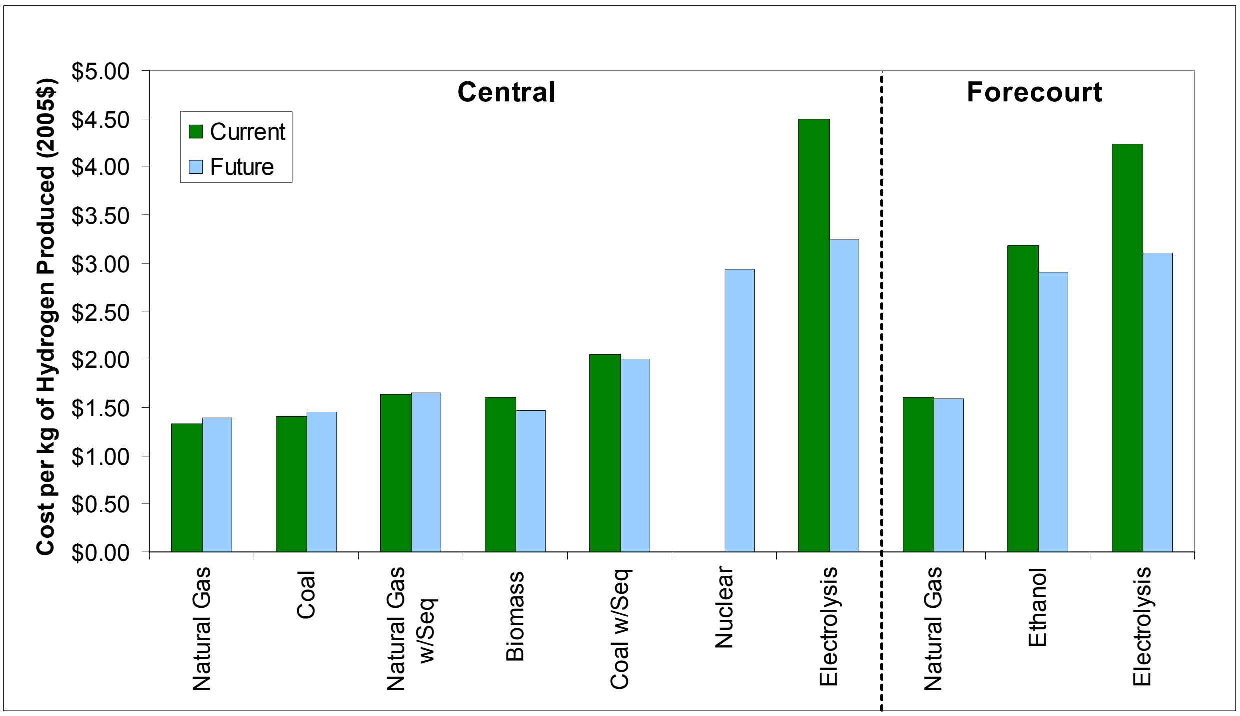
This report presents an analysis of the levelized cost of producing hydrogen via different pathways using the National Renewable Energy Laboratory's (NREL) H2A Hydrogen Production Model, Version 2. NREL developed the H2A Production Model in 2002 to help DOE's Hydrogen Program understand the cost of producing hydrogen for the transportation market. The model enables technical and economic analysis of central and distributed (i.e., at the fueling station, or "forecourt") hydrogen production technologies. Using a standard discounted cash flow rate of return methodology, it determines the levelized hydrogen cost, including a specified after-tax internal rate of return on investments. The cost calculation is based on a wide variety of inputs that characterize financial assumptions as well as capital, operating, maintenance, feedstock, utility, and replacement costs. The recently released Version 2 of the model features enhanced usability and functionality. Input fields are consolidated and simplified. New capabilities include performing sensitivity analyses and scaling analyses to various plant sizes.

To enhance long-term energy security, DOE envisions that hydrogen will be produced from various energy sources using a range of processing methods, including large-scale centralized and small-scale distributed production. Although the cost of producing hydrogen from these different pathways will vary from region to region based on feedstock availability and cost, a consistent analysis methodology is important for understanding and comparing costs associated with each pathway. With input from a team of hydrogen experts, NREL developed 19 standardized H2A production technology cases, enabling analysis of levelized hydrogen production costs from centralized and distributed facilities using eight general production pathways:

- Central biomass gasification
- Central grid electrolysis
- Central coal gasification (with and without carbon sequestration)
- Central natural gas reforming (with and without carbon sequestration)
- Central nuclear-based high-temperature electrolysis
- Distributed natural gas reforming
- Distributed grid electrolysis
- Distributed ethanol reforming.

For each pathway, currently available and expected future technologies were analyzed (except for the central nuclear case, which only includes a future case). Figure ES - 1 shows the calculated hydrogen production cost (per kilogram of hydrogen produced) for each technology case and timeframe. The values represent the levelized production cost of hydrogen (in 2005$), including a 10% after-tax internal rate of return on investments (this cost can also be thought of as the minimum selling price of hydrogen, although the actual hydrogen price will be driven by the market).

The levelized hydrogen production cost ranges from approximately $1.30 to $4.50 per kilogram, with the lowest production costs associated with traditional fossil fuel pathways such as coal and natural gas. However, once the costs associated with carbon capture and sequestration are included, production of hydrogen from renewable biomass resources becomes one of the most cost-competitive pathways.

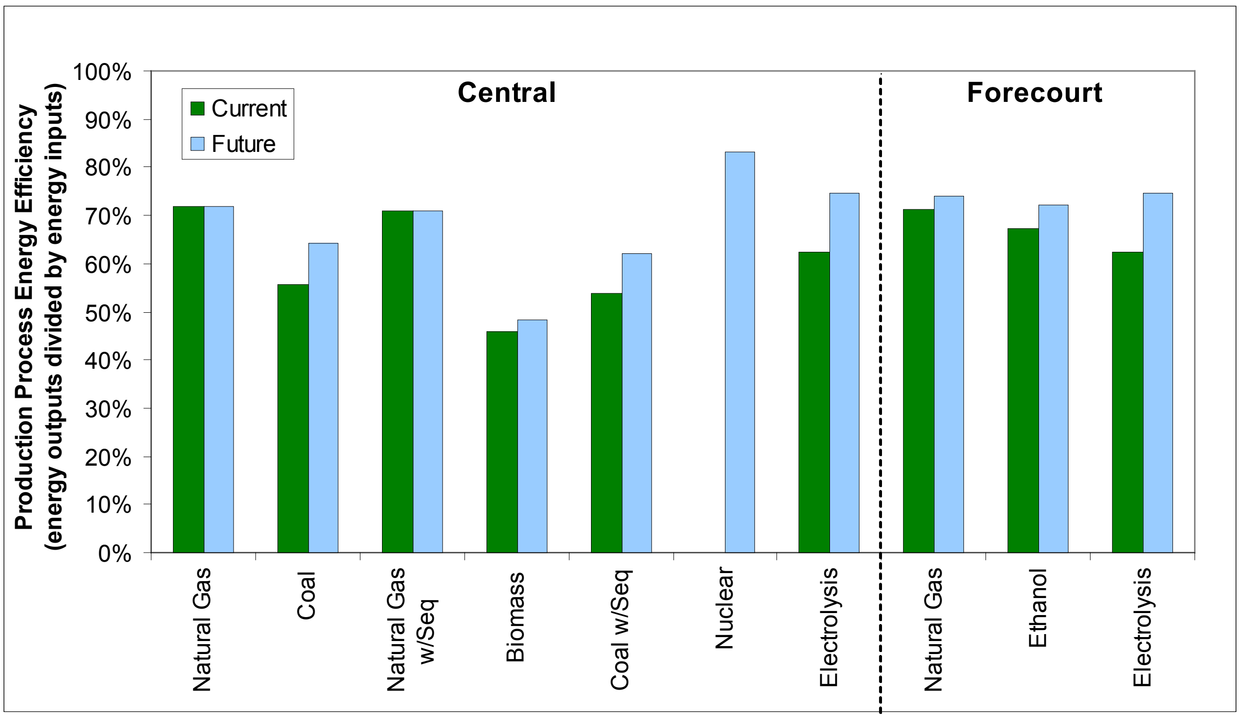


Costs are for hydrogen production and carbon sequestration only; delivery, compression, storage, and dispensing costs are not included.

Figure ES - 1. Levelized Hydrogen Production Cost: All Technology Cases.

DOE's goal is to reduce the untaxed levelized cost of hydrogen to $2.00 to $3.00 per gasoline gallon equivalent (GGE) delivered at the pump, regardless of the technology pathway used. (A kilogram of hydrogen has about the same energy content as a gallon of gasoline, i.e., it is about 1 GGE.) The H2A Production Model does not address delivery and dispensing costs for central cases, but it does analyze levelized hydrogen compression, storage, and dispensing (CSD) costs for forecourt cases because these functions will be included at the same fueling station site as the hydrogen production operations. Overall, CSD costs are expected to total about $1.90 per kilogram of hydrogen dispensed. Given these costs, which are not included in Figure ES - 1, the total levelized cost of dispensed hydrogen for the forecourt pathways ranges from about $3.50 per kilogram for onsite natural gas reformation to about $6.00 per kilogram for onsite electrolysis ($5.00/kg in the future). This cost might be improved in the future as better CSD technologies are developed and modeled.

This report also presents analyses of process energy efficiency—the ratio of process energy output to input—for the various hydrogen production pathways. Process energy efficiency indicates the quantity of resources needed to produce hydrogen and is therefore an important indicator of the efficient use of domestic resources. The future nuclear case has the highest process energy efficiency at 83%; the current biomass case has the lowest at 46% (Figure ES - 2).



Energy inputs and outputs are for hydrogen production and carbon sequestration only; compression, storage, and dispensing energy is not included. All values were calculated using LHVs.

Figure ES - 2. Process Energy Efficiency: All Technology Cases

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Introduction

In 2003, President Bush announced the Hydrogen Fuel Initiative to promote the development of hydrogen and fuel cell technologies. The Initiative's goal is to reduce U.S. dependence on foreign oil while addressing the environmental impacts of the transportation sector by reducing air pollution and greenhouse gas emissions associated with climate change. The U.S. Department of Energy's (DOE) Hydrogen, Fuel Cells, and Infrastructure Technologies Program is helping develop the hydrogen production, delivery, storage, and fuel cell technologies needed to commercialize fuel cell vehicles and hydrogen fuel infrastructure. A key component of the Hydrogen Program's R&D effort is investigation of the potential costs of these technologies, including the cost of hydrogen production.

This report presents an analysis of the levelized cost of producing hydrogen via different pathways using the National Renewable Energy Laboratory's (NREL) H2A Hydrogen Production Model, Version 2. NREL developed the H2A Production Model in 2002 to help the Hydrogen Program understand the cost of producing hydrogen for the transportation market. NREL recently completed a significant revision to the model. At the same time, with the input of a team of hydrogen experts, NREL developed 19 standardized hydrogen production technology cases, analyzing hydrogen production from centralized and distributed facilities.

The first section of this report provides background on the H2A Production Model's purpose and functions. This is followed by a summary of modeling results for the 19 H2A production technology cases (page 6). The remainder of the report includes input parameters and processes common to all H2A technology cases (page 16) as well as unique inputs and results for specific central (page 19) and forecourt (page 64) technology cases.

H2A Hydrogen Production Model

To enhance long-term energy security, DOE envisions hydrogen produced from various energy sources using a range of processing methods, including large-scale centralized and small-scale distributed production. Although the actual cost of producing hydrogen from these pathways will vary from region to region based on feedstock availability and cost, it is important to better understand current and potential costs of the range of hydrogen production technologies. The H2A Production Model is an analytical tool that provides a consistent analysis methodology and allows transparent understanding and reporting of hydrogen production costs.

The H2A Production Model enables the technical and economic analysis of central and distributed (i.e., fueling station or "forecourt") hydrogen production technologies. Using a standard discounted cash flow rate of return methodology[1], it determines the levelized hydrogen cost (minimum selling price) to achieve a net present value (NPV) of zero, including a specified after-tax internal rate of return on investments. The cost calculation is based on a wide variety of inputs that characterize financial assumptions as well as capital, operating, maintenance, feedstock, utility, and replacement costs. Model users can provide their own inputs or can begin with any of the 19 standard technology case studies developed in conjunction with the H2A

[1] Short, W.; Packey, D. J.; Holt, T, *A Manual for the Economic Evaluation of Energy Efficiency and Renewable Energy Technologies.* NREL/TP-462-5173. Golden, CO: National Renewable Energy Laboratory, March 1995.

model. With these technology cases, users can accept default technology input values or enter their own custom values. Users can also modify the model's financial inputs.

Version 2 of the H2A Production Model features enhanced usability and functionality. Input fields are consolidated and simplified. New capabilities include performing sensitivity analyses and scaling analyses to various plant sizes. Table 1 lists highlights of Version 2. Visit www.hydrogen.energy.gov/h2a_production.html to download Version 2 and its user guide.

Table 1. Highlights of H2A Production Model, Version 2

Enhanced usability
- Clarified definitions, assumptions, and default values
- Consolidated and simplified input fields

New functionality
- Flexible analysis capability
- Sensitivity analyses and tornado charts
- Easy plant scaling
- Upstream energy use and greenhouse gas emissions estimates (based on GREET Model[2])
- Process energy efficiency and greenhouse gas emissions estimates
- Carbon tax calculations based on feedstock carbon content and upstream energy use, credits for carbon sequestration
- Forecourt hydrogen compression, storage, and dispensing modeling (based on H2A Delivery Analysis Scenario and Carrier Components Models[3])
- Central CO_2 compression, transportation, and injection calculations
- Easy importing, exporting, and printing of data and results

New and updated technology cases
- Central production of hydrogen
  - Biomass gasification
  - Grid-based electrolysis
  - Coal gasification (with and without carbon sequestration)
  - Natural gas reforming (with and without carbon sequestration)
  - Nuclear—high-temperature electrolysis
- Distributed production of hydrogen
  - Electrolysis
  - Natural gas reforming
  - Ethanol reforming

H2A Production Model Functions

The H2A Production Model is actually two models: one Microsoft Excel workbook to analyze centralized hydrogen production technologies and another to analyze distributed/forecourt hydrogen production technologies for use at fueling stations. The two models are very similar. The primary difference is that the central model ends at the plant gate, i.e., it accounts for production but not distribution of hydrogen, whereas the forecourt model performs fueling

[2] Greenhouse gases, Regulated Emissions, and Energy use in Transportation (GREET) Model, www.transportation.anl.gov/software/GREET.
[3] H2A Delivery Scenario Analysis and Delivery Carrier Components Models, www.hydrogen.energy.gov/h2a_delivery.html.

station compression, storage, and dispensing calculations adapted from the H2A Delivery Scenario Analysis and Delivery Carrier Components models. The central model performs carbon sequestration calculations.

The model workbooks are organized into 19 or more worksheets, which have tabs color coded according to their function. Figure 1 shows the data flow among worksheets. In general, users can select from the slate of standard production cases or can input their simulated hydrogen production facility's technical operating parameters and specifications; financial input values; energy feedstocks, utilities, and byproducts; capital costs; and fixed and variable operating costs. The model uses these inputs—along with price data and physical properties of process materials—to calculate projected levelized hydrogen cost and approximate process energy use and greenhouse gases emissions. For detailed information about the model, download its user guide at www.hydrogen.energy.gov/h2a_production.html.

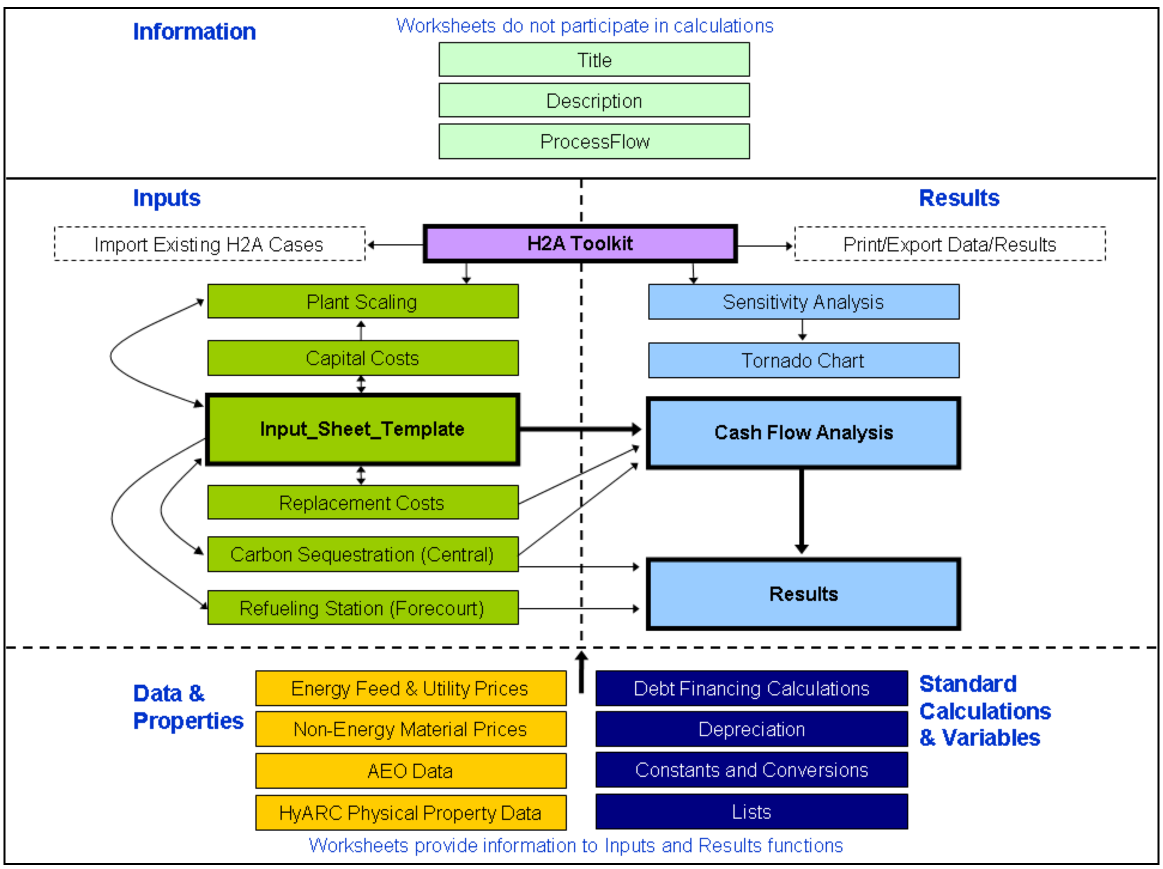


Figure 1. Schematic of Data Flow Among H2A Worksheets

H2A Production Technology Cases

As part of the H2A Production Model development, NREL worked with a team of "key industrial collaborators" from hydrogen-related industries as well as numerous government, academic, and consulting experts to model 19 specific hydrogen production pathways reflecting eight general production methods (Table 2). Input parameters were defined by experts in the design and advancement of technologies for each production method. Seven of the production methods were modeled based on current technology (assumed to be available as of 2005, the year in which the technology cases were first published) and future technology (assumed to be available in the 2020–2030 timeframe). The central coal and natural gas pathways were also modeled with and without carbon capture and sequestration. Hydrogen production from nuclear energy was modeled as a future case only (with an assumed startup in 2030).

As the transportation market for hydrogen develops, it is likely that hydrogen initially will be produced in a distributed fashion at the fueling station. Distributed production will be accomplished via onsite reformation of natural gas and renewable ethanol and via onsite water electrolysis. Once the hydrogen market is more established, these distributed production facilities will be augmented with larger, centralized hydrogen production facilities that make hydrogen using fossil fuels (including coal and natural gas), nuclear power, and renewable energy sources (including biomass and wind-generated electricity). Hydrogen produced at these central facilities will be delivered to fueling sites initially using trucks and, later, pipelines. The H2A technology cases were developed to reflect this full range of production techniques.

Table 2. H2A Hydrogen Production Technology Cases

Central production technologies
Biomass
1. Current biomass gasification (version 2.1.2)
2. Future biomass gasification (version 2.1.2)
Electrolysis
3. Current grid electrolysis (version 2.1.1)
4. Future grid electrolysis (version 2.1.1)
Coal
5. Current coal without carbon sequestration (version 2.1.1)
6. Future coal without carbon sequestration (version 2.1.1)
7. Current coal with carbon sequestration (version 2.1.1)
8. Future coal with carbon sequestration (version 2.1.1)
Natural gas
9. Current natural gas without carbon sequestration (version 2.1.1)
10. Future natural gas without carbon sequestration (version 2.1.1)
11. Current natural gas with carbon sequestration (version 2.1.1)
12. Future natural gas with carbon sequestration (version 2.1.1)
Nuclear
13. Future nuclear via high-temperature electrolysis (version 2.1.1)
Forecourt production technologies
Electrolysis
14. Current grid electrolysis (version 2.1.2)
15. Future grid electrolysis (version 2.1.2)
Natural gas
16. Current natural gas (version 2.1.1)
17. Future natural gas (version 2.1.1)
Ethanol
18. Current ethanol (version 2.1.2)
19. Future ethanol (version 2.1.2)

Figure 2 shows the wide variety of hydrogen production design capacities among the various H2A cases. The central cases were modeled to produce at least 50,000 kg of hydrogen per day, but the design capacities vary depending on the technology modeled. The current central cases have hydrogen production capacities ranging from approximately 52,000 to 820,000 kg of hydrogen per day, with each plant design optimized for its specific production capacity. At the high end, the minimum practical size requirements of a nuclear facility dictate that the production plant be designed to produce a very large amount of hydrogen. At the low end, for central electrolysis, multiple electrolyzers will need to be installed to attain 52,000 kg/day of production; the modular nature of the design could easily allow higher production rates, although the per-kilogram hydrogen cost is not expected to be lower. Biomass-based production facilities were modeled with a capacity of approximately 150,000 kg/day. This plant size is driven not by the biomass gasification technology but rather by the distance from which biomass feedstock can be delivered economically.

All the forecourt cases model facilities that produce 1,500 kg of hydrogen per day. In addition to their advantage of eliminating the need for an extensive hydrogen delivery infrastructure, distributed production facilities can be sized to match the hydrogen demand of the local area. To facilitate easy comparisons across technology pathways, the standard H2A cases are all based on a 1,500-kg/day production capacity.

The technology cases include not only the designed production capacity, discussed above, but also the potential plant output. Because of plant outages and scheduled maintenance, actual plant output will be less than design capacity. The default capacity factor—which is multiplied by the design capacity to calculate plant output—is 90% for central cases (with some exceptions) and about 85% for forecourt cases.

The capacity factor for the forecourt stations was calculated based on expected seasonal variations in demand and planned and unplanned hydrogen production equipment outages. The following calculations were used to derive the value (Source: Directed Technologies, Inc.).

$$CF = 100\% - R_{season} - R_{planned} - R_{unplanned} - R_{extra}$$

Where:

CF =	Operating capacity factor
R_{season} =	CF reduction for seasonal loads (winter to summer) Assumption = 10%
$R_{planned}$ =	CF reduction for planned shutdown 7 days per year for planned system shutdown (annual maintenance, etc.) $R_{planned}$ = 7 day/year ÷ 365 day/year = 1.92%
$R_{unplanned}$ =	CF reduction for unplanned shutdowns 6 "expected" unplanned system shutdowns per year (equipment failure, power outage, etc.) 14 hr system down for each unplanned shutdown (average): 2 hr to react to shutdown (also allows unit to cool) 6 hr to get repair personnel to site 4 hr to effect repairs (assumes replacement parts are in hand) 2 hr to bring unit back to full power and monitor for proper performance

$R_{unplanned}$ = 6 shutdowns/year × (14 hr/shutdown ÷ 8,760 hr/year) = 0.96%

R_{extra} = CF reduction for needing extra production capacity to refill storage tanks after unplanned shutdowns

H2 storage for hourly/daily demand fluctuations determined by the Chevron supplied hourly demand load calculations in the HDSAM/H2A model

Chevron demand based on highest daily demand of highest weekly demand (Friday in summer)

1,500 kg/day maximum rating of forecourt production system

14 hr system down for each unplanned shutdown (average)

30 days of design time between unplanned shutdowns

R_{extra} = 14 hr/shutdown ÷ (30 days × 24 hr/day) = 1.94%

H2 storage for unplanned shutdowns = 14 hr/shutdown ÷ 24 hr/day × 1,500 kg/day = 875 kg

CF = 100% – 10% –1.92% –0.96% –1.94% = 85.2%

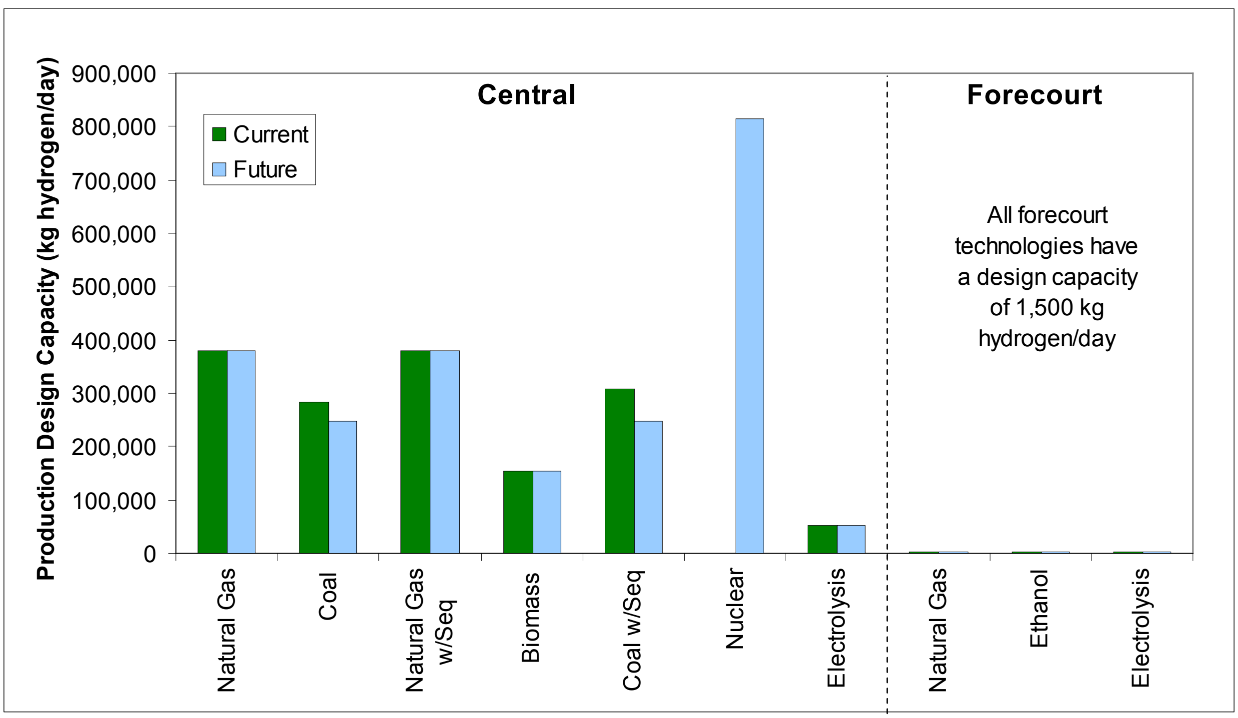


Figure 2. Hydrogen Production Design Capacity: All Technology Cases

Overview of Levelized Hydrogen Production Cost and Energy Use

Using the standard H2A production technology cases, the levelized costs of producing hydrogen from eight typical production pathways were investigated. The results of this cost analysis, as well as an analysis of the process energy required for the various production pathways, are presented below. The subsequent sections of this report describe the detailed inputs and results for each technology pathway. Visit www.hydrogen.energy.gov/h2a_prod_studies.html to download the cases.

Levelized Hydrogen Production Cost

DOE's goal is to reduce the untaxed levelized cost of hydrogen to $2.00 to $3.00 per gasoline gallon equivalent (GGE) delivered at the pump, regardless of the technology pathway used. (A kilogram of hydrogen has about the same energy content on a lower heating value [LHV] basis as a gallon of gasoline, thus a kilogram of hydrogen is approximately equal to 1 GGE.) The H2A Production Model and associated technology cases help researchers to understand the levelized hydrogen production cost component of this DOE goal. The central cases show only the production cost component, with costs reflecting produced hydrogen available "at the plant gate." The full cost of this centrally-produced hydrogen also needs to include the cost of hydrogen delivery and dispensing, which can be determined using DOE's Hydrogen Delivery Scenario Analysis Model.[4] The forecourt production cases include levelized hydrogen production costs by component and also provide the levelized cost of hydrogen compression, storage, and dispensing (CSD) because these functions will be included at the same fueling station site as the hydrogen production operations.

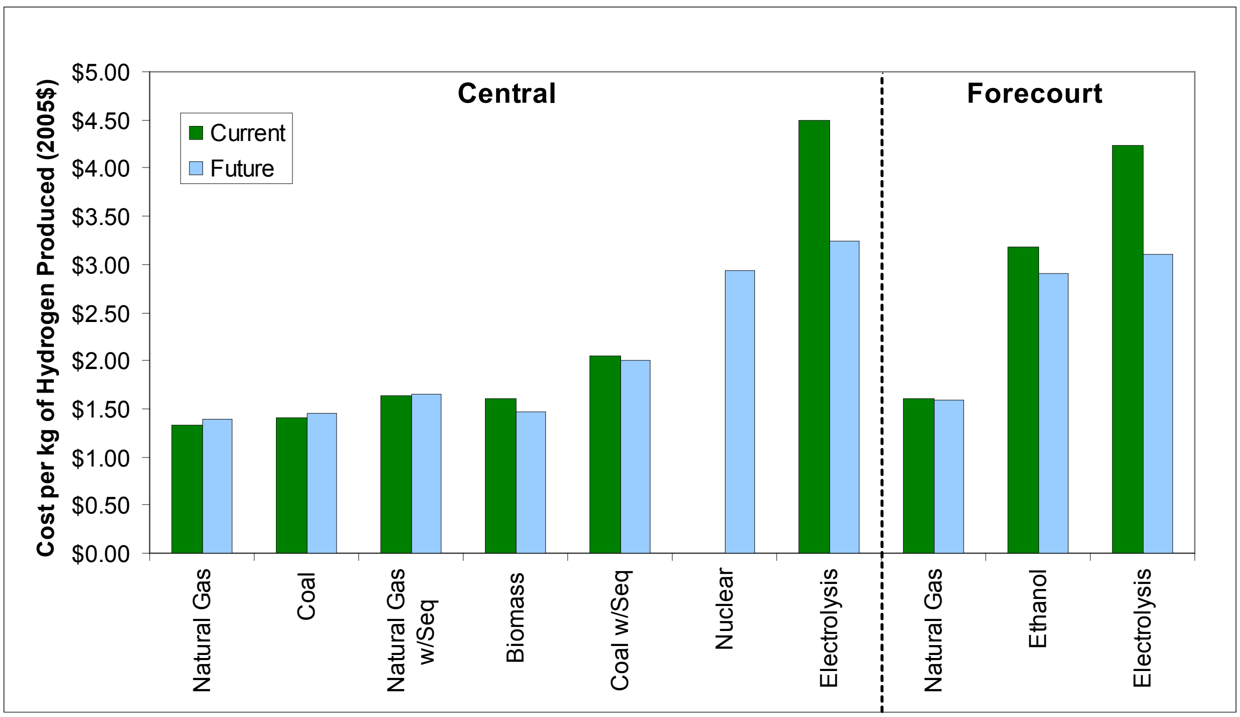
Figure 3 shows the calculated cost per kilogram of hydrogen produced for each technology case and timeframe. The values represent the levelized cost of hydrogen (in 2005$), including a 10% after-tax internal rate of return (this cost can also be thought of as the minimum hydrogen selling price, although the market will drive the actual hydrogen price). Although the model can accommodate a wide variety of financial strategies and assumptions, a standard set of financial assumptions was used for the cases to maintain consistency. Cost values are shown in 2005$ for all cases. To allow a consistent comparison of hydrogen production costs from central and forecourt technologies, CSD costs are not included in this figure; only production costs and carbon sequestration costs (for the two relevant central cases) are included. (See Figure 7 for the CSD costs of the three forecourt technology cases.)

The levelized hydrogen production cost ranges from approximately $1.30 to $4.50 per kg, with the lowest production costs associated with traditional fossil fuel pathways such as coal and natural gas. However, once the costs associated with carbon capture and sequestration are included, hydrogen production from renewable biomass resources becomes one of the most cost-competitive pathways. For most cases, the cost difference between current and future timeframes varies only slightly. The central and forecourt electrolysis cases realize a substantial price decrease in the future timeframe, primarily owing to a large expected decrease in the capital cost of the electrolyzer system. Overall, the costs of water electrolysis production are higher than other pathways. These costs are based on average U.S. electricity prices. Electrolysis production might be more cost competitive in regions with below-average electricity prices.

For most pathways, capital and feedstock costs are the dominant contributors to total cost, each typically contributing one-third to one-half—and up to more than three-quarters—of the total levelized production cost. Figure 4 and Figure 5 show the contribution of various cost components to hydrogen production cost. Feedstock costs are based primarily on energy cost

[4] H2A Delivery Scenario Analysis Model, www.hydrogen.energy.gov/h2a_delivery.html.

projections from the Energy Information Administration's Annual Energy Outlook (AEO).[5] To maintain consistency among cases, the AEO price projections for the 2005 "High A Case" were used for most analyses. The total feedstock cost for the production pathway also depends on the amount of process energy used and the production process efficiency, which are discussed below. Estimates of total capital costs were developed by the H2A team as part of developing the H2A production technology cases. In general, the production pathways investigated represent the use of pioneering technologies. To better represent the expected costs associated with capital investment in a commercialized market, the projected capital costs for central facilities were based on the "nth plant," where the technology has been proven on a large scale, the first-of-its-kind cost penalties have been overcome, and costs have stabilized. Similarly, capital cost projections for forecourt technologies are based on a production level of 500 units per year.



Costs are for hydrogen production and carbon sequestration only; compression, storage, and dispensing (CSD) costs are not included. Figure 7 shows CSD costs for forecourt technology cases. The H2A Production Model, Version 2.0, calculates CSD costs for forecourt technology cases only; however, hydrogen produced from central technologies also would incur CSD costs in order for the hydrogen to be usable in vehicles.

Figure 3. Levelized Hydrogen Production Cost: All Technology Cases

Beyond capital and feedstock costs, operating and maintenance costs, raw material costs, and other variable costs can play a significant role. Sequestration costs can be significant for the natural gas and coal cases that include carbon capture and sequestration. The costs of carbon sequestration for the coal pathway in particular are large because of coal's higher CO_2 emissions factor relative to natural gas. The total cost of hydrogen can be reduced as a result of revenue generated from the sale of material or energy byproducts of the hydrogen production process. For example, several coal pathways produce excess electricity, which is sold to generate revenue.

[5] Energy Information Administration Annual Energy Outlook, http://www.eia.doe.gov/oiaf/aeo/aeoref_tab.html.

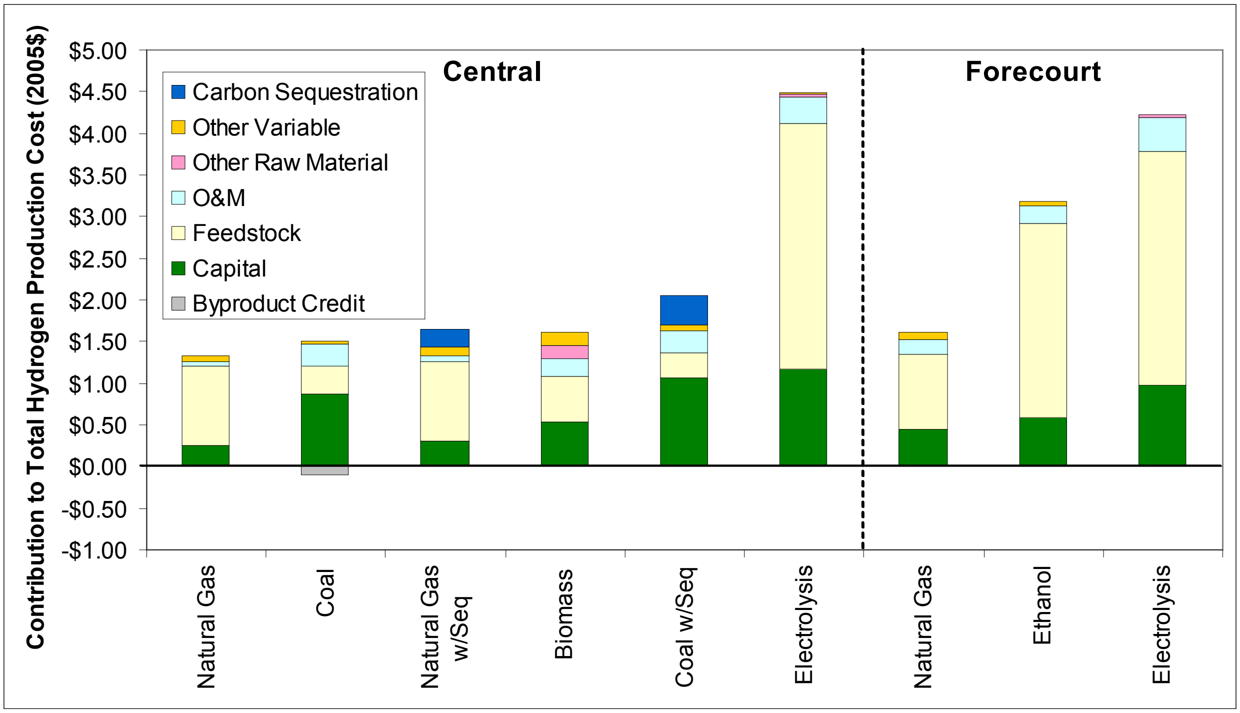
This is represented in the figure by the gray bar segments below the $0 line (the total hydrogen cost equals the segments above the $0 line minus the segment below the $0 line). Sale of byproducts other than electricity (e.g., oxygen from electrolysis) was not considered for these standardized cases.

To better understand the impact of capital investment on production cost, the amount of capital investment required can be compared to facility design capacity. Figure 6 shows the relative capital intensity of the technology cases, dividing total initial direct capital cost by hydrogen production design capacity. The coal-based technology cases are more capital intensive than most other cases; this is offset by the low cost of coal feedstock and the credit for byproduct electricity generated (see Figure 4 and Figure 5). Electrolysis is relatively capital intensive in the current timeframe, with a large reduction in capital cost in the future timeframe (owing to the lower-cost electrolyzer systems in the future).

The forecourt cases include the levelized cost of hydrogen production as well as the CSD costs at the fueling site, so these cases provide the full cost of delivered hydrogen, which can be compared to the DOE goal of $2.00 to $3.00 per GGE dispensed to the consumer (Figure 7). CSD costs constitute approximately one-third to one-half of total delivered hydrogen costs, about $1.90 per kilogram of hydrogen dispensed. Because the forecourt cases have identical hydrogen production capacities and fueling station characteristics, their CSD costs are almost identical. The only difference is the lower electricity cost for electrolysis. It was assumed that, because of their high electricity consumption, these facilities would be able to purchase electricity at a lower industrial rate.

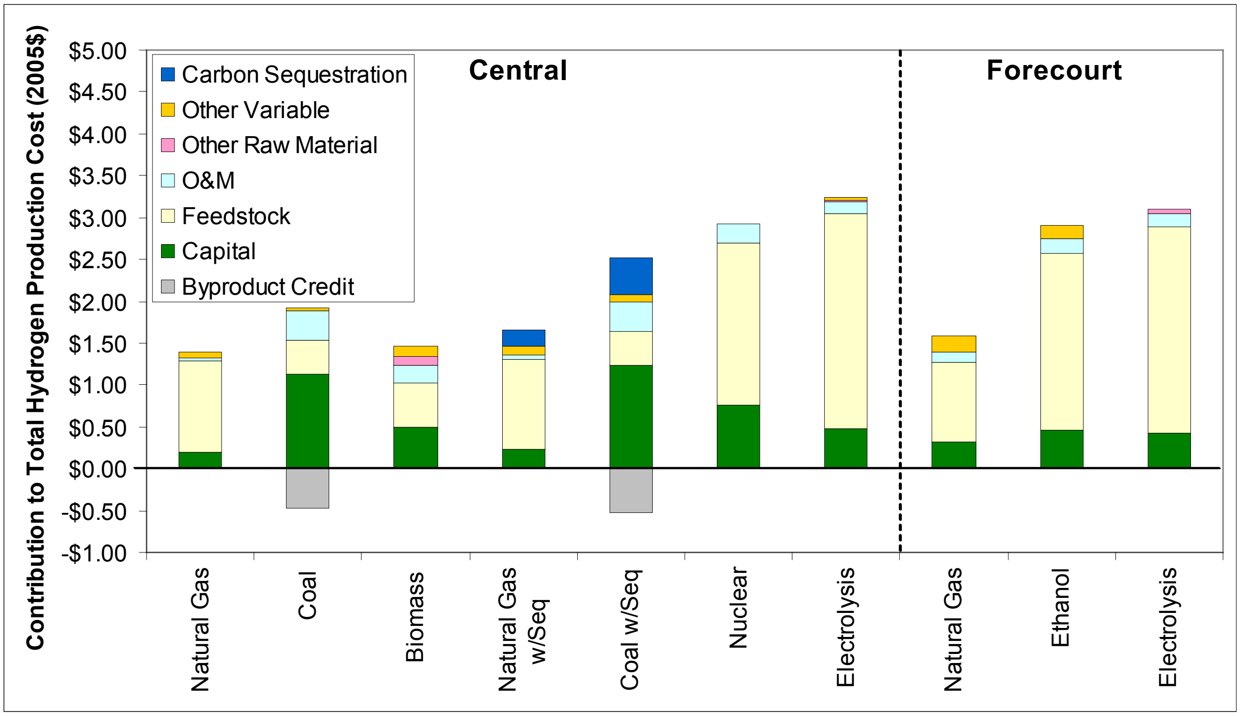
DOE expects forecourt hydrogen production at fueling stations to meet hydrogen demand during the early transition to hydrogen. As seen in Figure 7, the total levelized cost of dispensed hydrogen is expected to be about $3.50 per kilogram for onsite natural gas reformation (the lowest-cost pathway), which exceeds DOE's target price.[6] This total dispensed cost of onsite hydrogen might be improved in the future as better CSD technologies are developed and modeled. Currently, the H2A Model does not capture improved CSD technologies and lower CSD costs for future applications; thus, modeled current and future CSD costs are the same.

[6] Also see the 2006 systems integration independent review *Distributed Hydrogen Production from Natural Gas*, www.hydrogen.energy.gov/pdfs/40382.pdf, which used the H2A Production Model to project that the cost of distributed hydrogen production from natural gas at high volumes could meet the upper range of DOE's $2.00 to $3.00 per GGE cost target.



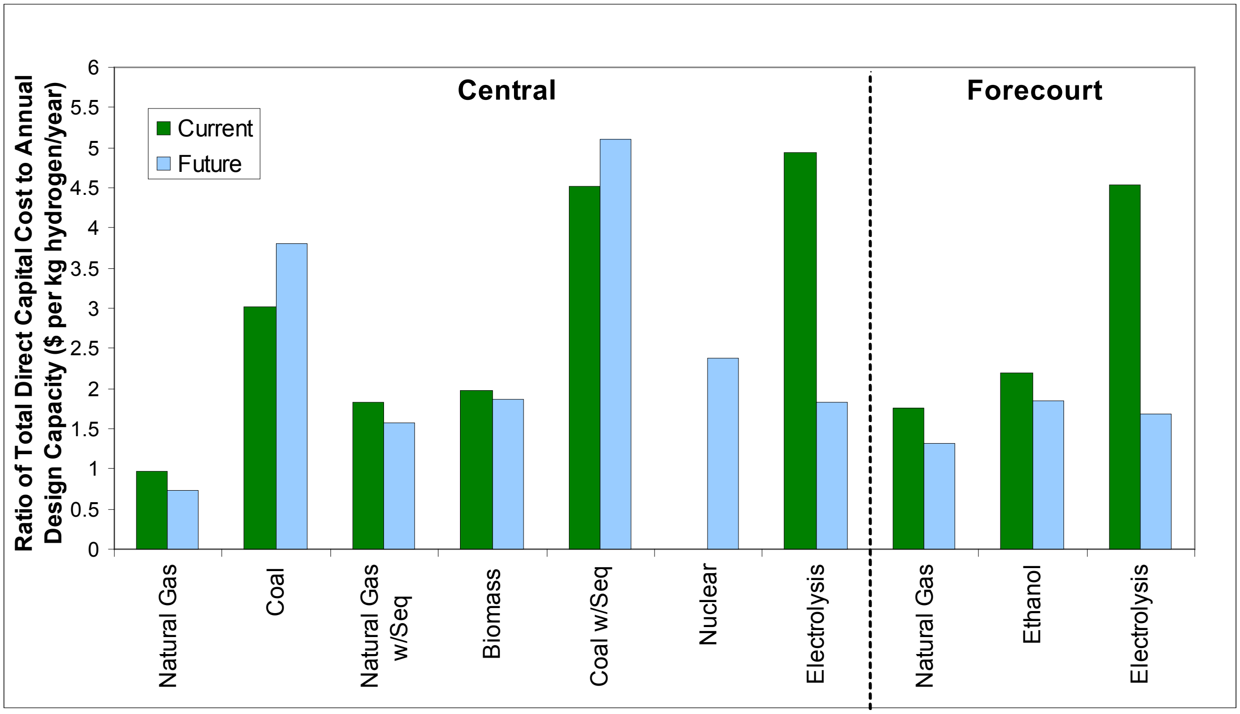
Costs are for hydrogen production and carbon sequestration only; compression, storage, and dispensing costs are not included. The gray segment below the $0 line for coal represents a credit for byproduct electricity; the total cost of hydrogen from this technology equals the colored segments above the $0 line minus the gray segment below the $0 line.

Figure 4. Contributions to Levelized Hydrogen Production Cost: Current Technology Cases



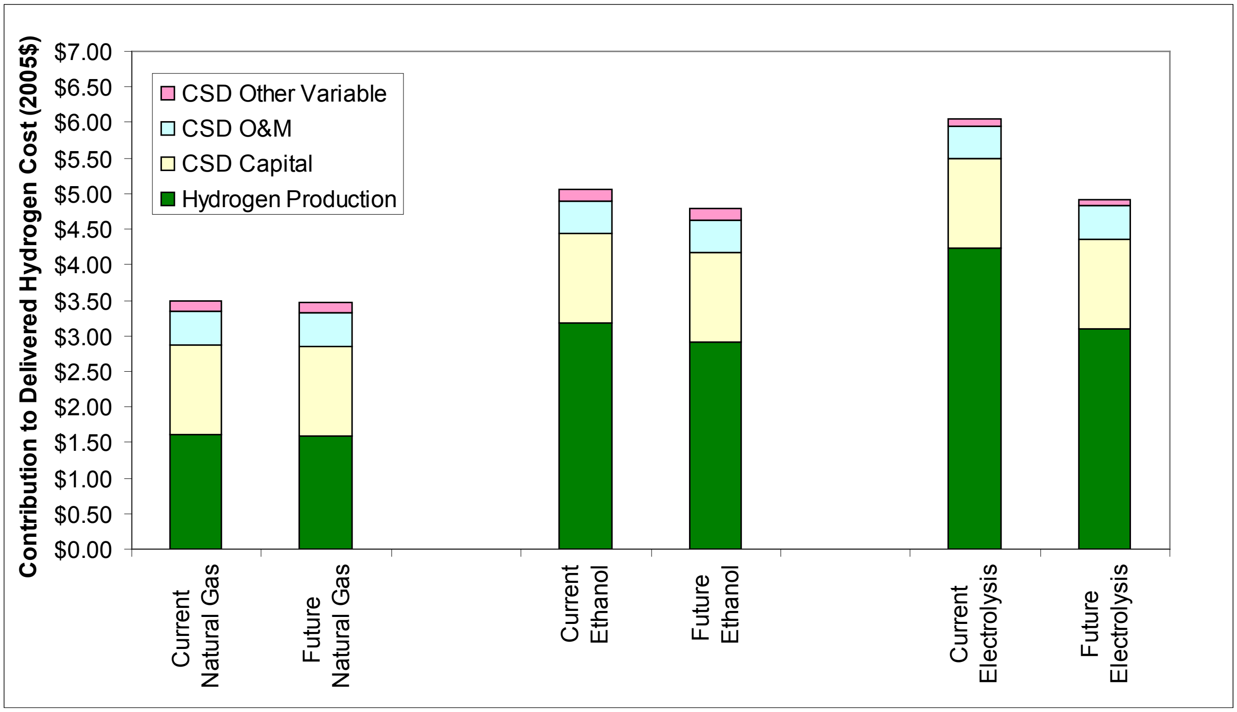
Costs are for hydrogen production and carbon sequestration only; compression, storage, and dispensing costs are not included. The gray segments below the $0 line for coal and coal with sequestration represent credits for byproduct electricity; the total cost of hydrogen from these technologies equals the colored segments above the $0 line minus the gray segment below the $0 line.

Figure 5. Contributions to Levelized Hydrogen Production Cost: Future Technology Cases



These are initial direct capital costs for hydrogen production and carbon sequestration only; compression, storage, and dispensing capital costs are not included, nor are replacement capital costs incurred over the production plants' lifetimes.

Figure 6. Capital Investment Relative to Production Design Capacity: All Technology Cases



This figure includes refueling station compression, storage, and dispensing (CSD) costs—unlike Figure 3, Figure 4, and Figure 5, which show hydrogen production costs only. The H2A Production Model, Version 2, calculates CSD costs for forecourt technology cases only; however, hydrogen produced from central technologies also would incur both delivery and CSD costs in order for the hydrogen to be made available for use by hydrogen fuel-cell vehicles.

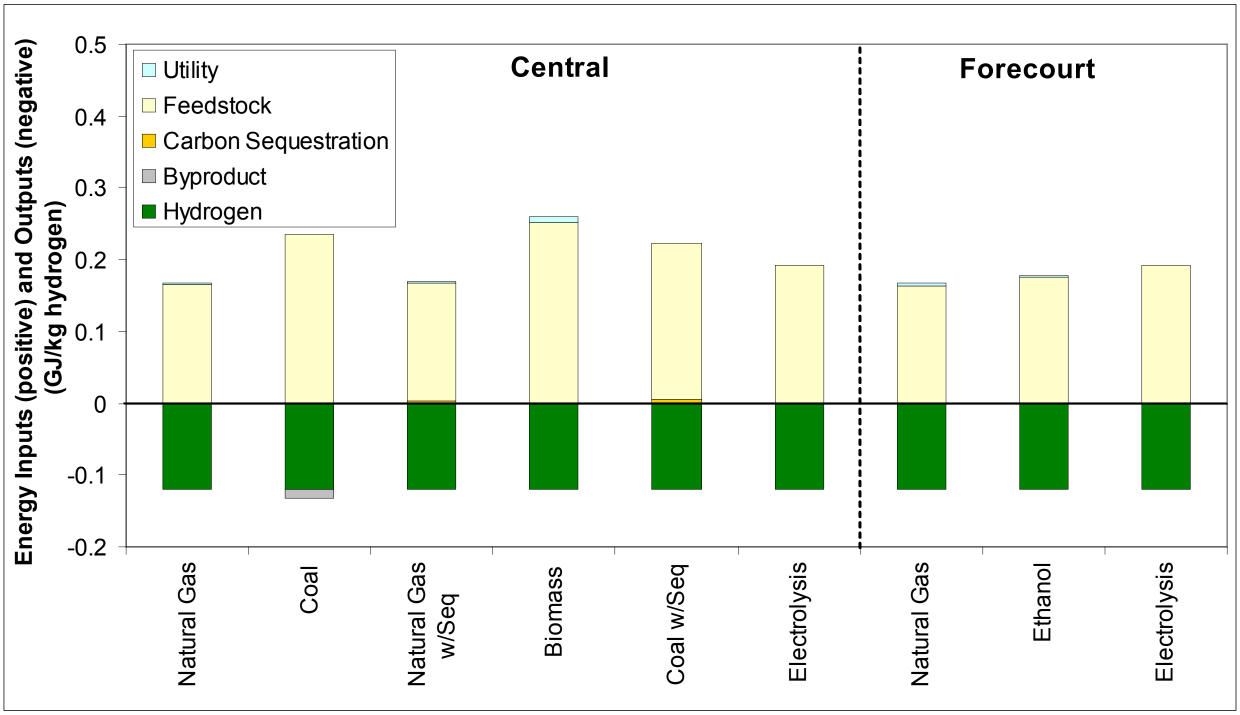
Figure 7. Contributions to Delivered Levelized Hydrogen Cost: Forecourt Technology Cases

Process Energy

Hydrogen can be derived from a variety of domestic energy resources, including fossil, nuclear, and renewable sources. Process energy efficiency—the ratio of process energy output to input—indicates the quantity of resources needed to produce hydrogen and is therefore an important indicator of the efficient use of domestic resources.

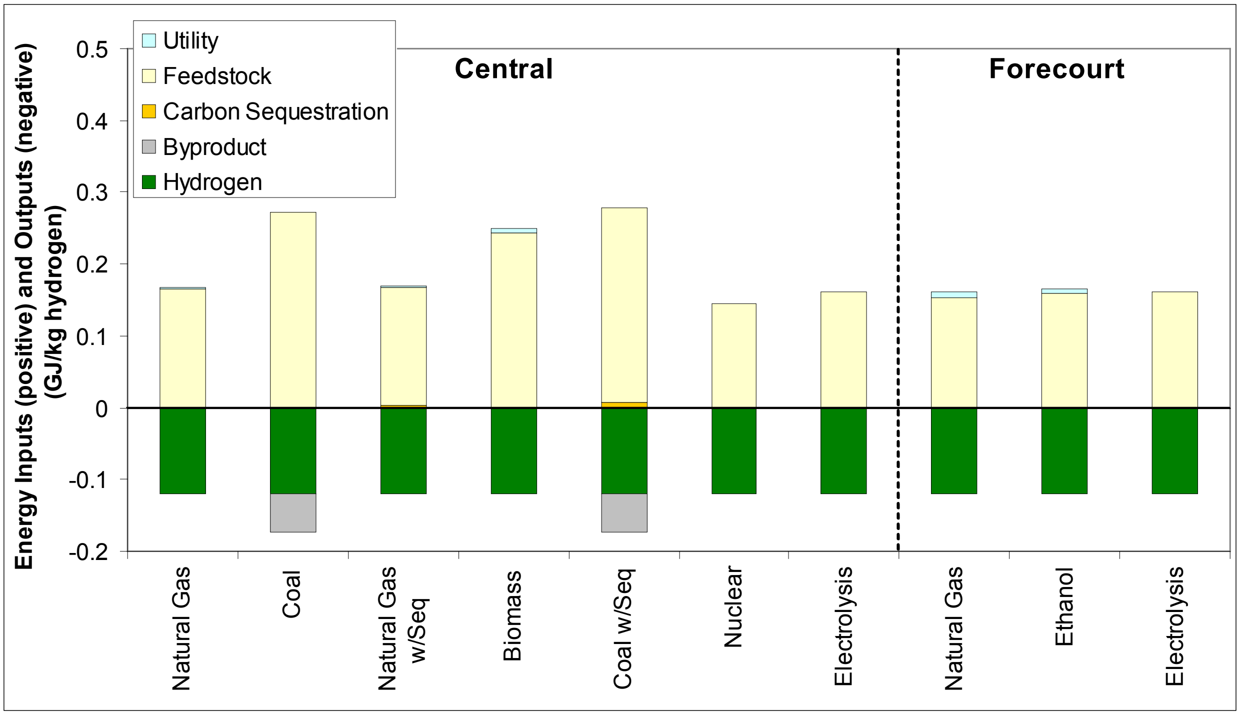
Figure 8 and Figure 9 show the process energy requirements for the different production pathways. The figures show the amount of input energy required (in the form of the energy contained in the feedstocks and the process energy used in the transformation) to produce one kilogram of hydrogen, which is shown as an output. For all cases, feedstocks constitute the vast majority of input energy. Byproduct energy flows are shown separately from the hydrogen output energy. The coal (current and future) and coal with carbon sequestration (future) cases produce excess electricity, which is another energy output for these cases. Energy used for carbon sequestration is identified separately from energy used for hydrogen production. The energy inputs shown do not include energy used upstream of the process—e.g., to extract fossil fuels and refine or produce feedstocks—or energy used for hydrogen delivery, compression, storage, and dispensing; thus they do not represent life-cycle or "well-to-wheel" energy use.

The future nuclear case has the highest ratio of process energy output to input (i.e., the highest process energy efficiency) at 83%. The current biomass case has the lowest process energy efficiency at 46%. Figure 10 shows the process energy efficiencies of all cases.



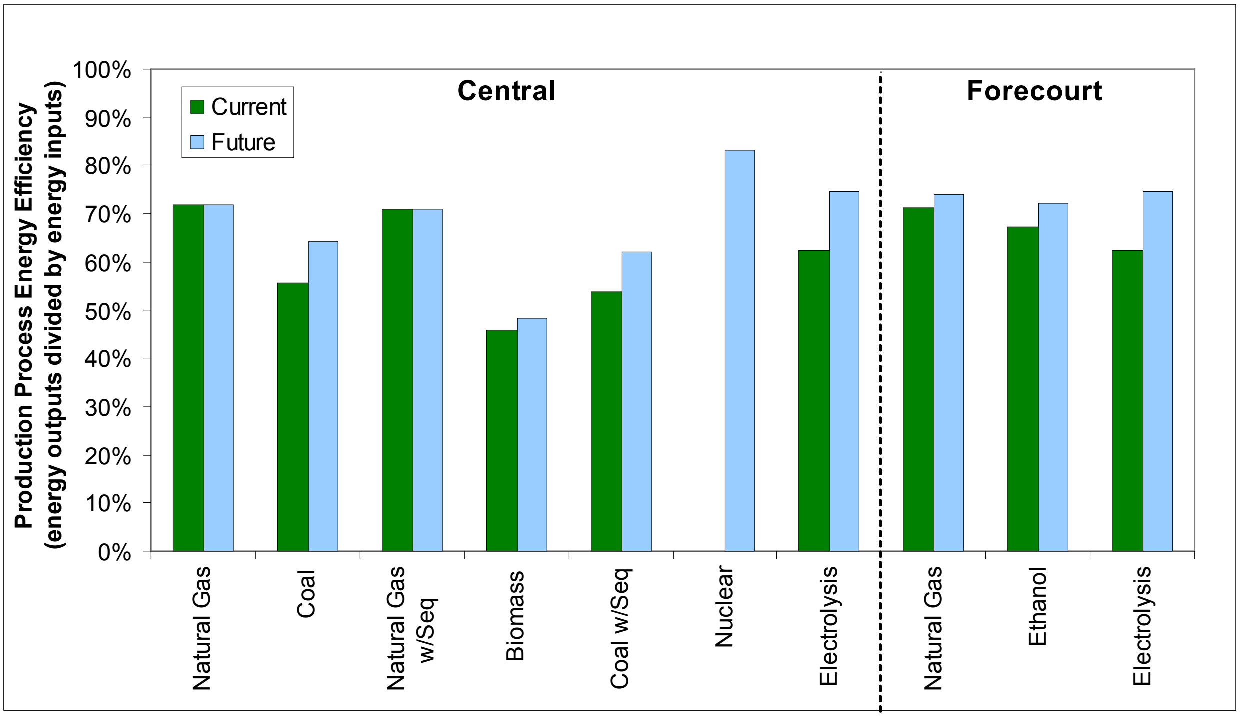
Energy inputs and outputs are for hydrogen production and carbon sequestration only; compression, storage, and dispensing energy is not included. The segments below the zero line represent the energy contained in the produced hydrogen and the energy in byproduct electricity. Net energy equals the segments below the zero line minus the segments above the zero line. All values were calculated using LHVs.

Figure 8. Process Energy Inputs and Outputs: Current Technology Cases



Energy inputs and outputs are for hydrogen production and carbon sequestration only; delivery, compression, storage, and dispensing energy are not included. The segments below the zero line represent the energy contained in the produced hydrogen and the energy in byproduct electricity. Net energy equals the segments below the zero line minus the segments above the zero line. All values were calculated using LHVs.

Figure 9. Process Energy Inputs and Outputs: Future Technology Cases



Energy inputs and outputs are for hydrogen production and carbon sequestration only; compression, storage, and dispensing energy is not included. All values were calculated using LHVs.

Figure 10. Process Energy Efficiency: All Technology Cases

Sensitivity Analysis

The H2A Production Model technology cases include sensitivity analyses to help users understand how the levelized hydrogen cost would change using different assumptions for key input parameters. For each case, key input variables were assigned three values: a likeliest value, a 10th-percentile value (i.e., the value at which 10% of the predicted inputs are at or below that value), and a 90th-percentile value (i.e., the value at which 90% of the predicted inputs are at or below that value). The variable values assigned were based on feedback from analysts consulted as part of the H2A development process and on ongoing DOE research into the uncertainties inherent to the various hydrogen production variables.

The likeliest, 10th-percentile, and 90th-percentile values for each single input variable were entered into the H2A Model while the other input variables were held constant at their likeliest values. The results of this analysis are ranges of levelized hydrogen production costs corresponding to the ranges of input variable values, which can be visualized in a "tornado chart." This indicates how much effect each input variable has on hydrogen cost when varied in isolation: the larger the range of hydrogen cost, the greater the effect of the variable. Figure 11 is an example tornado chart for the current coal without CO_2 sequestration case. In this example, total direct capital cost has the largest effect on hydrogen cost: a total direct capital cost of $290 million results in a hydrogen cost of about $1.35/kg, and a total direct capital cost of $430 million results in a hydrogen cost of about $1.75/kg.

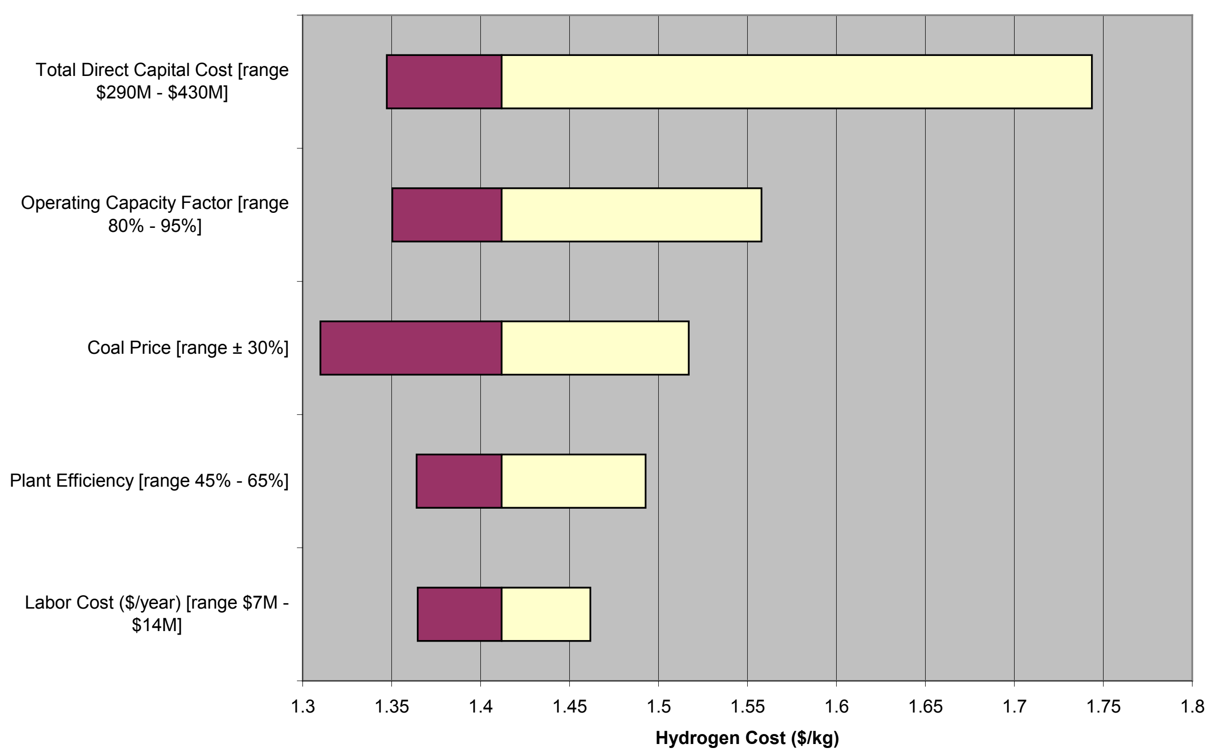
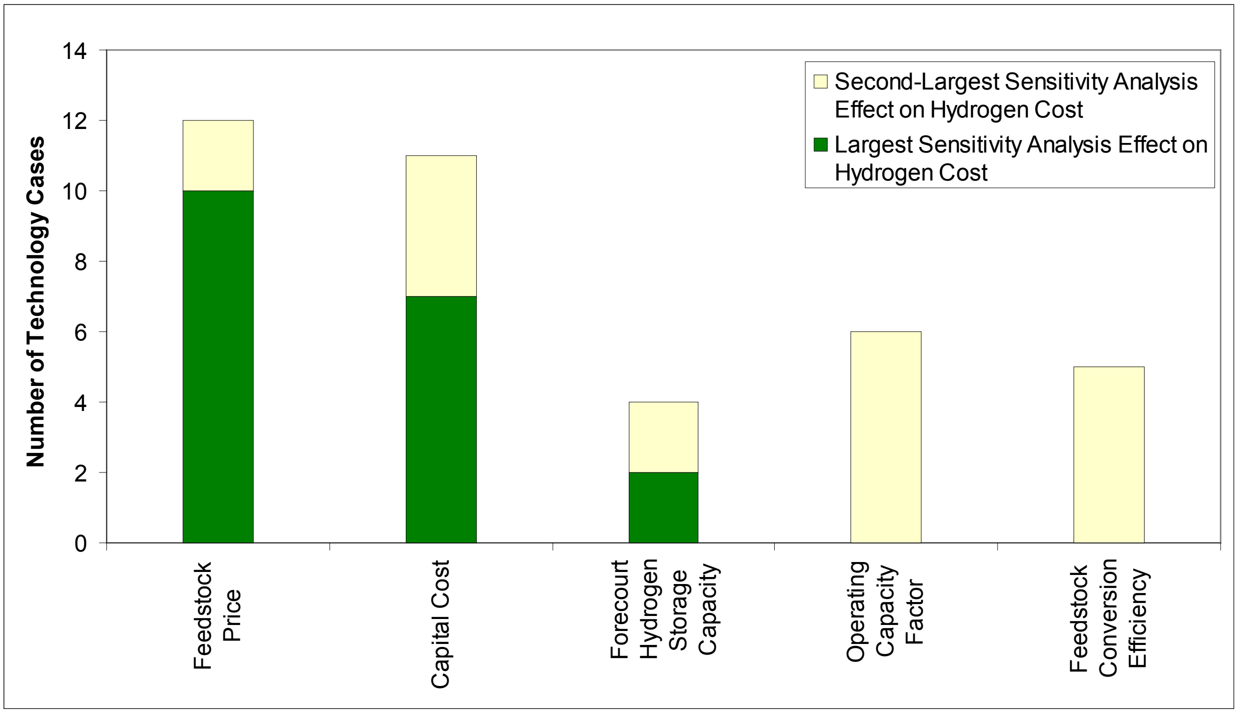


Figure 11. Example "Tornado Chart" (Current Coal without CO_2 Sequestration Case)

Figure 12 shows the input parameters that most frequently have the greatest effect on the modeled cost of hydrogen for all technology cases. Feedstock price has the greatest effect on hydrogen cost in 10 cases and the second-greatest effect in two cases. Capital costs have the greatest impact on hydrogen cost in seven cases and the second-greatest effect in four cases. For the forecourt cases, the amount of hydrogen storage has the greatest impact on hydrogen cost in two cases and the second-greatest impact in two cases. Operating capacity factor and feedstock conversion efficiency have the second-greatest impact in six and five cases, respectively.



Sensitivity analyses include hydrogen production variables for all cases; compression, storage, and dispensing variables for forecourt technology cases; and carbon sequestration variables for central cases.

Figure 12. Variables Producing the Largest Effects in Sensitivity Analyses

Detailed Levelized Hydrogen Production Cost Results

The results presented above provide an overview of the full range of hydrogen production cases developed using the H2A Production Model (Table 2 shows the cases studied). The sections below present additional details of these cases by production pathway, starting with the central cases and followed by the forecourt cases. Each case begins with a description of the hydrogen production process used, followed by a detailed accounting of the costs expected and the resulting levelized hydrogen cost. For full details, download the cases at www.hydrogen.energy.gov/h2a_prod_studies.html.

Common Input Parameters and Processes

The H2A technology cases have various default input parameters and processes, many of which are common to all or most of the cases. Table 3 lists these common values. Parameters and processes unique to each case are discussed in subsequent sections of this report.

Table 3. Common Input Parameters and Processes for Production Technology Cases

Analysis methodology	Discounted cash flow (DCF) model that calculates a levelized hydrogen cost yielding a prescribed IRR
Analysis period	40 years for central cases; 20 years for forecourt cases
Average burdened labor rate for staff	$50/hour for central cases; $10/hour for forecourt cases
Capacity factor	90% for central cases with exceptions; 85.2% for forecourt cases
Capital expenditure schedule—central	25%–75% of capital spent in first year of construction; 25%–75% in second year; 20%–30% in third year
Capital expenditure schedule—forecourt	100% of capital spent in first year of construction, with a 1-year total construction period.
Central storage	Optional buffer only as required for efficient operations
CO_2 capture credit	Not included in base cases (default value = 0)
CO_2 production taxes	Not included in base cases (default value = 0)
Construction period and cash flow	2–3 years for central cases; 1 year for forecourt cases
Co-produced and cogenerated electricity selling price	$30/MWh
Decommissioning	10% of initial capital for central cases, with exceptions; 0% for forecourt cases
Depreciation type and schedule for initial depreciable capital cost	MACRS: 20 years for central cases with exceptions; 5 years for compressors, 7 years for remainder of plant for forecourt cases
Energy and emissions data	From H2A Production Model *HyARC Physical Property Data* worksheet; various original sources
Energy feed and utility prices (standard)	Energy feed and utility prices vary over time based on values derived and extrapolated from AEO 2005 High A Case within H2A Production Model *Energy Feed & Utility Prices* worksheet; sourced from U.S. Energy Information Administration AEO Reference Case 2005 modified for high projected oil prices
Facility life	40 years for central cases with exceptions; 20 years for forecourt cases with exceptions
Fixed operating costs during startup	75%–100% of total for central cases; 75% for forecourt cases
Forecourt compressed hydrogen storage	120% of maximum daily production
G&A rate	20% of labor costs
Hydrogen pressure at central gate	300 psig; if higher pressure is inherent to the process, apply pumping power credit for pressure > 300 psig

Hydrogen purity	98% minimum; CO < 10 ppm, sulfur < 10 ppm
Hydrogen storage pressure at forecourt	6,250 psig
Income taxes	35% federal; 6% state; 38.9% effective
Inflation rate	1.9%, but with resultant price of hydrogen in reference year constant dollars
Land cost	$5,000/acre purchased for central cases; $0.3/sqft/month for long-term lease for forecourt cases
Location	Production facilities are assumed to be located in the United States
Non-energy material prices	From H2A Production Model *Non-Energy Material Prices* worksheet
Process contingency	% adjustment to the total initial capital cost such that the result incorporates the mean or expected overall performance. Process contingency was set at 0 for the H2A case studies
Project contingency	% adjustment to the total initial capital cost such that the result represents the mean or expected cost value; periodic replacement capital includes project contingency. Project contingency was set at 15% of direct capital costs for the central H2A case studies and 5% of direct capital costs for the forecourt H2A case studies
Property taxes and business insurance	2%/year of the total initial capital cost
Reference financial structure	100% equity with 10% IRR; model allows debt financing
Reference year dollars	2005, to be adjusted at half-decade increments (e.g., 2005, 2010)
Revenues during startup	50% of total
Sales tax	Not included on basis that facilities and related purchases are wholesale and through a general contractor entity
Salvage value	10% of initial capital for central cases with exceptions; 0% for forecourt cases
Sensitivity variables and ranges	Based on applying best judgment of 10% and 90% confidence limit extremes to the most significant baseline cost and performance parameters
Startup time	1–2 years for central cases; 0.5 years for forecourt cases
Startup year	2005 for current cases; 2025 for future cases (except central future nuclear case, for which it is 2030)
Technology development stage	All central and forecourt case cost estimates are based on mature, commercial facilities
Variable operating costs during startup	75% of total for central cases (except nuclear, for which it is 50%); 50% for forecourt cases
Working capital rate	15% of the annual change in total operating costs

Pressure swing adsorption (PSA) is used for hydrogen purification for all cases except electrolysis. The PSA unit separates the hydrogen from the other components in the shifted gas stream; mainly CO_2, unreacted CO, CH_4, and other hydrocarbons. The hydrogen purity achieved from a PSA unit can be greater than 99.99%. Based on past conversations with industrial gas producers, the shifted gas stream must contain at least 70 mol% hydrogen before it can be purified in the PSA unit. Purification of streams more dilute than this decreases the product purity and recovery of hydrogen. For these analyses, the concentration of hydrogen in the gas stream prior to the PSA may be less than 70 mol%. In these cases, part of the PSA hydrogen product stream is recycled back into the PSA feed. For a 70 mol% hydrogen PSA feed, a hydrogen recovery rate of 85% is typical with a product purity of 99.9 vol%.

Costs for carbon (CO_2) compression, transport to the sequestration site, and injection are calculated in a separate tab for the central production cases that include carbon sequestration. Capital and operating costs for CO_2 compression, installation, and maintenance of the pipeline for transporting the CO_2 to the sequestration site and injection well costs are included. Capital and operating costs for carbon capture are assumed to be included in the capital and operating costs of the production facility. For H2A analysis of hydrogen production processes, it is assumed that only carbon contained in the process feedstock can be captured. Carbon contained in fuels such as natural gas used for steam generation (i.e., as a utility) cannot be captured. However, emissions from fuels used as utilities are included in the total emissions from the process shown in the model's *Results* worksheet.

Central Cases

The following sections describe the processes, inputs, and results for the H2A central production technology cases (see Table 2 for a list of cases). Values that are common to all cases are not shown in these sections; see *Common Input Parameters and Processes* (above) for those values.

Biomass Gasification

Process Description

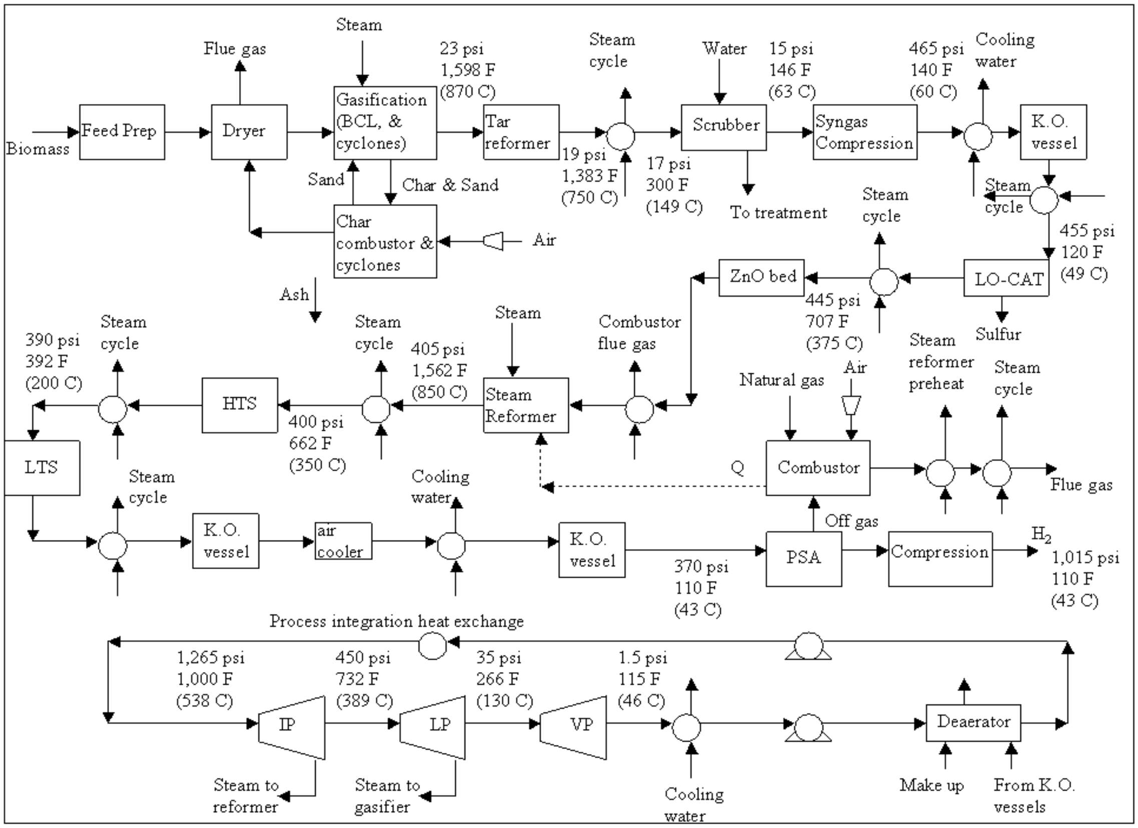
The central current (Figure 13) and future (Figure 14) biomass technology cases are based on the Battelle/FERCO indirectly heated biomass gasifier, conventional catalytic steam reforming, water gas shift, and PSA purification. The biomass feedstock is assumed to be a woody biomass, represented as hybrid poplar. Process energy sources include internally generated steam and electricity, industrial electricity, and commercial natural gas. Aspen Plus® was used to model the thermodynamics of the system.

The as-received wood is dried from 50 wt% moisture to 12 wt% with a rotary dryer. The dryer uses gas from the char combustor as the drying medium. Conveyors and hoppers feed the wood to the low-pressure, indirectly heated, entrained-flow gasifier. Heat for the endothermic gasification reactions is supplied by circulating hot synthetic olivine, which is a calcined magnesium silicate (primarily enstatite [$MgSiO_3$], forsterite [Mg_2SiO_3], and hematite [Fe_2O_3]) used as a sand for various applications, between the gasifier and a char combustor vessel. A small amount of MgO is added to the fresh olivine to keep from forming glass-like bed agglomerations that would result from biomass potassium interacting with the silicate compounds. The gasification medium is steam. The char formed in the gasifier is burned in the combustor to reheat the olivine. Particulate removal is performed through cyclone separators. Ash and any sand particles that are carried over are landfilled.

Reforming ($C_nH_m + nH_2O \leftrightarrow (n+m/2)H_2 + nCO$) and water-gas shift ($CO + H_2O \leftrightarrow CO_2 + H_2$) are the main reactions in the steam reformation process. The reformer is fueled by the PSA offgas, and a small amount of natural gas is added for burner control. The amount of natural gas added is equal to 10% of the heating value of the PSA offgas. The high-temperature shift (HTS) and low-temperature shift (LTS) reactors convert the majority of the CO into CO_2 and H_2 through the water-gas shift reaction.

The steam cycle is integrated with the biomass-to-hydrogen production process and produces power in addition to providing steam for the gasifier and reformer operations. There is an extraction steam turbine/generator, and steam is supplied to the reformer and gasifier from the intermediate- and low-pressure turbine sections, respectively. Preheaters, steam generators, and superheaters are integrated within the process design. The condensate from the syngas compressor and the condensate from the cooled shifted gas stream prior to the PSA are sent to the steam cycle, de-gassed, and combined with the makeup water. A pinch analysis was performed to determine the heat integration of the system.

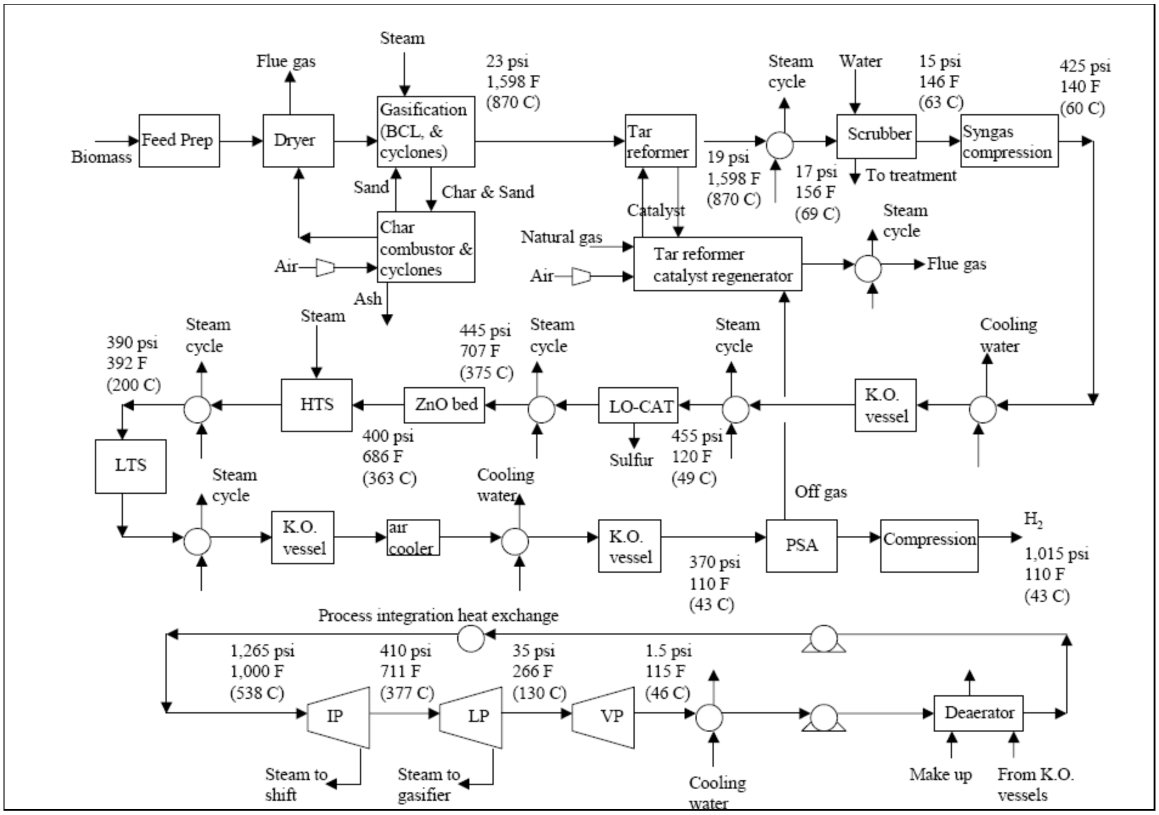
A cooling water system is also included in the Aspen Plus® model to determine the requirements of each cooling water heat exchanger within the hydrogen production system as well as the requirements of the cooling tower.



Although hydrogen compression appears in this diagram, compression energy use and capital costs were removed from the analysis for consistency with other H2A cases.

Figure 13. Process Flow Diagram—Current Biomass Gasification

The design of the future biomass gasification process differs from the current process design in that the tar reformer consists of a reactor vessel and a catalyst regeneration vessel. Additionally, because the tar reformer in the future design reforms a significant amount of the syngas methane, the steam reformer is eliminated from the design. The tar reforming reactor/catalyst regenerator system operates isothermally. The heat required for the tar reforming reactor/catalyst regenerator system is supplied by burning the PSA offgas along with some natural gas. The steam-to-carbon ratio for the shift conversion step is set at 2 mol H_2O/mol C. The biomass-to-hydrogen process is integrated with the steam cycle.



Although hydrogen compression appears in this diagram, compression energy use and capital costs were removed from the analysis for consistency with other H2A cases.

Figure 14. Process Flow Diagram—Future Biomass Gasification

H2A Model Inputs

The following tables show the H2A Model input values for the central current and future biomass technology cases. Default values that are common to all central cases are not shown in the tables below; see *Common Input Parameters and Processes* (page 16) for those values.

Table 4 shows the technical operating parameters and specifications. The central current and future biomass cases produce approximately 140,000 kg of hydrogen per day. Table 5 summarizes the cost inputs, including capital, fixed operating, and variable operating costs. Table 6 summarizes the feedstock and utility energy inputs.

Table 4. Central Biomass: Technical Operating Parameters and Specifications

	Current	Future
Plant design capacity (kg hydrogen/day)	155,000	155,000
Plant output (kg hydrogen/day)	140,000	139,000
Plant output (kg hydrogen/year)	**51,000,000**	**50,800,000**

Table 5. Central Biomass: Cost Input Summary

Capital Costs (million 2005$)		Current	Future
Initial direct	Feed handling and drying	20.1	20.1
	Gasification, tar reforming, quench	17.8	25.3
	Compression and sulfur removal	16.6	17.1
	Steam methane reforming, shift, PSA	32.1	—
	Shift and PSA	—	17.5
	Steam system and power generation	15.3	15.1
	Cooling water and other utilities	3.6	3.6
	Buildings and structures	6.4	6.4
	Total initial direct	*111.9*	*105.0*
Indirect depreciable	Site preparation	1.1	1.0
	Engineering and design	14.5	13.6
	Project contingency	16.8	15.7
	Up-front permitting	10.1	9.4
Indirect non-depreciable	Land cost	0.3	0.3
	Total indirect	*42.8*	*40.1*
	Total initial (initial direct + indirect)	***154.6***	***145.1***
Expected replacement[a]		5.9	8.1
	Total[b]	**160.5**	**153.2**

Fixed Operating Costs (million 2005$/year)		Current	Future
	Labor	5.6	5.2
	G&A	1.1	1.0
	Property taxes and insurance	3.1	2.9
	Material for maintenance and repairs	0.6	0.5
	Total[b]	**10.4**	**9.7**

Variable Operating Costs (million 2005$/year)[c]		Current	Future
Energy feedstocks, utilities, byproducts	Biomass feedstock	27.4	26.4
	Commercial natural gas	2.9	1.4
	Industrial electricity	2.8	2.7
Other materials and byproducts	Cooling water	0.3	0.4
	Process water	0.1	0.1
Other variable operating costs	Other variable[d]	0.1	0.1
	Other material[e]	7.4	5.0
	Waste treatment	1.3	1.2
	Solid waste disposal	0.8	0.7
	Total[b]	**43.2**	**38.0**

[a]Sum of expected replacement capital costs over the 40-year plant life, adjusted to the year in which they are incurred using an NPV calculation.
[b]Components might not add to total owing to rounding.
[c]These costs vary over time in accordance with H2A Model price tables; startup year costs are shown here.
[d]MgO, boiler chemicals, #2 diesel fuel, and cooling tower chemicals.
[e]Catalyst, olivine, and other materials.

Table 6. Central Biomass: Energy Input Summary

Feed	Lower Heating Value	Material Use (per kg hydrogen)	
		Current	Future
Biomass feedstock	19.6 MJ/kg	12.8 kg	12.4 kg
Commercial natural gas	36.6 MJ/Nm^3	0.2 Nm^3	0.1 Nm^3
Industrial electricity	3.6 MJ/kWh	1.0 kWh	0.9 kWh

H2A Model Results

The following tables and figures show the H2A Model results for the central current and future biomass technology cases. Table 7 summarizes the cost results. The central current biomass case produces hydrogen for $1.61/kg. The future case produces hydrogen for $1.47/kg. Capital costs and feedstock costs each account for approximately one-third of the hydrogen cost.

Table 7. Central Biomass: Cost Results Summary

Cost Component	Cost Contribution ($/kg hydrogen)		Percentage of Hydrogen Cost[a]	
	Current	Future	Current	Future
Capital	0.53	0.50	33%	34%
Fixed O&M	0.21	0.20	13%	14%
Feedstock	0.55	0.53	34%	36%
Other raw material	0.15	0.10	9%	7%
Other variable[b]	0.16	0.13	10%	9%
Total hydrogen cost	**1.61**	**1.47**	**100%**	**100%**

[a]Total might not add to 100% owing to rounding.
[b]Including utilities.

Table 8 summarizes the process energy results. Most of the energy input is in the form of biomass feedstock. The only energy output is hydrogen. The process energy efficiency on a lower heating value basis (LHV energy output divided by energy input) is 46.1% for the current case and 48.3% for the future case. These are process energy inputs only and do not include energy used upstream of the process, e.g., to extract fossil fuels and refine feedstocks or to collect and transport biomass resources. The results also do not include energy used for hydrogen compression, storage, and dispensing or for consumption in vehicles; thus they do not represent life-cycle or "well-to-wheel" energy use.

Table 8. Central Biomass: Process Energy Results Summary

Energy Component	Energy Input (MJ per kg hydrogen)		Energy Output (MJ per kg hydrogen)	
	Current	**Future**	**Current**	**Future**
Biomass feedstock	251	242	—	—
Commercial natural gas	6	3	—	—
Industrial electricity	4	3	—	—
Hydrogen	—	—	120	120
Total	**261**	**249**	**120**	**120**

Process energy efficiency (LHV)[a]
Current = 46.1%
Future = 48.3%

[a]Process energy efficiency = hydrogen energy output ÷ all feedstock and process energy inputs.

Table 9 (current) and Table 10 (future) show the values used in the sensitivity analyses. See *Sensitivity Analysis* (page 15) to learn how these values were chosen. Figure 15 (current) and Figure 16 (future) show the sensitivity analysis results. For both cases, total capital cost has the largest effect on hydrogen price, followed by feedstock price and production process energy efficiency. Operating capacity factor, total fixed operating cost, and labor requirement have smaller effects on hydrogen cost.

Table 9. Central Current Biomass: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Labor requirement (FTE)	25	54	70
Total fixed operating cost (million $)	7.3	10.4	13.5
Operating capacity factor	0.95	0.90	0.80
Plant efficiency	55%	46%	35%
Biomass feedstock price ($/kg)	0.029	0.042	0.054
Total capital investment (million $)	100.0	154.6	220.0

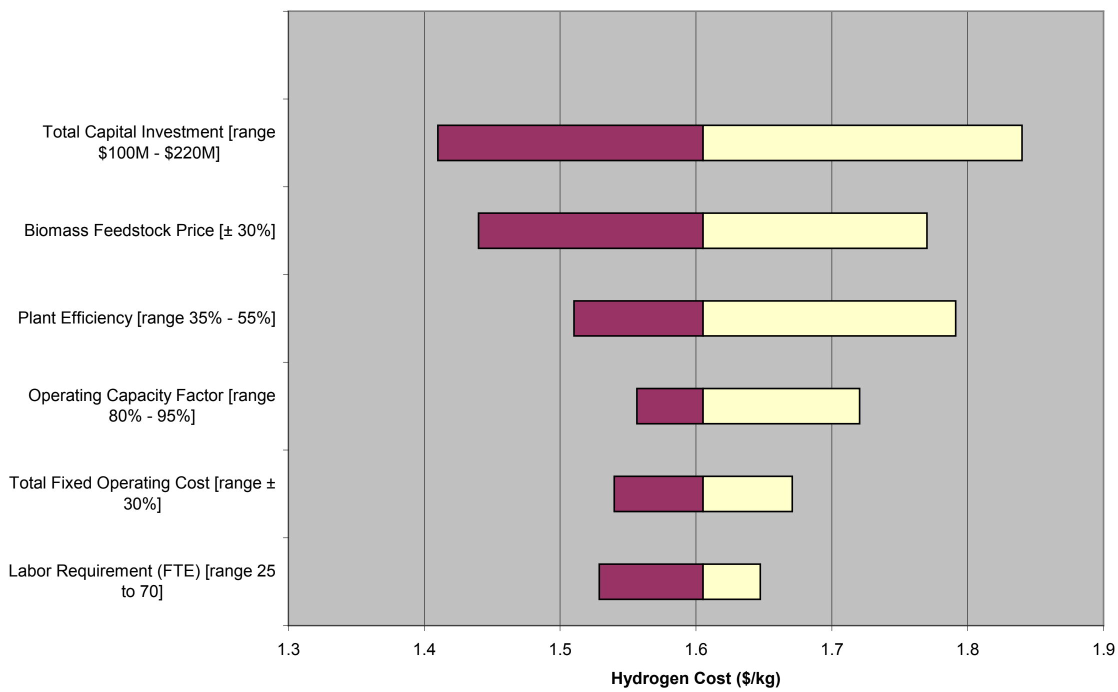


Figure 15. Central Current Biomass: Sensitivity Analysis Results

Table 10. Central Future Biomass: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Labor requirement (FTE)	20	50	60
Total fixed operating cost (million $)	6.8	9.7	12.6
Operating capacity factor	0.95	0.90	0.80
Plant efficiency	58%	48%	38%
Biomass feedstock price ($/kg)	0.029	0.042	0.054
Total capital investment (million $)	95.0	145.1	220.0

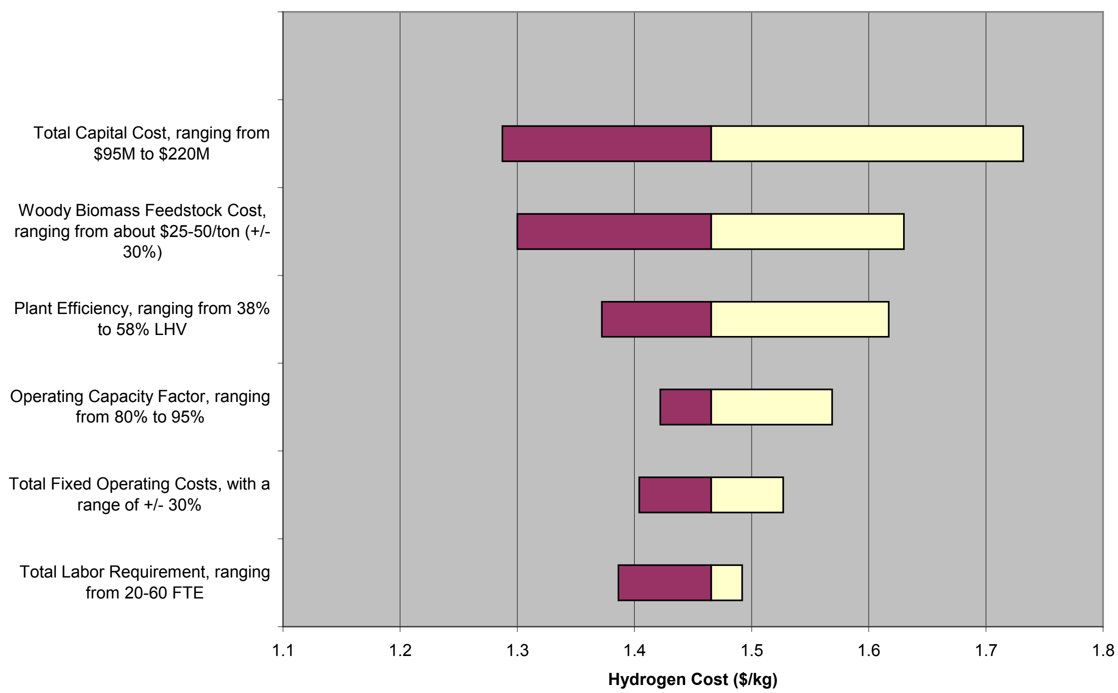


Figure 16. Central Future Biomass: Sensitivity Analysis Results

Grid Electrolysis

Process Description

The central current and future grid electrolysis technology cases (Figure 17) are based on a standalone grid-powered electrolyzer system with a total hydrogen production capacity of 52,000 kg/day. The system is based on the Hydro bi-polar alkaline electrolyzer system (Atmospheric Type No.5040 - 5150 Amp DC); for the future case, improvements in cost and performance were determined in consultation with the H2A development group. The total electrolyzer system consists of 50 electrolyzer units, each capable of producing 485 Nm^3 of hydrogen per hour. The electrolyzer units use high-purity process water for electrolysis. Potassium hydroxide (KOH) is needed for the electrolyte in the system. The system includes the following equipment: transformer, thyristor, electrolyzer unit, lye tank, feed water demineralizer, hydrogen scrubber, gas holder, two compressor units to 30 bar (435 psig), deoxidizer, and twin tower dryer.

The electrolyzer system receives AC grid electricity, which is converted via transformer and rectifier subsystems into DC electricity for use by the electrolyzer stack. The transformer subsystem is an oil-immersed, ambient air-cooled unit, manufactured to IEC-76. The rectifier subsystem converts the AC voltage to DC voltage using thyristors. Cooling is generally accomplished via forced air achieved by fans on the bottom of the rectifier cabinet but can also be accomplished with cooling water. The electrolyzer system uses 4.8 kWh (current case) or 4.0 kWh (future case) of electricity per Nm^3 of hydrogen produced, i.e., 53.4 kWh (current case) or 44.7 kWh (future case) per kg of hydrogen produced.

The electrolyzer system requires high-purity water to avoid deterioration of electrolyzer performance. Process water is demineralized and softened to a specific resistance of 1–2 megaohm/cm in the water demineralizer unit. The system requires 1 L/Nm^3 (2.939 gal/kg) of hydrogen produced.

The electrolyzer system produces hydrogen and oxygen from the electrolysis of feed water. The gas from each cell in the electrolyzer stack is collected in the hydrogen and oxygen flow channels and fed into the gas/lye (KOH) separators. The lye, separated from the produced gas, is recycled through the lye pump, through the lye cooler, and back into the lye tank. Excess heat in the electrolyzer is removed by the lye cooler. Oxygen is removed from the lye in the oxygen/lye separator. The system modeled does not capture the oxygen gas, but capture of the high-purity oxygen gas is a possibility. Saturated hydrogen gas from the hydrogen/lye separator is fed to the gas scrubber subsystem, which purifies the hydrogen. The hydrogen gas is held in a small gas holder unit and is compressed to 435 psig. Following compression, residual oxygen is removed from the hydrogen gas by the deoxidizer unit, and the hydrogen gas is dried in the twin tower dryer. The purity of the hydrogen gas coming off the electrolyzer stack is 99.9%. Following the gas purifier, deoxidizer, and dryer stages, the purity of hydrogen increases to 99.9998% (2 ppm).

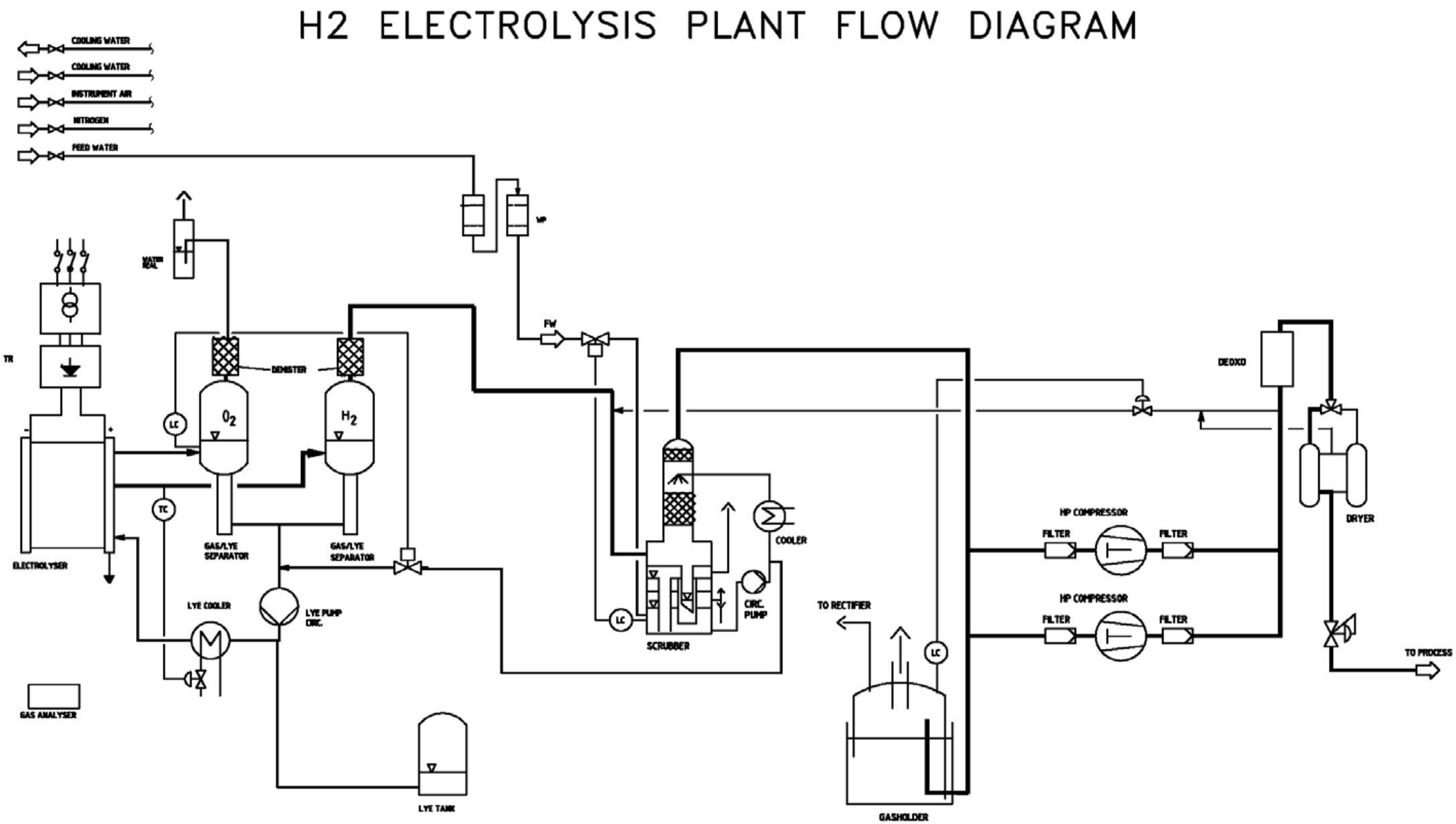


Figure 17. Process Flow Diagram—Grid Electrolysis

H2A Model Inputs

The following tables show the H2A Model input values for the central current and future grid electrolysis technology cases. Default values that are common to all central cases are not shown in the tables below; see *Common Input Parameters and Processes* (page 16) for those values.

Table 11 shows the technical operating parameters and specifications. The central current and future grid electrolysis cases produce approximately 51,000 kg of hydrogen per day. Table 12 summarizes the cost inputs, including capital, fixed operating, and variable operating costs. Table 13 summarizes the energy inputs.

Table 11. Central Grid Electrolysis: Technical Operating Parameters and Specifications

	Current	**Future**
Plant design capacity (kg hydrogen/day)	52,000	52,000
Plant output (kg hydrogen/day)	51,000	51,000
Plant output (kg hydrogen/year)	**19,000,000**	**19,000,000**

Table 12. Central Grid Electrolysis: Cost Input Summary

Capital Costs (million 2005$)		Current	Future
Initial direct	Electrolyzer units (50)	30.2	11.2
	Transformer/rectifier units (50)	5.7	2.1
	Compressor units to 30 bar/435 psig (100)	27.4	10.2
	Gas holders (50)	14.2	5.3
	Balance of plant[a]	17.0	6.3
	Total initial direct	*94.4*	*35.1*
Indirect depreciable	Site preparation	0.9	0.4
	Engineering and design	4.7	1.8
	Project contingency	9.4	3.5
	Up-front permitting	0.9	0.4
Indirect non-depreciable	Land cost	0.03	0.03
	Total indirect	*16.1*	*6.0*
	Total initial (initial direct + indirect)	***110.4***	***41.1***
Expected replacement[b]		13.1	7.1
	Total[c]	**123.5**	**48.2**

Fixed Operating Costs (million 2005$/year)		Current	Future
	Labor	0.3	0.3
	G&A	0.1	0.1
	Property taxes and insurance	2.2	0.8
	Material for maintenance and repairs	2.8	1.1
	Total[c]	**5.4**	**2.2**

Variable Operating Costs (million 2005$/year)[d]		Current	Future
Energy feedstocks, utilities, byproducts	Industrial electricity	54.9	46.8
Other materials and byproducts	Cooling water	0.4	0.4
	Process water	0.1	0.1
	Compressed inert gas	0.01	0.01
Other variable operating costs	Other material[e]	0.4	0.4
	Total[c]	**55.9**	**47.8**

[a]Includes gas purifier (hydrogen scrubber), feed-water purifier/demineralizer, lye tank, deoxidizer, and twin tower drier.
[b]Sum of expected replacement capital costs over the 40-year plant life, adjusted to the year in which they are incurred using an NPV calculation.
[c]Components might not add to total owing to rounding.
[d]These costs vary over time in accordance with H2A Model price tables; startup year costs are shown here.
[e]Electrolyte solution.

Table 13. Central Grid Electrolysis: Energy Input Summary

Feed	Lower Heating Value	Material Use (per kg hydrogen)	
		Current	Future
Industrial electricity	3.6 MJ/kWh	53.4 kWh	44.7 kWh

H2A Model Results

The following tables and figures show the H2A Model results for the central current and future grid electrolysis technology cases. Table 14 summarizes the cost results. The central current grid electrolysis case produces hydrogen for \$4.50/kg. The future case produces hydrogen for \$3.24/kg. Feedstock (electricity) is the largest expense for both cases. Capital costs decrease substantially from the current to the future timeframe.

Hydrogen from the central electrolysis cases is more expensive than from the forecourt electrolysis cases (\$4.23/kg for current and \$3.10 for future) because the forecourt electrolyzer is assumed to be “skid mounted” and inexpensive to install, whereas the installation and coordination of multiple electrolyzers for the central cases are assumed to be more costly.

Table 14. Central Grid Electrolysis: Cost Results Summary

Cost Component	Cost Contribution (\$/kg hydrogen)		Percentage of Hydrogen Cost[a]	
	Current	Future	Current	Future
Capital	1.16	0.47	26%	15%
Fixed O&M	0.32	0.13	7%	4%
Feedstock	2.96	2.58	66%	80%
Other raw material	0.02	0.02	1%	1%
Other variable[b]	0.03	0.03	1%	1%
Total hydrogen cost	**4.50**	**3.24**	**100%**	**100%**

[a]Total might not add to 100% owing to rounding.
[b]Including utilities.

Table 15 summarizes the process energy results. All the energy input is in the form of electricity feedstock. The only energy output is hydrogen. Results are reported based on both the lower heating value (LHV) and higher heating value (HHV) of hydrogen. The HHV—which accounts for the latent and sensible heat of vaporization of the combustion products (i.e., water vapor) between 150°C and 25°C—represents the actual amount of energy required to electrolyze water and is a more thermodynamically accurate measure for this production technology because liquid water (not water vapor) is produced. However, LHV—which assumes the latent and sensible heat of vaporization of the combustion products are not recovered between 150°C and 25°C—is also given because it is customary to use LHV to measure the performance of hydrogen production technologies. The HHV of hydrogen is 18% greater than the LHV. Only LHV results are given for the other hydrogen production technologies discussed in this report because LHV is a more accurate measure for those higher-temperature processes.

The process energy efficiency (energy output divided by energy input) is 73.8% for the current case and 88.2% for the future case (HHV). These are process energy inputs only and do not include energy used upstream of the process. The results also do not include energy used for hydrogen compression, storage, and dispensing or for consumption in vehicles; thus they do not represent life-cycle or "well-to-wheel" energy use.

Table 15. Central Grid Electrolysis: Process Energy Results Summary

Energy Component	Energy Input (MJ per kg hydrogen)		Energy Output (MJ per kg hydrogen)	
	Current	Future	Current	Future
Industrial electricity	192	161	—	—
Hydrogen	—	—	120	120
Total	**192**	**161**	**120**	**120**

Process energy efficiency (LHV)[a]
Current = 62.4%
Future = 74.6%

Process energy efficiency (HHV)[a]
Current = 73.8%
Future = 88.2%

[a]Process energy efficiency = Hydrogen energy output ÷ all feedstock and process energy inputs.

Table 16 (current) and Table 17 (future) show the values used in the sensitivity analyses. See *Sensitivity Analysis* (page 15) to learn how these values were chosen. Figure 18 (current) and Figure 19 (future) show the sensitivity analysis results. For both cases, electricity price has the largest effect on hydrogen price.

Table 16. Central Current Grid Electrolysis: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Total fixed operating cost (million $)	3.8	5.4	7.0
Operating capacity factor	0.98	0.97	0.85
Electrolyzer system efficiency (HHV)	79%	74%	69%
Uninstalled electrolyzer cost ($/kW)	575	675	775
Industrial electricity price ($/kWh)	0.039	0.055	0.072

Table 17. Central Future Grid Electrolysis: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Total fixed operating cost (million $)	1.6	2.2	2.9
Operating capacity factor	0.98	0.97	0.85
Electrolyzer system efficiency (HHV)	90%	88%	76%
Uninstalled electrolyzer cost ($/kW)	240	300	450
Industrial electricity price ($/kWh)	0.039	0.055	0.073

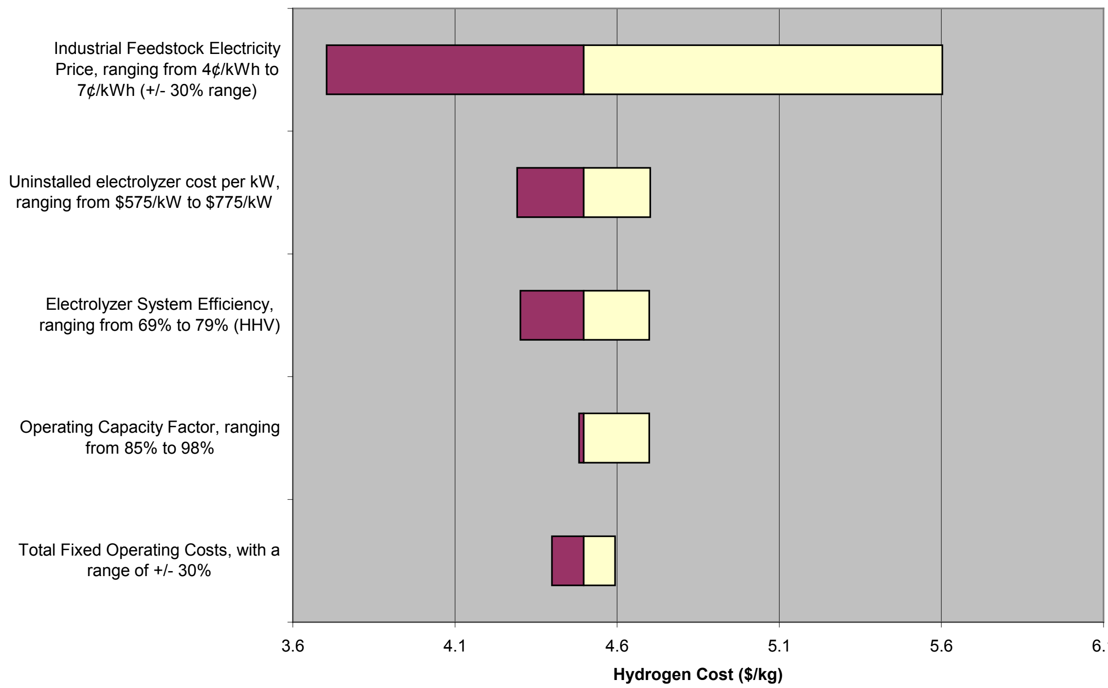


Figure 18. Central Current Grid Electrolysis: Sensitivity Analysis Results

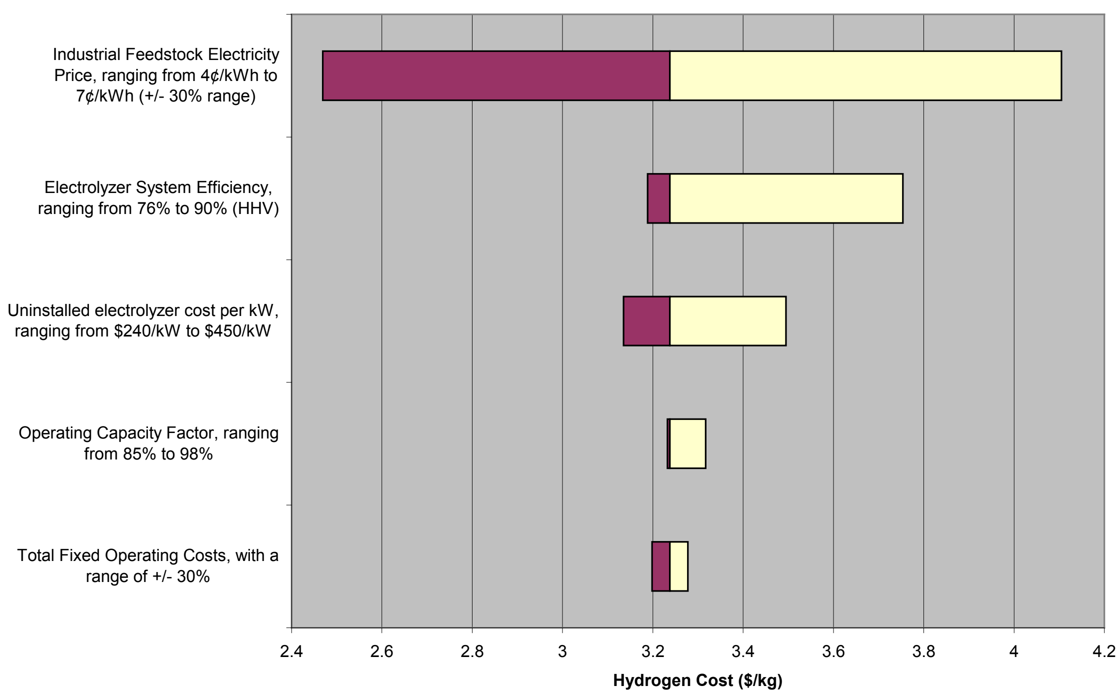


Figure 19. Central Future Grid Electrolysis: Sensitivity Analysis Results

Coal without CO_2 Sequestration

Process Description

The central current coal without CO_2 sequestration technology case is based on commercially available process technologies (Figure 20). The plant modeled uses a Wabash River–scale Conoco-Phillips (EGas) gasifier, conventional gas cooling, commercial shift conversion and acid gas cleanup, commercial sulfuric acid technology, and commercial PSA. The EGas gasifier is the gasifier of choice for this study because it has been operated on bituminous and sub-bituminous coals. The process design includes an air separation unit for supplying concentrated oxygen to the gasifier and an amine unit for separation of a CO2-rich stream (~ 93mol% CO2) from the hydrogen-rich stream. The up-front removal of nitrogen from the process and separation of a CO2-rich stream makes this process design amenable to carbon sequestration. The only additional process step required for CO2 sequestration would be compression and transport of the CO2 stream.

The future case is based on longer-term process technology (Figure 21). Hot raw gas from the transport gasifier is sent to the hot gas desulfurization process for desulfurization. Elemental sulfur is produced as a byproduct. The clean filtered hot gas then goes through a hydrogen separation membrane, where the shift reaction occurs and hydrogen is separated from CO_2. A portion of the hydrogen is fired to heat compressed air entering the ITM oxygen separation unit. Heat for the air is also extracted from the hot CO_2 stream with a high-temperature heat exchanger. Additional hydrogen is used to produce power from a combined cycle solid oxide fuel cell (SOFC) and gas turbine. Pure oxygen produced from the ITM is cooled and compressed for use in the gasifier. The remaining hydrogen is compressed for product delivery. The CO_2-rich stream is fired with oxygen and expanded to recover energy as power. Note that the CO2-rich stream shown as "CO2 Product" in Figure 21 is not captured for CO2 sequestration, and therefore it is a waste stream rather than a product stream.

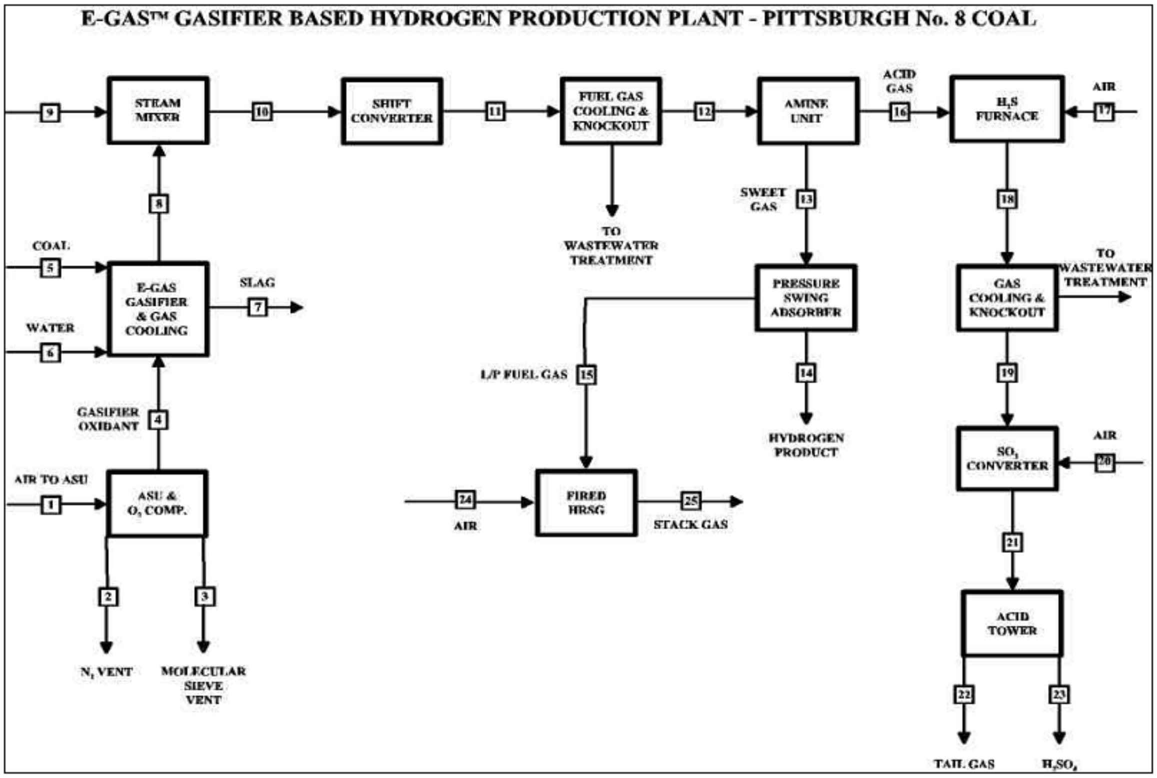


Figure 20. Process Flow Diagram—Current Coal without CO_2 Sequestration

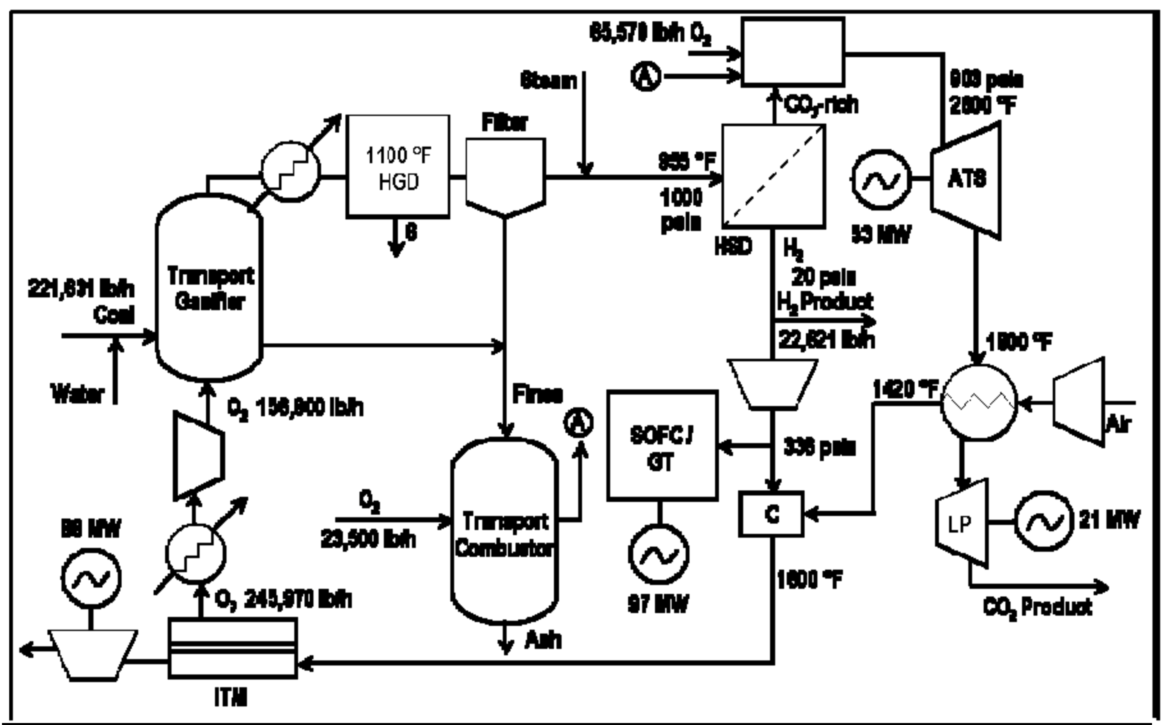


Figure 21. Process Flow Diagram—Future Coal without CO_2 Sequestration

H2A Model Inputs

The following tables show the H2A Model input values for the central current and future coal without CO_2 sequestration technology cases. Default values that are common to all central cases are not shown in the tables below; see *Common Input Parameters and Processes* (page 16) for those values.

Table 18 shows the technical operating parameters and specifications. The central coal without CO_2 sequestration cases produce approximately 255,000 (current) and 222,000 (future) kg of hydrogen per day. Table 19 summarizes the cost inputs, including capital, fixed operating, and variable operating costs. Table 20 summarizes the energy inputs.

Table 18. Central Coal without CO_2 Sequestration: Technical Operating Parameters and Specifications

	Current	Future
Plant design capacity (kg hydrogen/day)	284,000	246,000
Plant output (kg hydrogen/day)	255,000	222,000
Plant output (kg hydrogen/year)	**93,000,000**	**81,000,000**

Table 19. Central Coal without CO_2 Sequestration: Cost Input Summary

Capital Costs (million 2005$)		Current	Future
Initial direct	Coal handling prep and feed	31.0	35.5
	Feedwater and misc. BOP systems	5.3	9.4
	Gasifier and accessories	84.2	69.7
	Air separation unit	58.5	34.8
	Hydrogen separation and gas cleanup	55.4	54.0
	Expander/generators and SOFC/CT	—	68.8
	HRSG ducting and stack	18.8	—
	HRSG and steam turbine generator	—	15.1
	Steam turbine generator	15.8	—
	Cooling water system	6.4	2.3
	Ash handling system	8.0	6.5
	Accessory electric plant	12.2	26.5
	I&C	10.6	12.9
	Buildings and structures	6.0	6.3
	Zinc oxide polisher	0.6	—
	Total initial direct	*312.8*	*341.9*
Indirect depreciable	Site preparation	5.1	3.4
	Engineering and design	31.3	34.2
	Project contingency	46.9	85.5
	Up-front permitting	38.3	34.2
Indirect non-depreciable	Land cost	1.3	1.3
	Total indirect	*122.9*	*158.5*
	Total initial (initial direct + indirect)	***435.7***	***500.5***
Expected replacement[a]		14.8	24.8
	Total[b]	**450.5**	**525.3**

Fixed Operating Costs (million 2005$/year)		Current	Future
	Labor	10.4	12.5
	G&A	2.1	2.5
	Property taxes and insurance	8.7	10.0
	Material for maintenance and repairs	1.9	2.1
	Total[b]	**23.1**	**27.0**

Variable Operating Costs (million 2005$/year)[c]		Current	Future
Energy feedstocks, utilities, byproducts	Coal feedstock	29.7	30.6
	Industrial electricity byproduct[d]	(8.9)	(36.4)
Other materials and byproducts	Process water	0.5	0.4
	Sulfuric acid byproduct[d]	(0.0)	—
Other variable operating costs	Other variable	2.4	1.0
	Waste treatment	0.1	0.1
	Solid waste disposal	0.9	0.8
	Total[b,d]	**24.7**	**(3.5)**

[a]Sum of expected replacement capital costs over the 40-year plant life, adjusted to the year in which they are incurred using an NPV calculation.
[b]Components might not add to total owing to rounding.
[c]These costs vary over time in accordance with H2A Model price tables; startup year costs are shown here.
[d]Numbers in parentheses represent income.

Table 20. Central Coal without CO_2 Sequestration: Energy Input Summary

Feed/(Byproduct)	Lower Heating Value	Material Use/(Production) (per kg hydrogen)	
		Current	Future
Coal feedstock	27.7 MJ/kg	8.5 kg	9.8 kg
(Industrial electricity)	3.6 MJ/kWh	(3.2 kWh)	(15 kWh)

H2A Model Results

The following tables and figures show the H2A Model results for the central current and future coal without CO_2 sequestration technology cases. Table 21 summarizes the cost results. The central current coal without CO_2 sequestration case produces hydrogen for $1.41/kg. The future case produces hydrogen for $1.45/kg. Capital cost is the largest expense for both cases. Both cases partially offset costs with income from byproduct electricity generation.

Table 21. Central Coal without CO_2 Sequestration: Cost Results Summary

Cost Component	Cost Contribution ($/kg hydrogen)		Percentage of Hydrogen Cost[a]	
	Current	Future	Current	Future
Capital	0.88	1.14	62%	78%
Fixed O&M	0.26	0.35	18%	24%
Feedstock	0.33	0.41	24%	28%
Byproduct credits	(0.10)	(0.47)	(7%)	(33%)
Other variable[b]	0.04	0.03	3%	2%
Total hydrogen cost	1.41	1.45	**100%**	**100%**

[a]Total might not add to 100% owing to rounding.
[b]Including utilities.

Table 22 summarizes the process energy results. All the energy input is in the form of coal feedstock. The energy outputs are hydrogen and electricity. The process energy efficiency (energy output divided by energy input) is 55.8% for the current case and 64.1% for the future case. These are process energy inputs only and do not include energy used upstream of the process. The results also do not include energy used for hydrogen compression, storage, and dispensing or for consumption in vehicles; thus they do not represent life-cycle or "well-to-wheel" energy use.

Table 22. Central Coal without CO_2 Sequestration: Process Energy Results Summary

Energy Component	Energy Input (MJ per kg hydrogen)		Energy Output (MJ per kg hydrogen)	
	Current	Future	Current	Future
Coal feedstock	236	271	—	—
Industrial electricity byproduct	—	—	11	54
Hydrogen	—	—	120	120
Total	**236**	**271**	**131**	**174**

Process energy efficiency (LHV)[a]
Current = 55.8%
Future = 64.1%

[a]Process energy efficiency = process energy output ÷ process energy input.

Table 23 (current) and Table 24 (future) show the values used in the sensitivity analyses. See *Sensitivity Analysis* (page 15) to learn how these values were chosen. Figure 22 (current) and Figure 23 (future) show the sensitivity analysis results. For both cases, total direct capital cost has the largest effect on hydrogen price.

Table 23. Central Current Coal without CO_2 Sequestration: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Labor cost (million $/year)	7.0	10.4	14.0
Plant efficiency	65%	56%	45%
Coal price ($/kg)	0.026	0.037	0.049
Operating capacity factor	0.95	0.90	0.80
Total direct capital cost (million $)	290	313	430

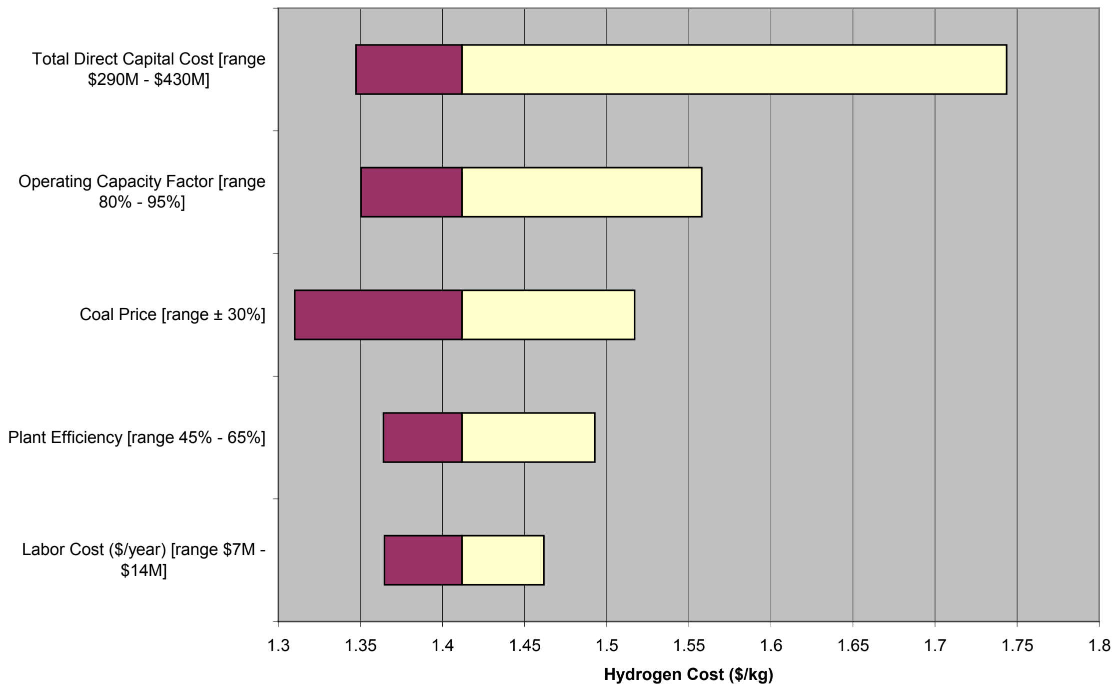


Figure 22. Central Current Coal without CO_2 Sequestration: Sensitivity Analysis Results

Table 24. Central Future Coal without CO_2 Sequestration: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Labor cost (million $/year)	9.0	12.5	15.0
Plant efficiency	70%	64%	50%
Coal price ($/kg)	0.027	0.039	0.050
Operating capacity factor	0.95	0.9	0.8
Total direct capital cost (million $)	310	342	440

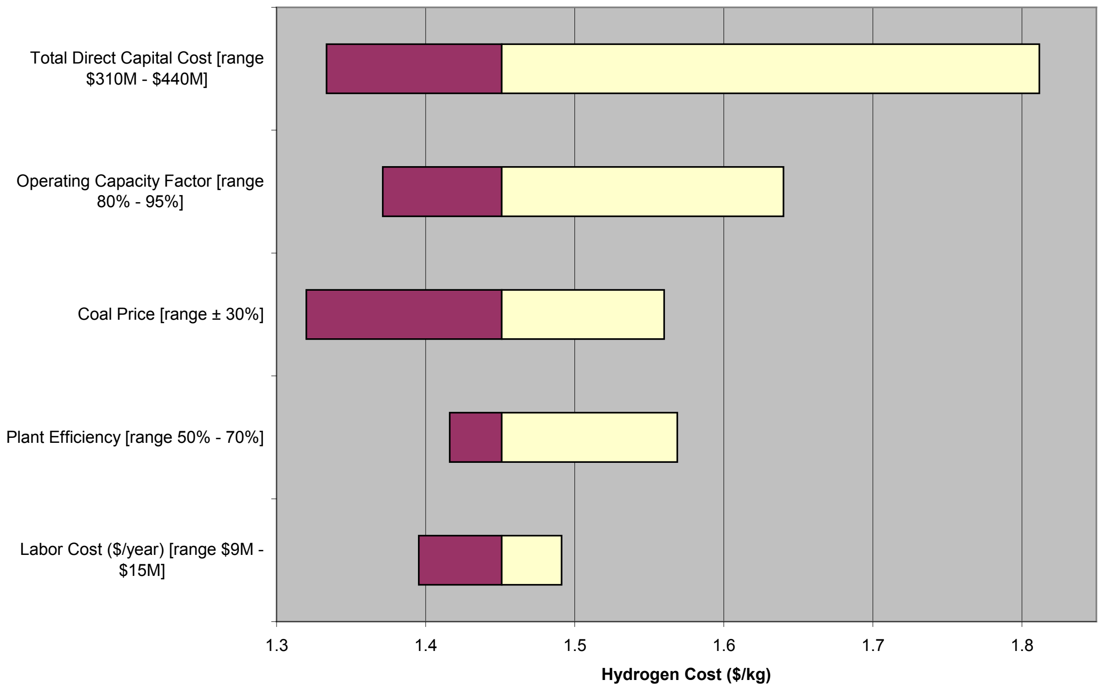


Figure 23. Central Future Coal without CO_2 Sequestration: Sensitivity Analysis Results

Coal with CO_2 Sequestration

Process Description

The central current coal with CO_2 sequestration technology case is based on commercially available process technologies (Figure 24). The plant modeled includes a Wabash River–scale Destec (EGas) gasifier, conventional gas cooling, commercial shift conversion and acid gas cleanup, commercial sulfuric acid technology, and commercial PSA. Two-stage Selexol is used to remove CO_2. CO_2 is compressed to 2,200 psi for sequestration. The EGas gasifier is the gasifier of choice for this study because it has been operated on bituminous and sub-bituminous coals.

For the future case, hot raw gas from the transport gasifier is sent to the hot gas desulfurization process for desulfurization (Figure 25). Elemental sulfur is produced as a byproduct. The clean, filtered hot gas then goes through a hydrogen separation membrane where the shift reaction occurs and hydrogen is separated from CO_2. A portion of the hydrogen is fired to heat compressed air entering the ITM oxygen separation unit. Heat for the air is also extracted from the hot CO_2 stream with a high-temperature heat exchanger. Additional hydrogen is used to produce power from a combined-cycle SOFC. Pure oxygen produced from the ITM is cooled and compressed for use in the gasifier. The remaining hydrogen is compressed for product delivery. The CO_2-rich stream is fired with oxygen and expanded to recover energy as power. The CO_2 exhaust is compressed for pipeline delivery.

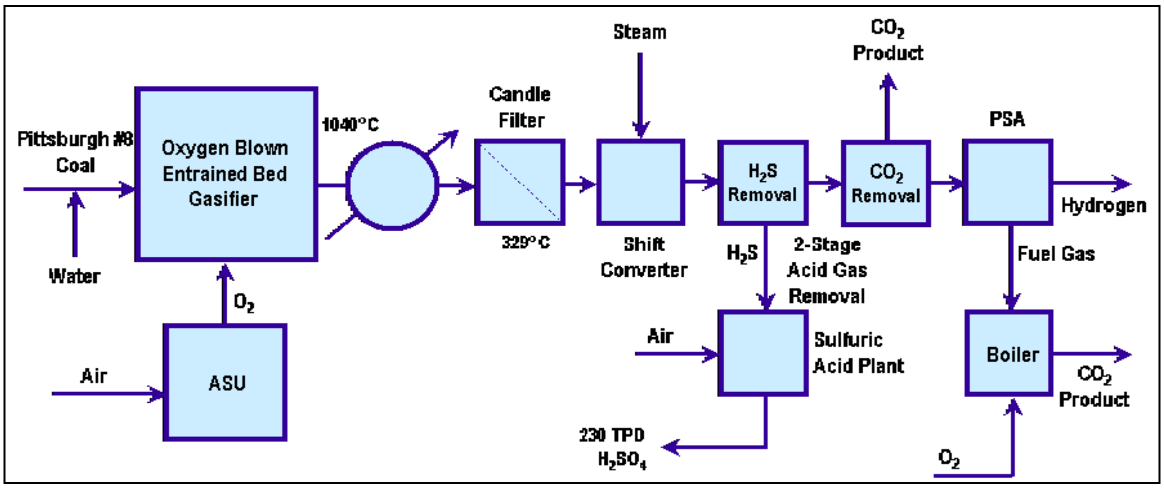


Figure 24. Process Flow Diagram—Current Coal with CO_2 Sequestration

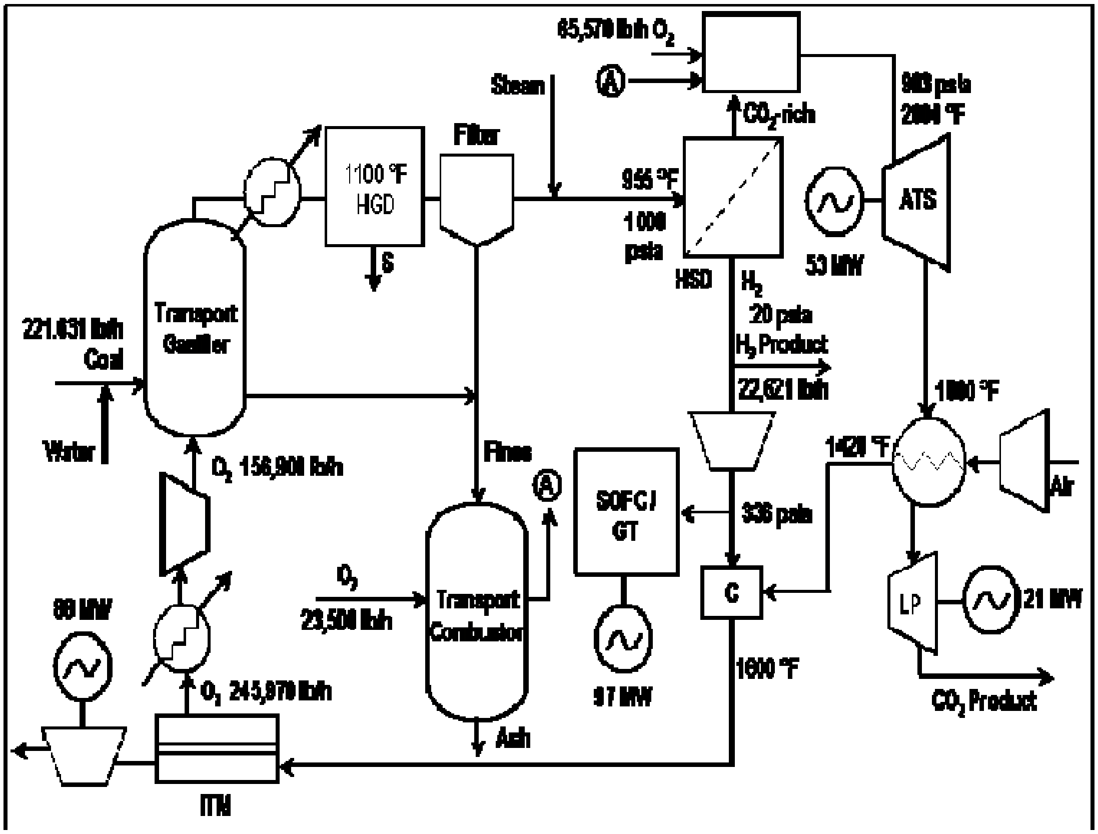


Figure 25. Process Flow Diagram—Future Coal with CO_2 Sequestration

H2A Model Inputs

The following tables show the H2A Model input values for the central current and future coal with CO_2 sequestration technology cases. Default values that are common to all central cases are not shown in the tables below; see *Common Input Parameters and Processes* (page 16) for those values.

Table 25 shows the technical operating parameters and specifications. The central coal with CO_2 sequestration cases produce approximately 277,000 (current) and 222,000 (future) kg of hydrogen per day. Table 26 summarizes the cost inputs, including capital, fixed operating, and variable operating costs. Table 27 summarizes the energy inputs.

Table 25. Central Coal with CO_2 Sequestration: Technical Operating Parameters and Specifications

	Current	Future
Plant design capacity (kg hydrogen/day)	308,000	246,000
Plant output (kg hydrogen/day)	277,000	222,000
Plant output (kg hydrogen/year)	**101,000,000**	**81,000,000**

Table 26. Central Coal with CO_2 Sequestration: Cost Input Summary

Capital Costs (million 2005$)		Current	Future
Initial direct	Coal handling prep and feed	31.0	35.5
	Feedwater and misc. BOP systems	5.3	9.4
	Gasifier and accessories	71.0	69.7
	Air separation unit	77.5	34.8
	Hydrogen separation and gas cleanup	106.3	54.0
	Expander/generators and SOFC/CT	—	68.8
	HRSG ducting and stack	18.8	—
	HRSG and steam turbine generator	—	15.1
	Steam turbine generator	13.6	—
	Cooling water system	5.6	2.3
	Ash handling system	8.0	6.5
	Accessory electric plant	13.2	26.5
	I&C	10.6	12.9
	Buildings and structures	5.6	6.3
	Zinc oxide polisher and CO_2 comp.	24.4	—
	CO_2 compressor	35.7	35.7
	CO_2 injection (site and wells)	2.4	2.4
	CO_2 pipeline	78.6	78.6
	Total initial direct	*507.7*	*458.7*
Indirect depreciable	Site preparation	6.5	4.6
	Engineering and design	50.8	45.9
	Project contingency	76.1	114.7
	Up-front permitting	49.0	45.9
Indirect non-depreciable	Land cost	1.3	1.3
	Total indirect	*183.7*	*212.2*
	Total initial (initial direct + indirect)	***691.4***	***670.9***
Expected replacement[a]		138.0	133.9
	Total[b]	**829.4**	**804.8**

Fixed Operating Costs (million 2005$/year)		Current	Future
	Labor	10.4	12.5
	G&A	2.1	2.5
	Property taxes and insurance	13.8	13.4
	Material for maintenance and repairs	2.3	2.2
	Total[b]	**28.7**	**30.6**

Variable Operating Costs (million 2005$/year)[c]		Current	Future
Energy feedstocks, utilities, byproducts	Coal feedstock	29.7	30.6
	Industrial electricity (for CO_2 sequestration)	9.6	9.8
	Industrial electricity byproduct[d]	—	(9.8)
	Commercial electricity byproduct[d]	—	(30.6)
Other materials and byproducts	Process water	0.5	0.4
Other variable operating costs	Other variable	2.7	1.0
	Waste treatment	0.1	0.1
	Solid waste disposal	0.8	0.8
	CO_2 sequestration O&M	3.5	3.5
	Total[b]	**46.9**	**5.8**

[a]Sum of expected replacement capital costs over the 40-year plant life, adjusted to the year they are incurred using an NPV calculation.
[b]Components might not add to total owing to rounding.
[c]These costs vary over time in accordance with H2A Model price tables; startup year costs are shown here.
[d]Numbers in parentheses represent income.

Table 27. Central Coal with CO_2 Sequestration: Energy Input Summary

Feed/(Byproduct)	Lower Heating Value	Material Use/(Production) (per kg hydrogen)	
		Current	Future
Coal feedstock	27.7 MJ/kg	7.8 kg	9.8 kg
Industrial electricity[a]	3.6 MJ/kWh	1.7 kWh	2.1 kWh
(Industrial electricity)	3.6 MJ/kWh	—	(2.2 kWh)
(Commercial electricity)	3.6 MJ/kWh	—	(12.6 kWh)

[a]Electricity used for CO_2 sequestration.

H2A Model Results

The following tables and figures show the H2A Model results for the central current and future coal with CO_2 sequestration technology cases. Table 28 summarizes the cost results. The central current coal with CO_2 sequestration case produces hydrogen for \$2.05/kg. The future case produces hydrogen for \$2.00/kg. Capital cost is the largest expense for both cases. The future case partially offsets costs with income from byproduct electricity generation.

Table 28. Central Coal with CO_2 Sequestration: Cost Results Summary

Cost Component	Cost Contribution ($/kg hydrogen)		Percentage of Hydrogen Cost[a]	
	Current	Future	Current	Future
Production capital[b]	1.06	1.23	52%	62%
CO_2 sequestration capital[b]	0.21	0.29	10%	14%
Production fixed O&M[c]	0.27	0.36	13%	18%
CO_2 sequestration O&M[c]	0.03	0.04	1%	2%
Feedstock	0.31	0.41	15%	20%
Byproduct credits	—	(0.52)	—	(26%)
Other variable[d]	0.07	0.08	4%	4%
CO_2 sequestration energy	0.10	0.12	5%	6%
Total hydrogen cost	**2.05**	**2.00**	**100%**	**100%**

[a]Total might not add to 100% owing to rounding.
[b]CO_2 capture capital costs are included as part of production capital costs. CO_2 sequestration capital costs include compression, injection, and pipeline costs.
[c]CO_2 capture O&M costs are included as part of production fixed O&M costs. CO_2 sequestration O&M costs include compression, injection, and pipeline costs.
[d]Including utilities.

Table 29 summarizes the process energy results. The energy inputs are in the form of coal feedstock and electricity used for CO_2 sequestration. The energy outputs are hydrogen and electricity. The process energy efficiency (energy output divided by energy input) is 53.7% for the current case and 62.0% for the future case. These are process energy inputs only and do not include energy used upstream of the process. The results also do not include energy used for hydrogen compression, storage, and dispensing or for consumption in vehicles; thus they do not represent life-cycle or "well-to-wheel" energy use.

Table 29. Central Coal with CO_2 Sequestration: Process Energy Results Summary

Energy Component	Energy Input (MJ per kg hydrogen)		Energy Output (MJ per kg hydrogen)	
	Current	Future	Current	Future
Coal feedstock	217	271	—	—
Industrial electricity[a]	6	8	—	—
Industrial electricity byproduct	—	—	—	8
Commercial electricity byproduct	—	—	—	45
Hydrogen	—	—	120	120
Total	**223**	**279**	**120**	**173**

Process energy efficiency (LHV)[b]
Current = 53.7%
Future = 62.0%

[a]Electricity used for CO_2 sequestration.
[b]Process energy efficiency = process energy output ÷ process energy input.

Table 30 (current) and Table 31 (future) show the values used in the sensitivity analyses. See *Sensitivity Analysis* (page 15) to learn how these values were chosen. Figure 26 (current) and Figure 27 (future) show the sensitivity analysis results. For both cases, production process total direct capital cost has the largest effect on hydrogen price.

Table 30. Central Current Coal with CO_2 Sequestration: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Labor cost (million \$/year)	7.0	10.4	14.0
Plant efficiency	60%	54%	45%
Coal price (\$/kg)	0.026	0.037	0.049
Operating capacity factor	0.95	0.9	0.8
CO_2 sequestration capital cost (million \$)	75	117	175
Production process total direct capital cost (million \$)	350	391	500

Table 31. Central Future Coal with CO_2 Sequestration: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Labor cost (million \$/year)	9.0	12.5	15.0
Plant efficiency	65%	62%	50%
Coal price (\$/kg)	0.027	0.039	0.050
Operating capacity factor	0.95	0.9	0.8
CO_2 sequestration capital cost (million \$)	70	117	170
Production process total direct capital cost (million \$)	310	342	420

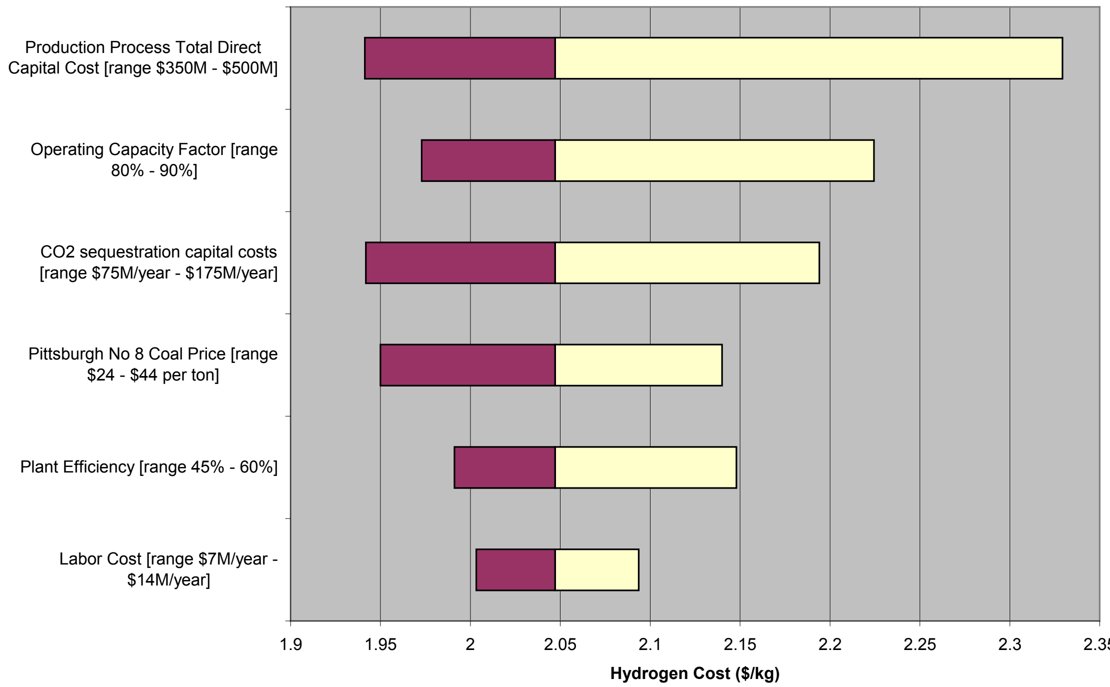


Figure 26. Central Current Coal with CO_2 Sequestration: Sensitivity Analysis Results

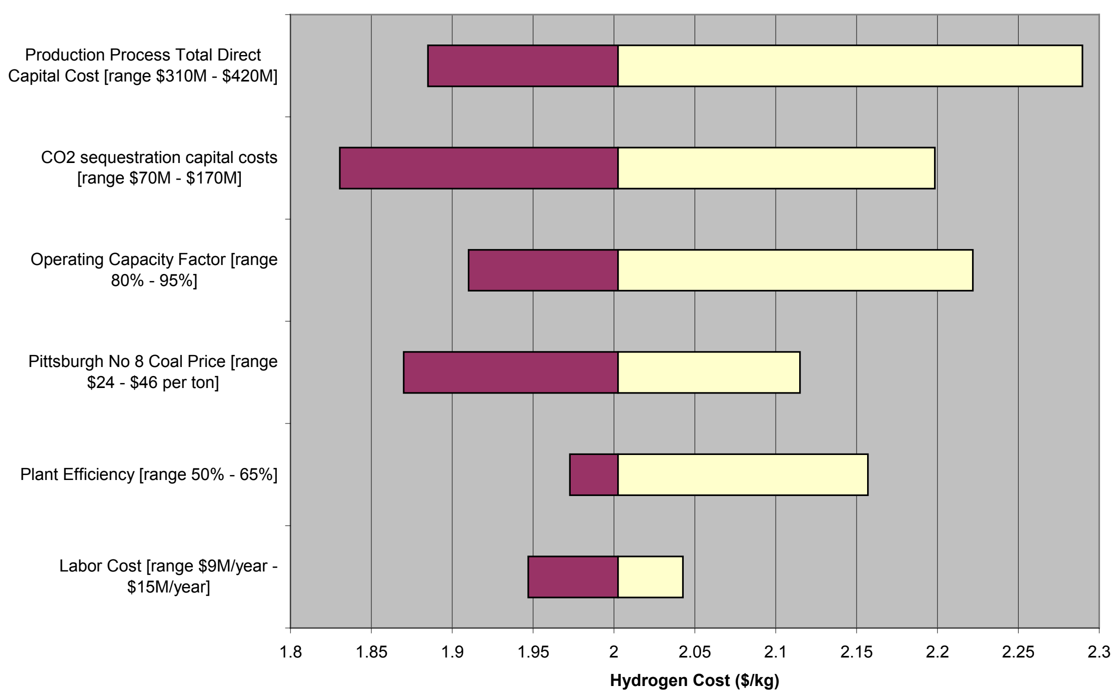


Figure 27. Central Future Coal with CO_2 Sequestration: Sensitivity Analysis Results

Natural Gas without CO_2 Sequestration

Process Description

In the central natural gas without CO_2 sequestration technology cases, natural gas is fed to the plant from the pipeline at a pressure of 450 psia (Figure 28). The gas is generally sulfur free, but odorizers with mercaptans must be cleaned from the gas to prevent contamination of the reformer catalyst. The desulfurized natural gas feedstock is mixed with process steam to be reacted over a nickel-based catalyst contained inside a system of high-alloy steel tubes. The reforming reaction is strongly endothermic, and the metallurgy of the tubes usually limits the reaction temperature to 1,400°–1,700°F. The flue gas path of the fired reformer is integrated with additional boiler surfaces to produce about 700,000 lb/hour steam. Of this, about 450,000 lb/hour is superheated to 450 psia and 750°F to be added to the incoming natural gas. Additional steam from the boiler is sent off site. After the reformer, the process gas mixture of CO and hydrogen passes through a heat-recovery step and is fed into a water gas shift reactor to produce additional hydrogen.

The PSA process is used for hydrogen purification based on its ease of operation and ability to produce high-purity hydrogen and low amounts of CO and CO_2. Shifted gas is fed directly to the PSA unit, where hydrogen is purified up to approximately 99.6%.

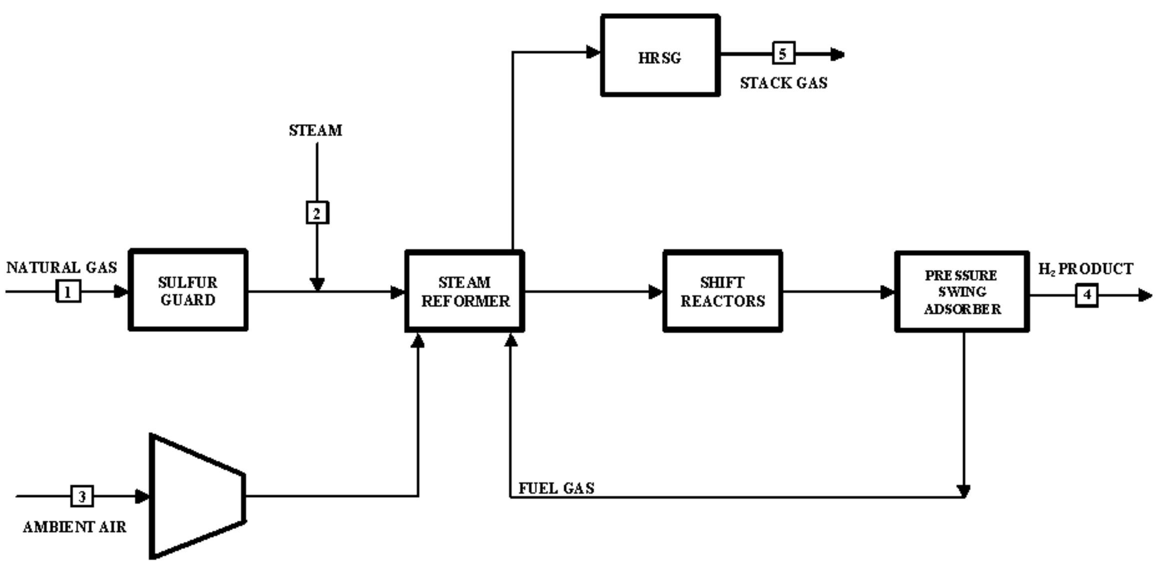


Figure 28. Process Flow Diagram—Natural Gas without CO_2 Sequestration

H2A Model Inputs

The following tables show the H2A Model input values for the central current and future natural gas without CO_2 sequestration technology cases. Default values that are common to all central cases are not shown in the tables below; see *Common Input Parameters and Processes* (page 16) for those values.

Table 32 shows the technical operating parameters and specifications. The central natural gas without CO_2 sequestration cases produce approximately 341,000 kg of hydrogen per day. Table 33 summarizes the cost inputs, including capital, fixed operating, and variable operating costs. Table 34 summarizes the energy inputs.

Table 32. Central Natural Gas without CO_2 Sequestration: Technical Operating Parameters and Specifications

	Current	Future
Plant design capacity (kg hydrogen/day)	379,000	379,000
Plant output (kg hydrogen/day)	341,000	341,000
Plant output (kg hydrogen/year)	**125,000,000**	**125,000,000**

Table 33. Central Natural Gas without CO_2 Sequestration: Cost Input Summary

Capital Costs (million 2005$)		Current	Future
Initial Direct	Process plant equipment	95.9	72.0
	Balance of plant and offsites	38.3	28.7
	SCR NO_x control on stack	0.6	0.4
	Total Initial Direct	*134.8*	*101.1*
Indirect depreciable	Site preparation	1.4	1.1
	Engineering and design	13.5	10.1
	Project contingency	20.2	15.2
	Up-front permitting	10.5	7.9
Indirect non-depreciable	Land cost	0.1	0.1
	Total Indirect	*45.7*	*34.3*
	Total[a]	**180.5**	**135.4**

Fixed Operating Costs (million 2005$/year)		Current	Future
	Labor	2.1	2.1
	G&A	0.4	0.4
	Property taxes and insurance	3.6	2.7
	Material for maintenance and repairs	0.8	0.6
	Total[a]	**6.9**	**5.8**

Variable Operating Costs (million 2005$/year)[b]		Current	Future
Energy feedstocks, utilities, byproducts	Natural gas feedstock	136.2	124.0
	Industrial electricity	3.9	4.2
Other materials and byproducts	Demineralized water	2.1	2.1
	Cooling water	0.0	0.0
Other variable operating costs	Other variable[c]	2.1	2.1
	Total[a]	**144.4**	**132.5**

[a]Components might not add to total owing to rounding.
[b]These costs vary over time in accordance with H2A Model price tables; startup year costs are shown here.
[c]SMR catalyst, SCR, PSA sorbent, and shift catalyst.

Table 34. Central Natural Gas without CO_2 Sequestration: Energy Input Summary

Feed	Lower Heating Value	Material Use (per kg hydrogen)	
		Current	Future
Natural gas feedstock	36.6 MJ/Nm^3	4.5 Nm^3	4.5 Nm^3
Electricity	3.6 MJ/kWh	0.6 kWh	0.6 kWh

H2A Model Results

The following tables and figures show the H2A Model results for the central current and future natural gas without CO_2 sequestration technology cases. Table 35 summarizes the cost results. The central current natural gas without CO_2 sequestration case produces hydrogen for $1.32/kg. The future case produces hydrogen for $1.40/kg. Feedstock cost is the largest expense for both cases and accounts for the hydrogen cost increase in the future case. The AEO 2005 high oil case projected price of industrial natural gas increases only about 6% between 2005 and 2045, the timeframe for the current case, but increases nearly 41% during the timeframe for the future case.

Table 35. Central Natural Gas without CO_2 Sequestration: Cost Results Summary

Cost Component	Cost Contribution ($/kg hydrogen)		Percentage of Hydrogen Cost[a]	
	Current	Future	Current	Future
Capital	0.25	0.20	19%	14%
Fixed O&M	0.06	0.05	4%	3%
Feedstock	0.95	1.08	72%	77%
Other variable[b]	0.07	0.07	5%	5%
Total hydrogen cost	**1.32**	**1.40**	**100%**	**100%**

[a]Total might not add to 100% owing to rounding.
[b]Including utilities.

Table 36 summarizes the process energy results. Energy inputs are in the form of natural gas feedstock and electricity. The energy output is hydrogen. The process energy efficiency (energy output divided by energy input) is 71.9% for the current case and 71.8% for the future case. These are process energy inputs only and do not include energy used upstream of the process. The results also do not include energy used for hydrogen compression, storage, and dispensing or for consumption in vehicles; thus they do not represent life-cycle or "well-to-wheel" energy use.

Table 36. Central Natural Gas without CO_2 Sequestration: Process Energy Results Summary

Energy Component	Energy Input (MJ per kg hydrogen)		Energy Output (MJ per kg hydrogen)	
	Current	Future	Current	Future
Natural gas feedstock	165	165	—	—
Electricity	2	2	—	—
Hydrogen	—	—	120	120
Total	**167**	**167**	**120**	**120**

Process energy efficiency (LHV)[a]
Current = 71.9%
Future = 71.8%

[a]Process energy efficiency = process energy output ÷ process energy input.

Table 37 (current) and Table 38 (future) show the values used in the sensitivity analyses. See *Sensitivity Analysis* (page 15) to learn how these values were chosen. Figure 29 (current) and Figure 30 (future) show the sensitivity analysis results. For both cases, natural gas feedstock price has the largest effect on hydrogen price.

Table 37. Central Current Natural Gas without CO_2 Sequestration: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Fixed operating cost (million $/year)	4.8	6.9	9.0
Operating capacity factor	0.95	0.9	0.8
Total direct capital cost (million $)	110	135	190
Natural gas use (Nm^3/kg hydrogen)	4.0	4.5	6.5
Feedstock price ($/$Nm^3$)	0.17	0.24	0.32

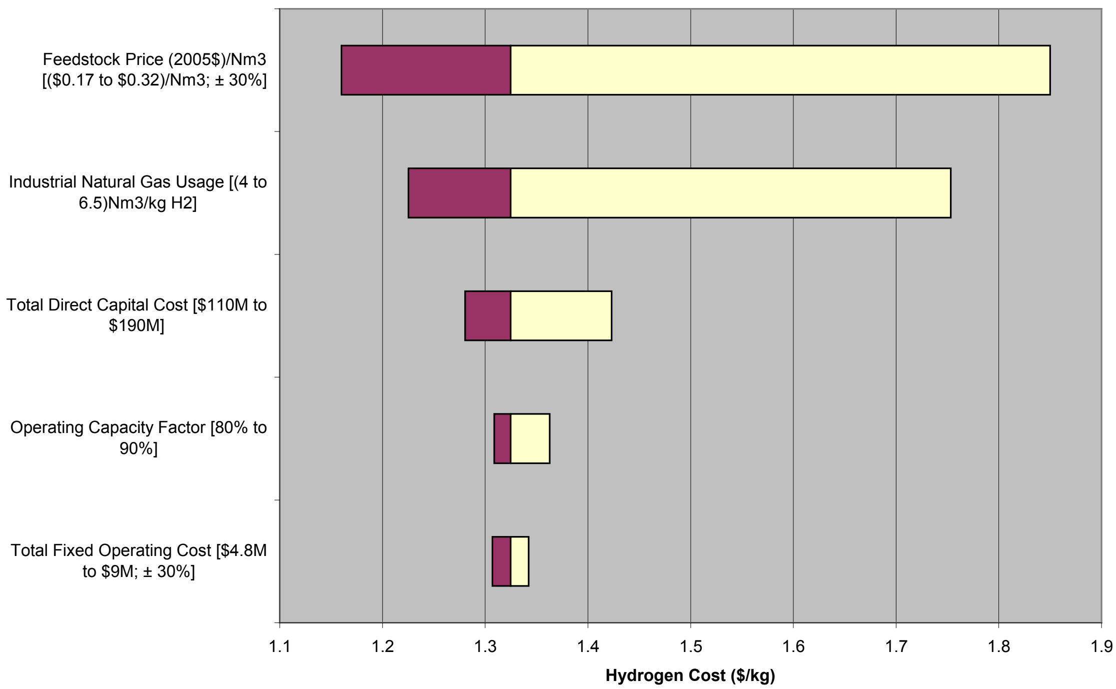


Figure 29. Central Current Natural Gas without CO_2 Sequestration: Sensitivity Analysis Results

Table 38. Central Future Natural Gas without CO_2 Sequestration: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Fixed operating cost (million $/year)	4.1	5.8	7.5
Operating capacity factor	0.95	0.9	0.8
Total direct capital cost (million $)	80	101	160
Natural gas use (Nm^3/kg hydrogen)	4.0	4.5	6.5
Feedstock price ($/$Nm^3$)	0.16	0.22	0.29

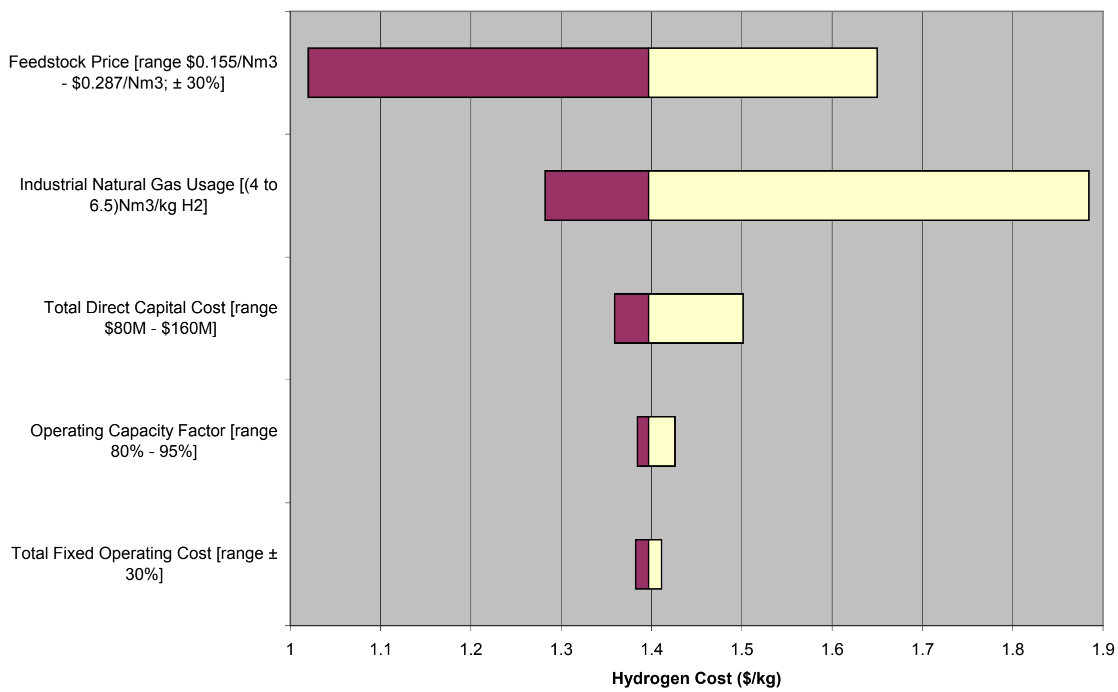


Figure 30. Central Future Natural Gas without CO2 Sequestration: Sensitivity Analysis Results

Natural Gas with CO_2 Sequestration

Process Description

In the central natural gas with CO_2 sequestration technology cases, natural gas is fed to the plant from the pipeline at a pressure of 450 psia (Figure 31). The desulfurized natural gas feedstock is mixed with process steam to be reacted over a nickel-based catalyst contained inside a system of high-alloy steel tubes. The reforming reaction is strongly endothermic, and the metallurgy of the tubes usually limits the reaction temperature to 1400°–1700°F. The flue gas path of the fired reformer is integrated with additional boiler surfaces to produce about 700,000 lb/hour of steam. Of this, about 450,000 lb/hour is superheated to 450 psia and 750°F to be added to the incoming natural gas. Additional steam from the boiler is used to regenerate the CO_2. After the reformer, the process gas mixture of CO and hydrogen passes through a heat recovery step and is fed into a water gas shift reactor to produce additional hydrogen.

The PSA process is used for hydrogen purification based on its ease of operation and ability to produce high-purity hydrogen and low amounts of CO and CO_2. Shifted gas is fed directly to the PSA unit, where hydrogen is purified up to approximately 99.6%. This plant uses a proprietary amine-based process to remove and recover 99% of the CO_2 from the syngas stream. From the shift reactor, gas is passed through an amine tower, where it is contacted counter-currently with a circulating stream of lean aqueous amine solution. The rich amine from the absorber is then sent to a stripper column where the amine is regenerated with a steam reboiler to remove the CO_2 by fractionation. The regenerated CO_2 stream is recovered at 27 psia and 121°F and is compressed to be sent off site. To reach 90% CO_2 removal, a secondary MEA treatment process is installed on the reformer stack.

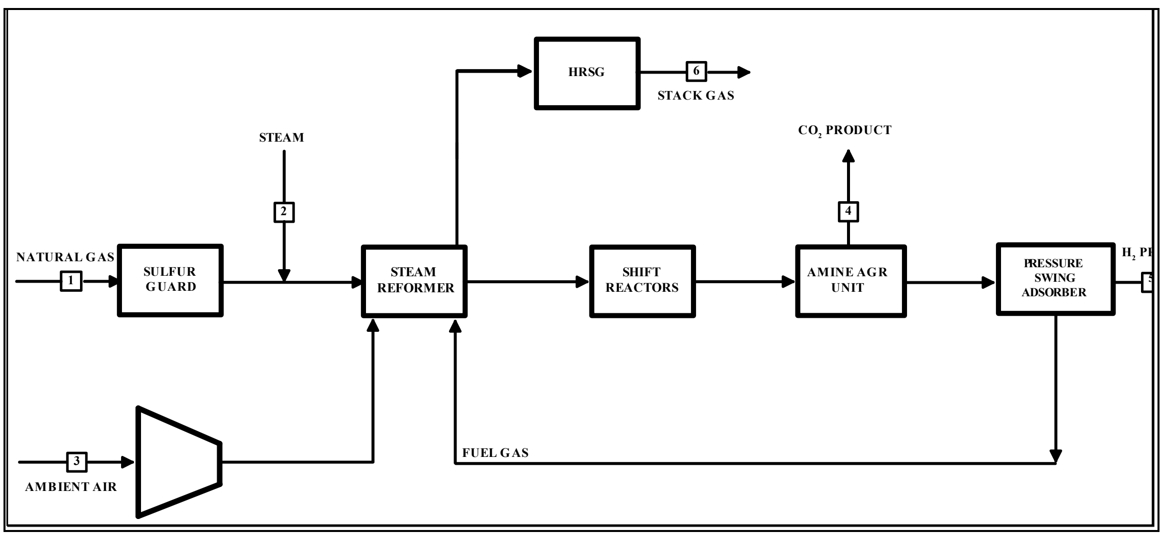


Figure 31. Process Flow Diagram—Natural Gas with CO_2 Sequestration

H2A Model Inputs

The following tables show the H2A Model input values for the central current and future natural gas with CO_2 sequestration technology cases. Default values that are common to all central cases are not shown in the tables below; see *Common Input Parameters and Processes* (page 16) for those values.

Table 39 shows the technical operating parameters and specifications. The central natural gas with CO_2 sequestration cases produce approximately 341,000 kg of hydrogen per day. Table 40 summarizes the cost inputs, including capital, fixed operating, and variable operating costs. Table 41 summarizes the energy inputs.

Table 39. Central Natural Gas with CO_2 Sequestration: Technical Operating Parameters and Specifications

	Current	Future
Plant design capacity (kg hydrogen/day)	379,000	379,000
Plant output (kg hydrogen/day)	341,000	341,000
Plant output (kg hydrogen/year)	**125,000,000**	**125,000,000**

Table 40. Central Natural Gas with CO_2 Sequestration: Cost Input Summary

Capital Costs (million 2005$)		**Current**	**Future**
Initial Direct	Process plant equipment (reformer)	95.9	76.8
	Balance of plant and offsites	38.3	30.7
	Process CO_2 removal	11.7	9.3
	Stack CO_2 removal	10.3	8.2
	CO_2 compressor	28.7	25.2
	CO_2 injection (site and wells)	2.4	2.4
	CO_2 pipeline	64.9	64.9
	Total Initial Direct	*252.1*	*217.4*
Indirect depreciable	Site preparation	2.2	1.8
	Engineering and design	25.2	10.1
	Project contingency	37.8	15.2
	Up-front permitting	16.5	7.9
Indirect non-depreciable	Land cost	0.1	0.1
	Total Indirect	*81.8*	*35.1*
	Total[a]	**334.0**	**252.6**
Fixed Operating Costs (million 2005$/year)		**Current**	**Future**
	Labor	2.6	2.6
	G&A	0.5	0.5
	Property taxes and insurance	6.7	5.1
	Material for maintenance and repairs	1.0	0.8
	Total[a]	**10.8**	**9.0**
Variable Operating Costs (million 2005$/year)[b]		**Current**	**Future**
Energy feedstocks, utilities, byproducts	Natural gas feedstock	135.9	123.7
	Industrial electricity (production)	4.1	4.2
	Industrial electricity (CO_2 sequestration)	5.6	4.9
Other materials and byproducts	Demineralized water	2.1	2.1
	Cooling water	0.0	0.0
Other variable operating costs	Other variable[c]	4.6	4.6
	CO_2 sequestration O&M	2.9	2.7
	Total[a]	**155.2**	**142.3**

[a]Components might not add to total owing to rounding.
[b]These costs vary over time in accordance with H2A Model price tables; startup year costs are shown here.
[c]SMR catalyst, SCR, MEA, PSA sorbent, and shift catalyst.

Table 41. Central Natural Gas with CO_2 Sequestration: Energy Input Summary

Feed	Lower Heating Value	Material Use (per kg hydrogen)	
		Current	Future
Natural gas feedstock	36.6 MJ/Nm^3	4.5 Nm^3	4.5 Nm^3
Electricity (production)	3.6 MJ/kWh	0.6 kWh	0.6 kWh
Electricity (CO_2 sequestration)	3.6 MJ/kWh	0.8 kWh	0.7 kWh

H2A Model Results

The following tables and figures show the H2A Model results for the central current and future natural gas with CO_2 sequestration technology cases. Table 42 summarizes the cost results. The central current natural gas with CO_2 sequestration case produces hydrogen for \$1.64/kg. The future case produces hydrogen for \$1.65/kg. As for the natural gas cases without carbon sequestration, feedstock costs account for the hydrogen cost increase in the future case.

Table 42. Central Natural Gas with CO_2 Sequestration: Cost Results Summary

Cost Component	Cost Contribution ($/kg hydrogen)		Percentage of Hydrogen Cost[a]	
	Current	Future	Current	Future
Production capital[b]	0.31	0.23	19%	14%
CO_2 sequestration capital[b]	0.14	0.12	9%	7%
Production fixed O&M[c]	0.07	0.05	4%	3%
CO_2 sequestration O&M[c]	0.02	0.02	1%	1%
Feedstock	0.94	1.08	58%	65%
Other variable[d]	0.11	0.11	7%	7%
CO_2 sequestration energy	0.04	0.04	2%	2%
Total hydrogen cost	**1.64**	**1.65**	**100%**	**100%**

[a]Total might not add to 100% owing to rounding.
[b]CO_2 capture capital costs are included as part of production capital costs. CO_2 sequestration capital costs include compression, injection, and pipeline costs.
[c]CO_2 capture O&M costs are included as part of production fixed O&M costs. CO_2 sequestration O&M costs include compression, injection, and pipeline costs.
[d]Including utilities.

Table 43 summarizes the process energy results. Energy inputs are in the form of natural gas feedstock and electricity. The energy output is hydrogen. The process energy efficiency (energy output divided by energy input) is 70.8% for the current case and 71.0% for the future case. These are process energy inputs only and do not include energy used upstream of the process. The results also do not include energy used for hydrogen compression, storage, and dispensing or for consumption in vehicles; thus they do not represent life-cycle or "well-to-wheel" energy use.

Table 43. Central Natural Gas with CO_2 Sequestration: Process Energy Results Summary

Energy Component	Energy Input (MJ per kg hydrogen)		Energy Output (MJ per kg hydrogen)	
	Current	Future	Current	Future
Natural gas feedstock	164	164	—	—
Electricity (production)	2	2	—	—
Electricity (CO_2 sequestration)	3	3	—	—
Hydrogen	—	—	120	120
Total	**169**	**169**	**120**	**120**

Process energy efficiency (LHV)[a]
Current = 70.8%
Future = 71.0%

[a]Process energy efficiency = process energy output ÷ process energy input.

Table 44 (current) and Table 45 (future) show the values used in the sensitivity analyses. See *Sensitivity Analysis* (page 15) to learn how these values were chosen. Figure 32 (current) and Figure 33 (future) show the sensitivity analysis results. For both cases, natural gas feedstock price has the largest effect on hydrogen price.

Table 44. Central Current Natural Gas with CO_2 Sequestration: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Fixed operating cost (million \$/year)	7.6	10.8	14.1
Operating capacity factor	0.95	0.9	0.8
Direct production capital cost (million \$)	150	156	220
CO_2 sequestration capital cost (million \$)	75	96	175
Natural gas use (Nm^3/kg hydrogen)	4.2	4.5	6.4
Feedstock price (\$/Nm^3)	0.17	0.24	0.32

Table 45. Central Future Natural Gas with CO_2 Sequestration: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Fixed operating cost (million \$/year)	6.3	8.9	11.6
Operating capacity factor	0.95	0.9	0.8
Direct production capital cost (million \$)	105	125	190
CO_2 sequestration capital cost (million \$)	70	93	170
Natural gas use (Nm^3/kg hydrogen)	4.2	4.5	6.4
Feedstock price (\$/Nm^3)	0.16	0.22	0.29

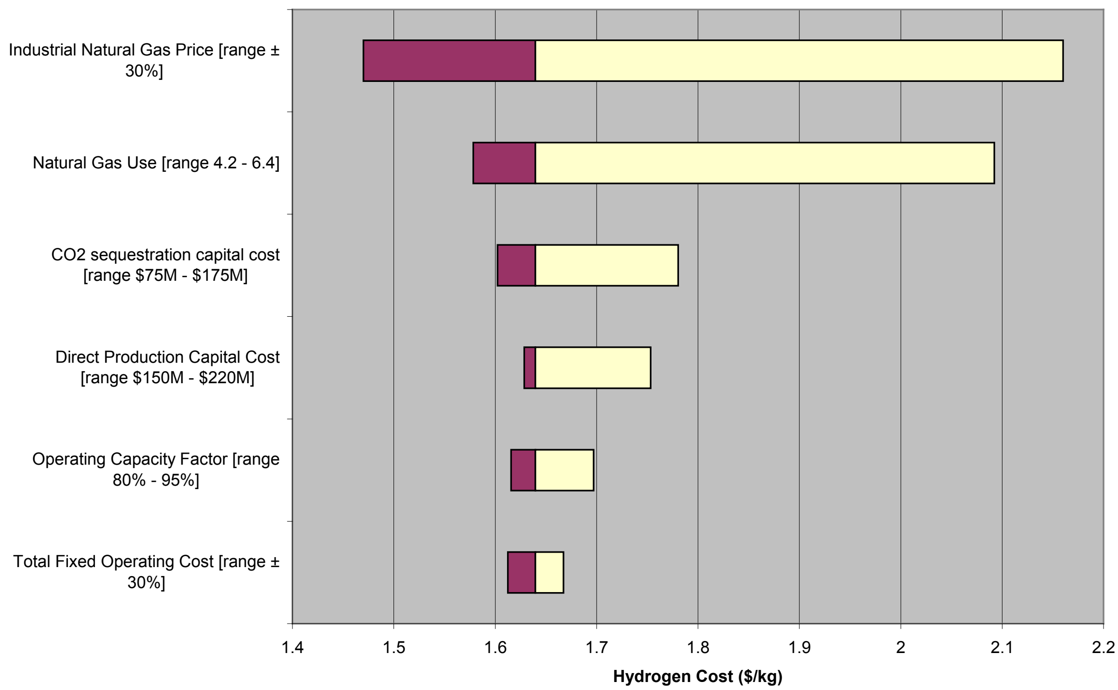


Figure 32. Central Current Natural Gas with CO_2 Sequestration: Sensitivity Analysis Results

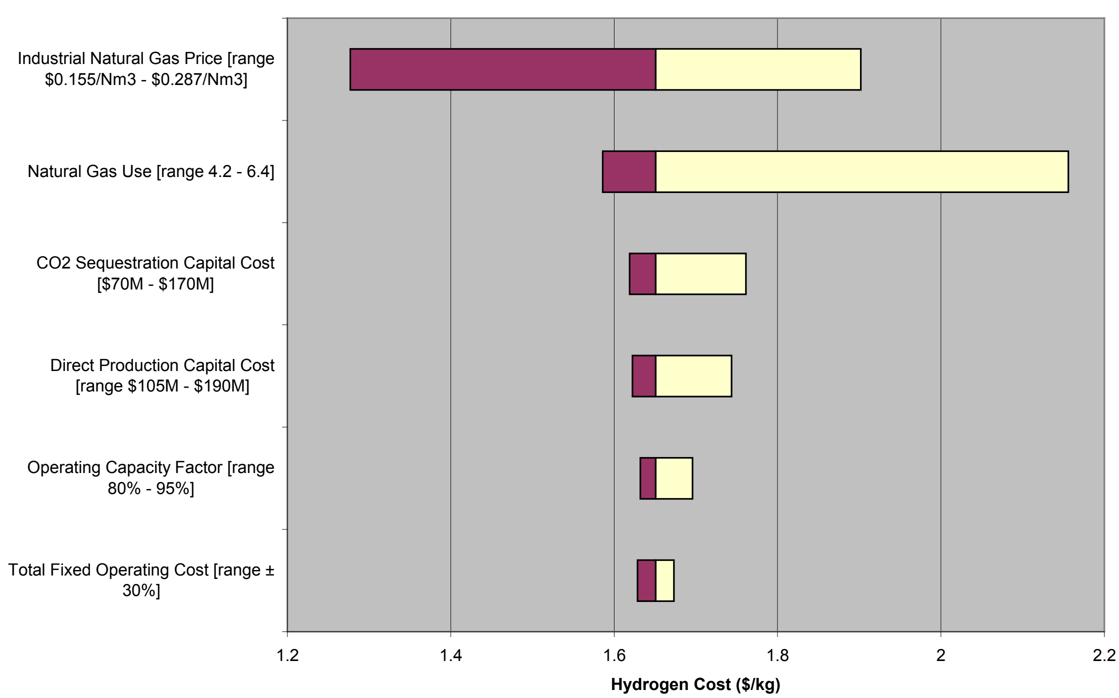


Figure 33. Central Future Natural Gas with CO_2 Sequestration: Sensitivity Analysis Results

Nuclear Energy via High-Temperature Electrolysis

Process Description

The central nuclear energy via high-temperature electrolysis technology case is based on the application of an advanced nuclear plant providing high-efficiency electric power and heat (high temperature steam) to a central high-temperature electrolysis (HTE) plant (Figure 34). The nuclear plant capital and operating costs are not modeled in this case. Instead, heat and electricity are purchased from the nuclear plant. HTE operation and performance is modeled on the design being developed by Idaho National Laboratory (INL), including published pilot-scale plant parameters. Only a future case (startup year 2030) is considered for this technology.

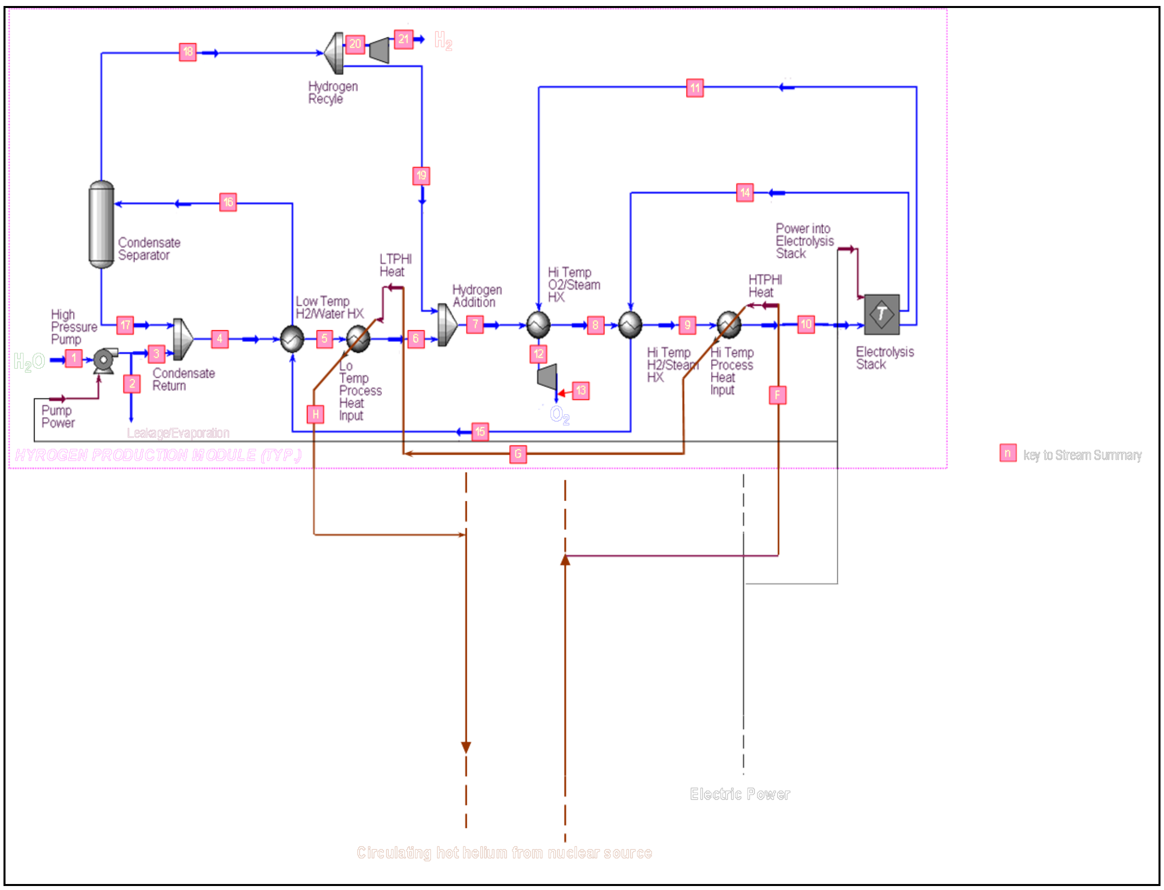


Figure 34. Process Flow Diagram—Nuclear Energy via High-Temperature Electrolysis

H2A Model Inputs

The following tables show the H2A Model input values for the central nuclear energy via high-temperature electrolysis technology case. Default values that are common to all central cases are not shown in the tables below; see *Common Input Parameters and Processes* (page 16) for those values.

Table 46 shows the technical operating parameters and specifications. The central nuclear energy via high-temperature electrolysis technology case produces approximately 734,000 kg of hydrogen per day. Table 47 summarizes the cost inputs, including capital, fixed operating, and variable operating costs. Table 48 summarizes the energy inputs.

Table 46. Central Nuclear Energy via High-Temperature Electrolysis: Technical Operating Parameters and Specifications

	Future
Plant design capacity (kg hydrogen/day)	816,000
Plant output (kg hydrogen/day)	734,000
Plant output (kg hydrogen/year)	**268,000,000**

Table 47. Central Nuclear Energy via High-Temperature Electrolysis: Cost Input Summary

Capital Costs (million 2005$)		Future
Initial direct	Steam generator/superheater	27.9
	Oxygen recuperator	22.4
	Hydrogen recuperator	24.5
	Sweep heater	7.5
	High-temperature (electric) heater	5.3
	Power recovery system	20.0
	Water supply system	4.0
	Miscellaneous plant equipment	10.0
	Electrolyzer system	584.0
	Total initial direct	*705.6*
Indirect depreciable	Project contingency	141.1
	Other depreciable[a]	141.1
Indirect non-depreciable	Land cost	1.0
	Total indirect	*283.2*
	Total initial (initial direct + indirect)	***988.8***
Expected replacement[b]		216.9
	Total[c]	**1,205.7**

Fixed Operating Costs (million 2005$/year)		Future
	Labor	20.0
	G&A	3.0
	Property taxes and insurance	19.8
	Material for maintenance and repairs	13.1
	Total[c]	**55.8**

Variable Operating Costs (million 2005$/year)[d]		Future
Energy feedstocks, utilities, byproducts	Thermal energy	36.6
	Industrial electricity	491.1
Other materials and byproducts	Demineralized water	3.2
	Total[c]	**530.9**

[a]Covers site preparation, engineering and design, licensing, permitting, etc.
[b]Sum of expected replacement capital costs over the 40-year plant life, adjusted to the year in which they are incurred using an NPV calculation.
[c]Components might not add to total owing to rounding.
[d]These costs vary over time in accordance with H2A Model price tables; startup year costs are shown here.

Table 48. Central Nuclear Energy via High-Temperature Electrolysis: Energy Input Summary

Feed	Lower Heating Value	Material Use (per kg hydrogen)
		Future
Thermal energy	3.6 mmBtu/MWh	6.8 kWh
Industrial electricity	3.6 MJ/kWh	33.2 kWh

H2A Model Results

The following tables and figure show the H2A Model results for the central nuclear energy via high-temperature electrolysis technology case. Table 49 summarizes the cost results. This technology case produces hydrogen for \$2.93/kg. Feedstock (electricity and thermal energy) is the largest expense.

Table 49. Central Nuclear Energy via High-Temperature Electrolysis: Cost Results Summary

Cost Component	Cost Contribution (\$/kg hydrogen)	Percentage of Hydrogen Cost[a]
	Future	Future
Capital	0.77	26%
Fixed O&M	0.22	7%
Feedstock	1.94	66%
Other variable[b]	0.01	0.4%
Total hydrogen cost	**2.93**	**100%**

[a]Total might not add to 100% owing to rounding.
[b]Including utilities.

Table 50 summarizes the process energy results. The energy inputs are in the form of thermal energy and electricity. The energy output is hydrogen. The process energy efficiency (energy output divided by energy input) is 83.3%. These are process energy inputs only and do not include energy used upstream of the process. The results also do not include energy used for hydrogen compression, storage, and dispensing or for consumption in vehicles; thus they do not represent life-cycle or "well-to-wheel" energy use.

Table 50. Central Nuclear Energy via High-Temperature Electrolysis: Process Energy Results Summary

Energy Component	Energy Input (MJ per kg hydrogen)	Energy Output (MJ per kg hydrogen)
	Future	Future
Thermal energy	25	—
Industrial electricity	119	—
Hydrogen	—	120
Total	**144**	**120**
Process energy efficiency (LHV)[a] Future = 83.3%		

[a]Process energy efficiency = process energy output ÷ process energy input.

Table 51 shows the values used in the sensitivity analysis. See *Sensitivity Analysis* (page 15) to learn how these values were chosen. Figure 35 shows the sensitivity analysis results. Total direct capital cost has the largest effect on hydrogen price.

Table 51. Central Nuclear Energy via High-Temperature Electrolysis: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Thermal energy cost ($/MWh)	14	20	26
Total fixed operating cost (million $)	41.9	55.8	67.0
Operating capacity factor	0.95	0.90	0.85
Total direct capital cost (million $)	529	706	847

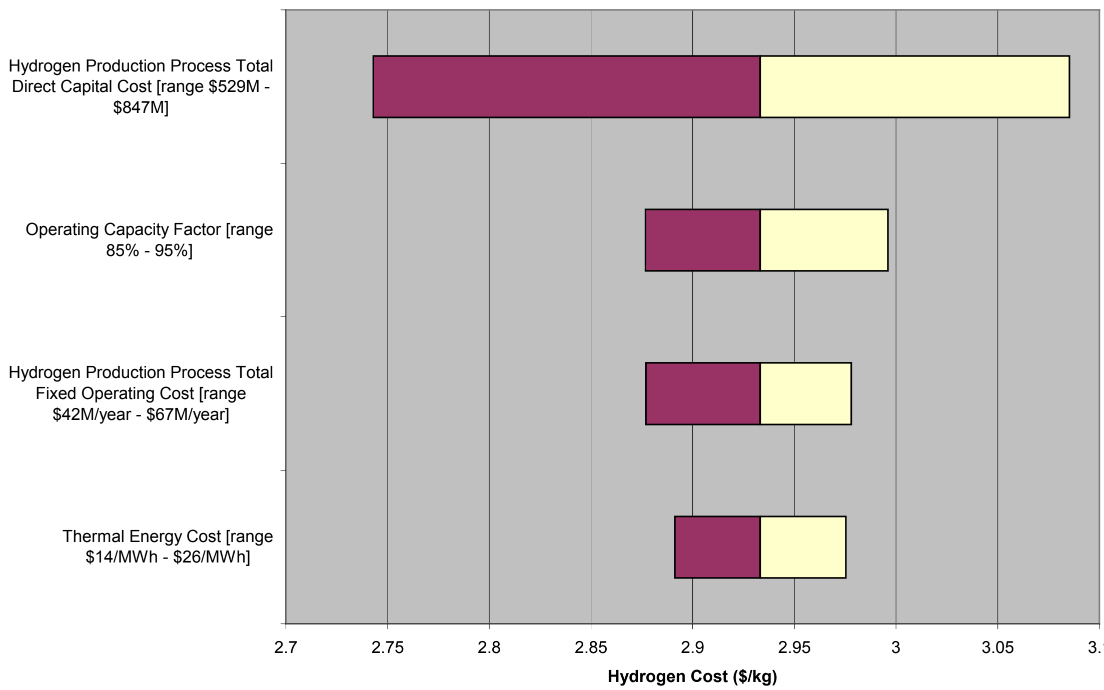


Figure 35. Central Nuclear Energy via High-Temperature Electrolysis: Sensitivity Analysis Results

Forecourt Cases

The following sections describe the processes, inputs, and results for the H2A forecourt production technology cases (see Table 2 for a list of cases). Values common to all cases are not shown in these sections; see *Common Input Parameters and Processes* (page 16) for those values. All forecourt technology cases have a design capacity of 1,500 kg of hydrogen per day and an output of 1,278 kg/day (85.2% capacity factor). Annual hydrogen production is 466,470 kg/year.

Unlike the central technology cases, the forecourt cases include refueling station compression, storage, and dispensing costs and energy use. It is important to keep this in mind when comparing results from forecourt cases with results from central cases. Hydrogen produced from central technologies would also incur CSD costs and energy use in order for the hydrogen to be delivered and dispensed for use in vehicles.

The refueling station subsystem is based on the assumptions and layout of DOE's HDSAM model. That model, which applies only to current/2007 technology, consists of hydrogen compression for storage in low-pressure (up to 2,500 psi) gas cylinders, followed by further compression to 6,250 psi for transfer to a 4-stage, high-pressure cascade system to allow rapid filling of 5,000-psi onboard vehicular hydrogen tanks.

Grid Electrolysis

Process Description

The forecourt current and future grid electrolysis technology cases are based on a standalone grid-powered electrolyzer system with a total hydrogen production capacity of 1,500 kg/day (Figure 36). The system is based on the Hydro bi-polar alkaline electrolyzer system (Atmospheric Type No. 5040–5150 Amp DC) which produces 485 Nm^3 of hydrogen per hour; for the future case, improvements in cost and performance were determined in consultation with the H2A development group. The electrolyzer system modeled is a skid-mounted unit, including the electrolyzer system and necessary auxiliary subsystems. The electrolyzer units use process water and electricity input for electrolysis. Potassium hydroxide (KOH, or lye) is needed for the electrolyte in the system. The system includes the following equipment: transformer, thyristor, electrolyzer unit, lye tank, feed-water demineralizer, hydrogen scrubber, gas holder, two compressor units to 30 bar (435 psig), deoxidizer, and twin tower dryer.

The electrolyzer system receives AC grid electricity, which is converted via transformer and rectifier subsystems into DC electricity for use by the electrolyzer stack. The transformer subsystem is an oil-immersed, ambient air-cooled unit, manufactured to IEC-76. The rectifier subsystem converts the AC voltage to DC voltage using thyristors. Cooling is generally accomplished via forced air achieved by fans on the bottom of the rectifier cabinet but can also be accomplished with cooling water. The electrolyzer system uses 4.8 kWh (current case) or 4.0 kWh (future case) of electricity per Nm^3 of hydrogen produced, i.e., 53.4 kWh (current case) or 44.7 kWh (future case) per kg of hydrogen produced.

The electrolyzer system requires high-purity water to avoid deterioration of electrolyzer performance. Process water is demineralized and softened to a specific resistance of 1–2 megaohm/cm in the water demineralizer unit. The system requires 1 L/Nm^3 (2.939 gal/kg) of hydrogen produced.

The electrolyzer system produces hydrogen and oxygen from the electrolysis of feed water. The gas from each cell in the electrolyzer stack is collected in the hydrogen and oxygen flow channels and fed into the gas/lye (KOH) separators. The lye, separated from the produced gas, is recycled through the lye pump, through the lye cooler, and back into the lye tank. Excess heat in the electrolyzer is removed by the lye cooler. Oxygen is removed from the lye in the oxygen/lye separator. The system modeled does not capture the oxygen gas, but capture of the high-purity oxygen gas is a possibility. Saturated hydrogen gas from the hydrogen/lye separator is fed to the gas scrubber subsystem, which purifies the hydrogen. The hydrogen gas is held in a small gas holder unit and is compressed to 435 psig. Following compression, residual oxygen is removed from the hydrogen gas by the deoxidizer unit, and the hydrogen gas is dried in the twin tower dryer. The purity of the hydrogen gas coming off the electrolyzer stack is 99.9%. Following the gas purifier, deoxidizer, and dryer stages, the purity of hydrogen increases to 99.9998% (2 ppm).

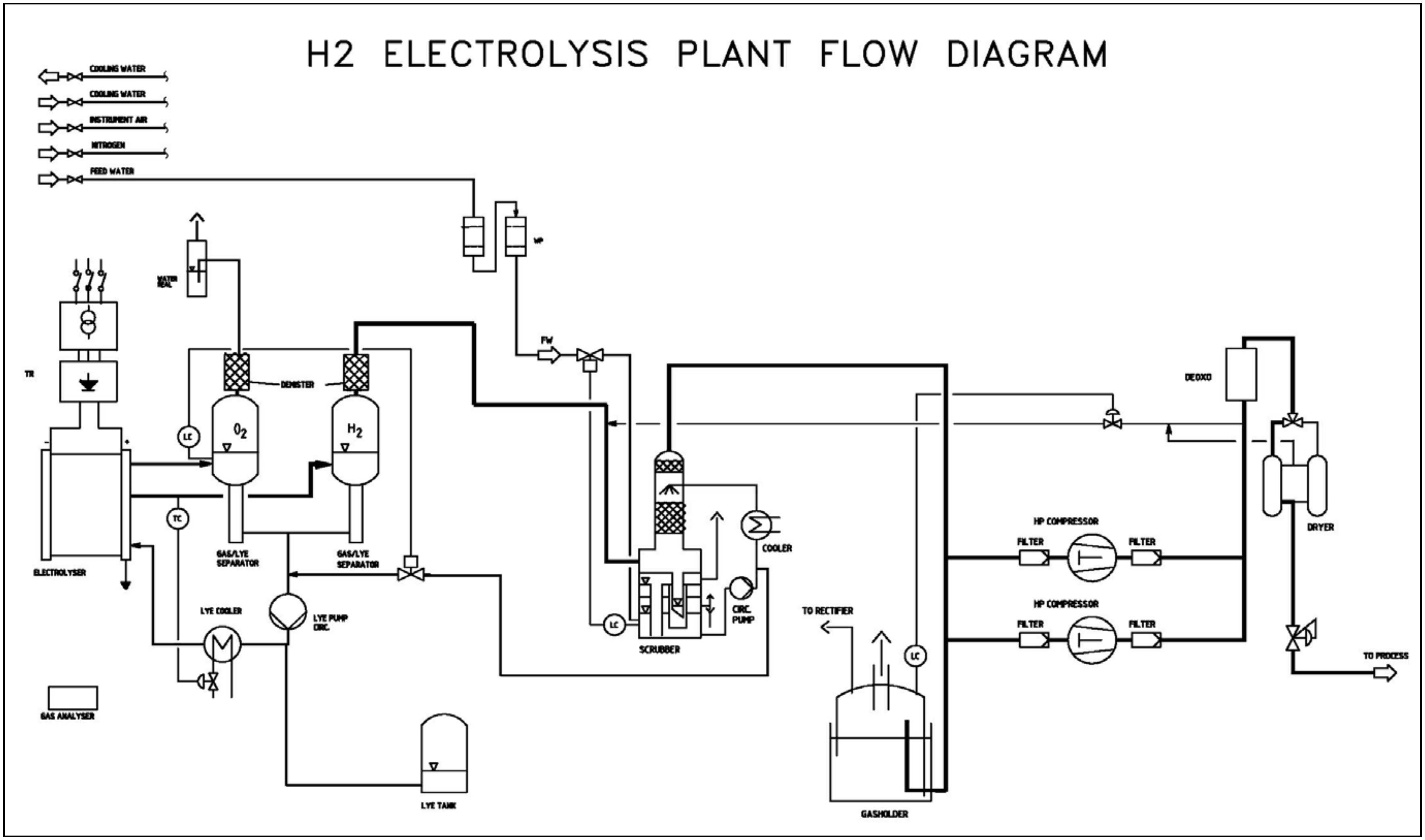


Figure 36. Process Flow Diagram—Grid Electrolysis

H2A Model Inputs

The following tables show the H2A Model input values for the forecourt current and future grid electrolysis technology cases. Default values that are common to all central cases are not shown in the tables below; see *Common Input Parameters and Processes* (page 16) for those values.

Table 52 summarizes the cost inputs, including capital, fixed operating, and variable operating costs; note that the values in this table are in units of thousand 2005$, unlike the central case cost input tables, which have units of million 2005$. Table 53 summarizes the energy inputs.

Table 52. Forecourt Grid Electrolysis: Cost Input Summary

Capital Costs (thousand 2005$)		**Current**	**Future**
Initial direct production	Electrolyzer unit	790	300
	Transformer/rectifier unit	150	60
	Compressor units to 30 bar/435 psig (2)	720	270
	Gas holder	370	140
	Balance of plant[a]	450	170
	Total initial direct production[b]	*2,480*	*920*
Initial direct CSD	Compressors (3)	940	940
	Dispensers (2)	50	50
	Cascade storage (325 kg)	390	390
	Low-pressure storage (16 × 89 kg)	1,670	1,670
	Electrical upgrading	80	80
	Trenching for higher-voltage lines	120	120
	Total initial direct CSD[b]	*3,240*	*3,240*
	Total initial direct[3]	***5,720***	***4,170***
Indirect depreciable (production + CSD)	Site preparation	240	240
	Engineering and design	350	350
	Project contingency	290	210
	Up-front permitting	130	130
	Total indirect[b]	*1,000*	*930*
	Total initial (direct + indirect)[3]	***6,730***	***5,090***
Expected replacement (production + CSD)[c]		690	580
	Total[3]	**7,420**	**5,680**
Fixed Operating Costs (thousand 2005$/year)		**Current**	**Future**
Production + CSD	Licensing, permits, and fees	10	10
	Property taxes and insurance	130	90
	Rent	60	60
	Operating, maintenance, and repairs	170	90
	Labor	40	40
	Overhead and G&A	10	10
	Total[b]	**400**	**290**
Variable Operating Costs (thousand 2005$/year)[d]		**Current**	**Future**
Energy feedstocks, utilities, byproducts	Industrial electricity (feedstock + CSD)	1,430	1,220
Other materials and byproducts	Process water	0	0
	Compressed inert gas	0	0
Other variable operating costs	Other variable operating[e]	0	0
	Other material[f]	20	20
	Total[b]	**1,460**	**1,250**

[a]Includes gas purifier (hydrogen scrubber), feed-water purifier/demineralizer, lye tank, deoxidizer, and twin tower drier.
[b]Components might not add to total owing to rounding (all numbers rounded to the nearest $10,000; all values less than $10,000 are entered as zero).
[c]Sum of expected replacement capital costs over the 20-year plant life, adjusted to the year in which they are incurred using an NPV calculation.
[d]These costs vary over time in accordance with H2A Model price tables; startup year costs are shown here.
[e]Waste disposal costs, non-feedstock fuels, environmental surcharges, etc.
[f]Electrolyte solution.

Table 53. Forecourt Grid Electrolysis: Energy Input Summary

Feed	Lower Heating Value	Material Use (per kg hydrogen)	
		Current	Future
Industrial electricity[1]	3.6 MJ/kWh	55.2 kWh	46.4 kWh

[1] Electricity used for feedstock and compression.

H2A Model Results

The following tables and figures show the H2A Model results for the forecourt current and future grid electrolysis technology cases. Table 54 summarizes the cost results. The forecourt current grid electrolysis case produces hydrogen for $4.23/kg, with a total delivered hydrogen cost of $6.05/kg. The future case produces hydrogen for $3.10/kg, with a total delivered hydrogen cost of $4.92/kg. Feedstock (electricity) is the largest expense for both cases. Capital costs decrease substantially from the current to the future timeframe. CSD costs account for approximately one-third of the delivered hydrogen cost for both cases.

Hydrogen from the forecourt electrolysis cases is less expensive than from the central electrolysis cases ($4.50/kg for current and $3.24 for future) because the forecourt electrolyzer is assumed to be skid mounted and inexpensive to install, whereas the installation and coordination of multiple electrolyzers for the central cases are assumed to be more costly.

Table 54. Forecourt Grid Electrolysis: Cost Results Summary

	Cost Component	Cost Contribution ($/kg hydrogen)		Percentage of Hydrogen Cost[a]	
		Current	Future	Current	Future
Production	Capital	0.98	0.43	16%	9%
	Fixed O&M	0.40	0.16	7%	3%
	Feedstock	2.80	2.47	46%	50%
	Other raw material	0.04	0.04	1%	1%
	Other variable[2]	0.01	0.01	0%	0%
	Total production cost	*4.23*	*3.10*	*70%*	*63%*
CSD	Capital	1.26	1.26	21%	26%
	Fixed O&M	0.46	0.46	8%	9%
	Other variable[b]	0.10	0.10	2%	2%
	Total CSD cost	*1.82*	*1.82*	*30%*	*37%*
Total delivered hydrogen cost		**6.05**	**4.92**	**100%**	**100%**

[a] Total might not add to 100% owing to rounding.
[b] Including utilities.

Table 55 summarizes the process energy results. The energy inputs are in the form of electricity feedstock and electricity used for hydrogen compression. The only energy output is hydrogen. Results are reported based on both the LHV and HHV of hydrogen. The HHV—which accounts for the latent and sensible heat of vaporization of the combustion products (i.e., water vapor) between 150°C and 25°C—represents the actual amount of energy required to electrolyze water and is a more thermodynamically accurate measure for this production technology because liquid water (not water vapor) is produced. However, LHV—which assumes the latent and sensible heat of vaporization of the combustion products are not recovered between 150°C and 25°C—is also given because it is customary to use LHV to measure the performance of hydrogen production technologies. The HHV of hydrogen is 18% greater than the LHV. Only LHV results are given for the other hydrogen production technologies discussed in this report because LHV is a more accurate measure for those higher-temperature processes.

The process energy efficiency (energy output divided by energy input) is 71.6% for the current case and 85.0% for the future case (HHV). These are process energy inputs only and do not include energy used upstream of the process or consumed in vehicles; thus they do not represent life-cycle or "well-to-wheel" energy use.

Table 55. Forecourt Grid Electrolysis: Process Energy Results Summary

Energy Component	Energy Input (MJ per kg hydrogen)		Energy Output (MJ per kg hydrogen)	
	Current	Future	Current	Future
Industrial electricity[a]	198	167	—	—
Hydrogen	—	—	120	120
Total	**198**	**167**	**120**	**120**

Process energy efficiency (LHV)[b]
Current = 60.5%
Future = 71.9%

Process energy efficiency (HHV)[b]
Current = 71.6%
Future = 85.0%

[a] Electricity used for feedstock and compression.
[b] Process energy efficiency = process energy output ÷ process energy input.

Table 56 (current) and Table 57 (future) show the values used in the sensitivity analyses. See *Sensitivity Analysis* (page 15) to learn how these values were chosen. Figure 37 (current) and Figure 38 (future) show the sensitivity analysis results. For both cases, electricity price has the largest effect on hydrogen price.

Table 56. Forecourt Current Grid Electrolysis: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Production fixed operating cost (thousand $)	130	184	260
Storage system capital cost (thousand $)	1,000	1,665	2,200
Industrial electricity use (kWh/kg hydrogen)	50	53.4	57
Forecourt hydrogen storage capacity (% of daily production capacity)	70	117	200
Operating capacity factor	0.95	0.85	0.6
Production system total direct capital cost (thousand $)	1,000	2,480	4,000
Industrial electricity price ($/kWh)	0.039	0.055	0.072

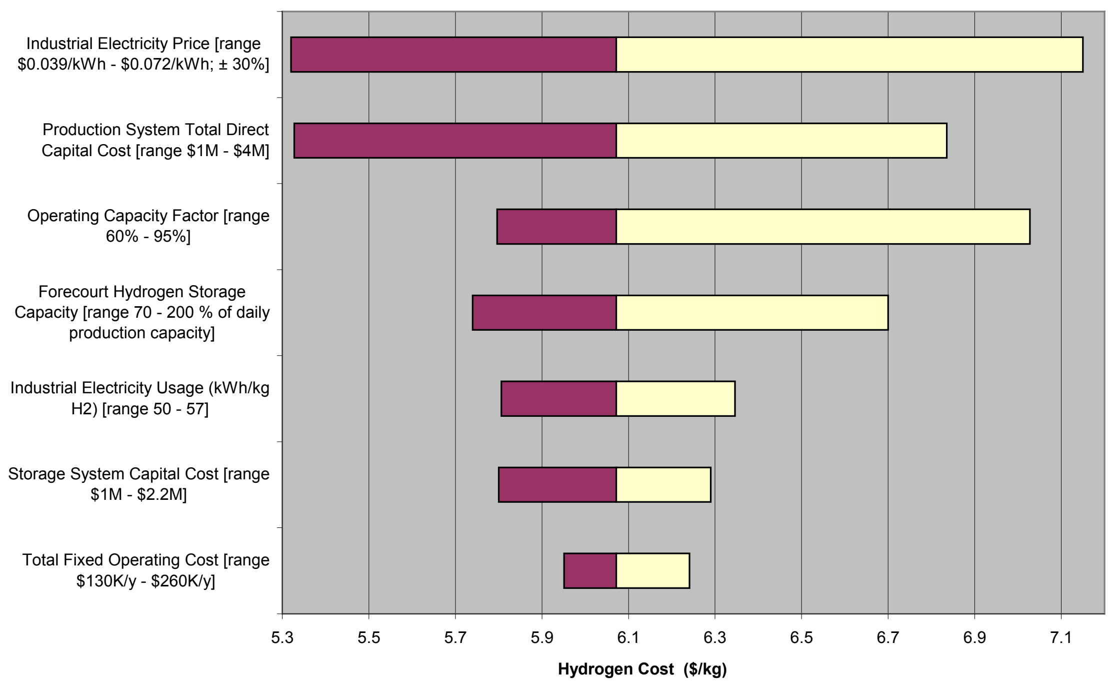


Figure 37. Forecourt Current Grid Electrolysis: Sensitivity Analysis Results

Table 57. Forecourt Future Grid Electrolysis: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Production fixed operating cost (thousand $)	50	73	100
Storage system capital cost (thousand $)	600	1,665	1,950
Industrial electricity use (kWh/kg hydrogen)	44	44.7	52
Forecourt hydrogen storage capacity (% of daily production capacity)	70	117	200
Operating capacity factor	0.95	0.85	0.6
Production system total direct capital cost (thousand $)	500	922	1,700
Industrial electricity price ($/kWh)	0.040	0.057	0.074

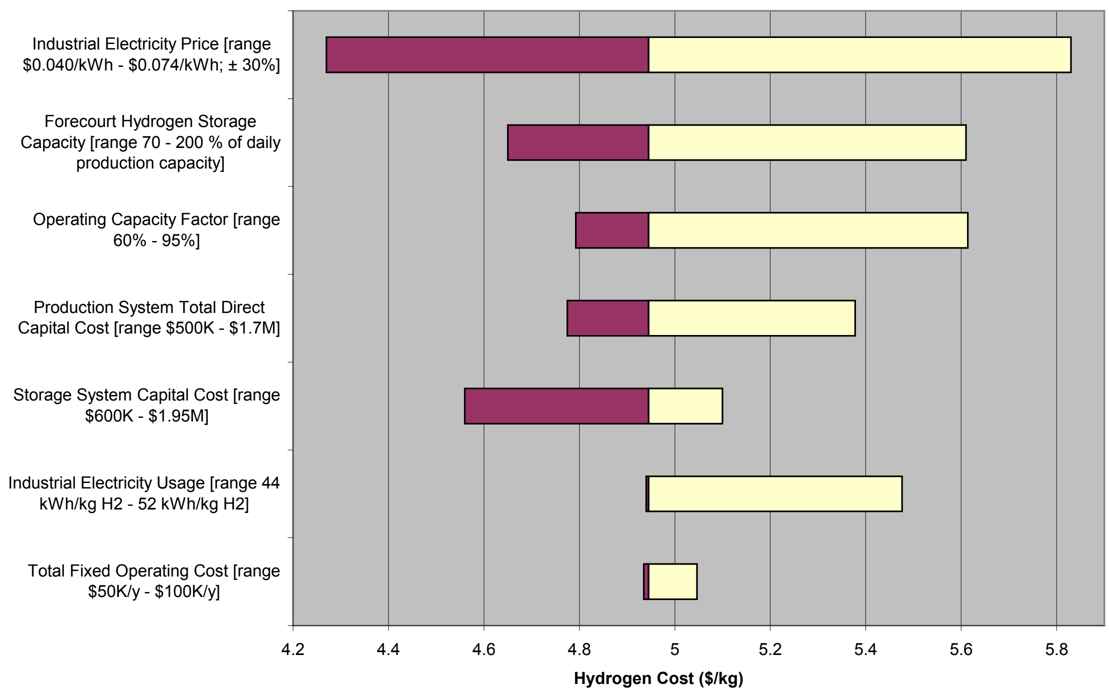


Figure 38. Forecourt Future Grid Electrolysis: Sensitivity Analysis Results

Natural Gas

Process Description

The forecourt current natural gas technology case is based on a 20-atm conventional tube-in-shell steam methane reactor (SMR) with hydro-desulfurization pretreatment and PSA gas cleanup (Figure 39). The PSA is based on a 4-bed Batta cycle achieving 75% hydrogen recovery. The unit is assumed to be factory built (as opposed to onsite construction) and is skid mounted for easy and rapid installation. A single 1,500-kg/day unit is assumed (as opposed to the previous H2A assumption of parallel 750-kg/day units.) The system is assumed to be air cooled (and thus requires no cooling water flow). The product hydrogen exits the PSA at 300 psi and is compressed for storage in metal cylinder storage tanks (2,500 psi maximum pressures). The hydrogen is next compressed to 6,250 psi (maximum) for transfer into a 4-bed high-pressure cascade system to allow rapid filling of 5,000-psi onboard vehicular hydrogen tanks.

The forecourt future natural gas technology case is based on a 20-atm integrated membrane stream reformer (reforming catalyst, water gas shift catalyst, and Pd-alloy membrane tubes integrated into a single vessel) (Figure 40). A 1:7.5 (by volume) admixture of reforming catalyst (Ni-Al-Ru at $150/kg, 2 g/cc) and water gas shift catalyst (Fe/Cr Ox at $7/kg, 1 g/cc) is assumed. Gas hourly specific space velocity (GHSV) of the reactor catalyst system is 1,344 per hour and is based on a 50% reduction of combined reformer/WGS catalyst volume compared with a non-integrated natural gas steam reforming configuration. The reactor vessel is based on a 4-pass annular heat exchange reformer configuration. Maximum process gas temperature is 550°C.

Natural gas and water are fed directly to the reactor at a 3:1 steam/C ratio without use of a pre-reformer. The membrane separator tubes are modeled as thin Pd-alloy layer supported on 1.27-cm diameter porous stainless steel support tubes. Hydrogen permeance is 527 scf/(hr•ft^2•$atm^{0.5}$). Overall membrane surface area is 34.1 ft^2 with a 90% hydrogen recovery. The required membrane surface area is calculated by a 1-D differential model based on a single pass, non-reacting chemistry and a membrane separator configuration wherein the reformate enters the membrane tubes and has hydrogen removed according to the permeance and differential hydrogen pressures across the membrane. Cost of the Pd-coated porous stainless-steel membrane tubes is estimated at $700/$ft^2$ to be consistent with the DOE target of $1,000/$ft^2$ for a complete standalone membrane system (tubes plus housing, manifolds, etc.).

Natural gas is the sole feedstock and is also used as a supplemental fuel to the burner. Flue gas is exhausted at 160°C as excessive waste heat is generated that cannot be used for recuperation. The unit is assumed to be factory built (as opposed to onsite construction) and is skid mounted for easy and rapid installation. A single 1,500-kg/day unit is assumed (as opposed to the previous H2A assumption of parallel 750-kg/day units.) The system is assumed to be air cooled (and thus requires no cooling water flow). The product hydrogen exits the reactor at 15 psi and, after cooling, is compressed to 300 psi before transfer to the refueling subsystem. The refueling subsystem further compresses the hydrogen for storage in metal cylinder storage tanks (2,500 psi maximum pressure) and then additionally compresses the hydrogen to 6,250 psi (maximum) for transfer into a 4-bed high-pressure cascade system to allow rapid filling of 5,000 psi onboard vehicular hydrogen tanks.

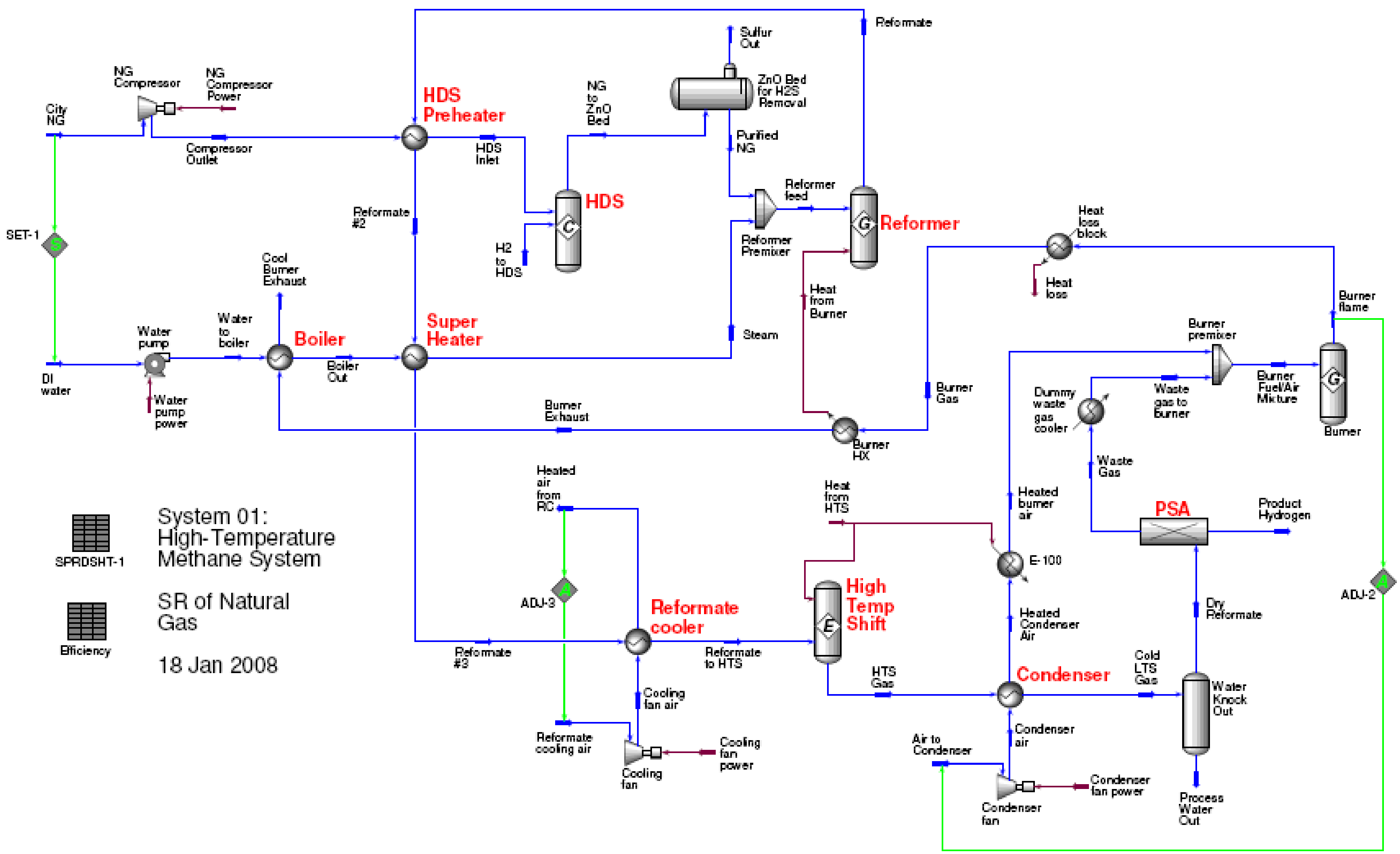


Figure 39. Process Flow Diagram—Current Natural Gas

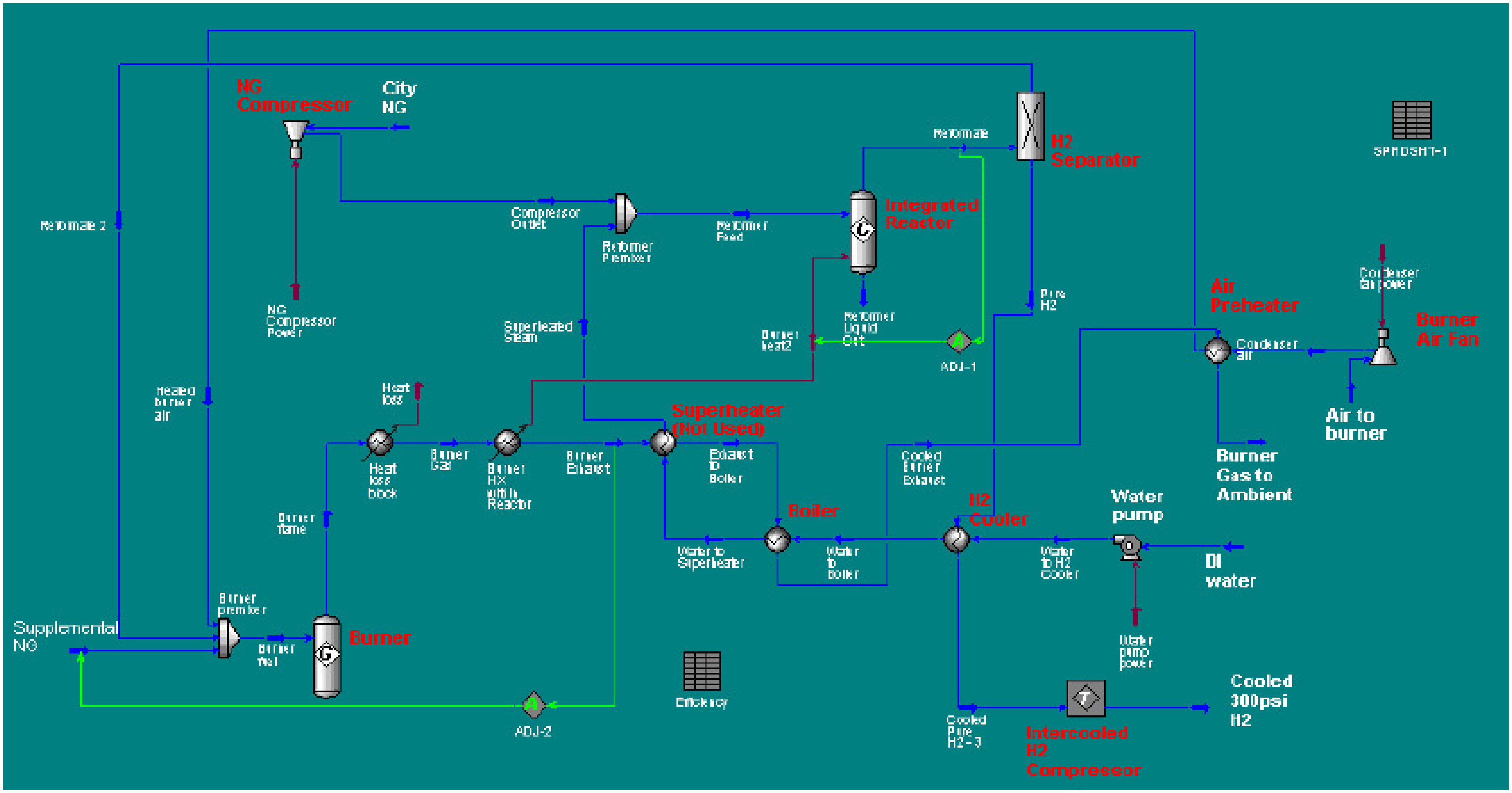


Figure 40. Process Flow Diagram—Future Natural Gas

H2A Model Inputs

The following tables show the H2A Model input values for the forecourt current and future natural gas technology cases. Default values that are common to all central cases are not shown in the tables below; see *Common Input Parameters and Processes* (page 16) for those values.

Table 58 summarizes the cost inputs, including capital, fixed operating, and variable operating costs; note that the values in this table are in units of thousand 2005$, unlike the central case cost input tables, which have units of million 2005$. Table 59 summarizes the energy inputs.

Table 58. Forecourt Natural Gas: Cost Input Summary

Capital Costs (thousand 2005$)		**Current**	**Future**
Initial direct production	Water feed system	0	0
	Primary feed system	10	10
	Burner feed system	—	0
	Boiler	30	10
	Superheater	10	—
	HDS & absorbent bed	10	—
	Burner	0	0
	Annular Ref-WGS-MS	—	150
	Reformer	200	—
	Water gas shift	170	—
	HDS preheater	0	—
	Primary air feed system	0	0
	Hydrogen cooler	—	10
	Reformate cooler	30	—
	Condenser	40	—
	Air feed system for condenser	0	—
	PSA unit	70	—
	Water purification	30	30
	Structural supports	20	20
	Controls system	40	40
	System assembly	210	160
	Miscellaneous	90	50
	Hydrogen compressor	—	230
	Total initial direct production[a]	*960*	*720*
Initial direct CSD	Compressors (3)	940	940
	Dispensers (2)	50	50
	Cascade storage (325 kg)	390	390
	Low-pressure storage (16 × 89 kg)	1,670	1,670
	Electrical upgrading	80	80
	Trenching for higher-voltage lines	120	120
	Total initial direct CSD[a]	*3,250*	*3,250*
	Total initial direct[a]	***4,200***	***3,970***
Indirect depreciable (production + CSD)	Site preparation	240	240
	Engineering and design	350	350
	Project contingency	210	200
	Up-front permitting	130	130
	Total indirect[a]	*930*	*920*

	Total initial (direct + indirect)[a]	***5,130***	***4,890***
Expected replacement (production + CSD)[b]		680	530
	Total[a]	**5,810**	**5,420**

Fixed Operating Costs (thousand 2005$/year)		**Current**	**Future**
Production + CSD	Licensing, permits, and fees	0	0
	Property taxes and insurance	90	90
	Rent	50	50
	Operating, maintenance, and repairs	90	80
	Labor	40	40
	Overhead and G&A	10	10
	Total[a]	**280**	**270**

Variable Operating Costs (thousand 2005$/year)[c]		**Current**	**Future**
Energy feedstocks, utilities, byproducts	Industrial natural gas	510	430
	Commercial electricity (production + CSD)	120	160
Other materials and byproducts	Process water	0	0
Other variable operating costs	Other variable operating[d]	0	0
	Total[2]	**630**	**590**

[a]Components might not add to total owing to rounding (all numbers rounded to the nearest $10,000; all values less than $10,000 are entered as zero).
[b]Sum of expected replacement capital costs over the 20-year plant life, adjusted to the year in which they are incurred using an NPV calculation.
[c]These costs vary over time in accordance with H2A Model price tables; startup year costs are shown here.
[d]Waste disposal costs, non-feedstock fuels, environmental surcharges, etc.

Table 59. Forecourt Natural Gas: Energy Input Summary

Feed	**Lower Heating Value**	**Material Use (per kg hydrogen)**	
		Current	**Future**
Industrial natural gas	36 MJ/Nm^3	4.5 Nm^3	4.2 Nm^3
Commercial electricity[a]	3.6 MJ/kWh	3.1 kWh	4.2 kWh

[a]Electricity used for production and compression.

H2A Model Results

The following tables and figures show the H2A Model results for the forecourt current and future natural gas technology cases. Table 60 summarizes the cost results. The forecourt current natural gas case produces hydrogen for $1.61/kg, with a total delivered hydrogen cost of $3.50/kg. The future case produces hydrogen for $1.59/kg, with a total delivered hydrogen cost of $3.47/kg. CSD capital costs are the largest expense for both cases; feedstock (natural gas) is the largest production expense. CSD costs account for more than half of the delivered hydrogen cost for both cases.

Table 60. Forecourt Natural Gas: Cost Results Summary

	Cost Component	Cost Contribution ($/kg hydrogen)		Percentage of Hydrogen Cost[a]	
		Current	Future	Current	Future
Production	Capital	0.45	0.32	13%	9%
	Fixed O&M	0.16	0.13	5%	4%
	Feedstock	0.91	0.96	26%	28%
	Other variable[b]	0.10	0.19	3%	5%
	Total production cost	*1.61*	*1.59*	*46%*	*46%*
CSD	Capital	1.26	1.26	36%	36%
	Fixed O&M	0.46	0.46	13%	13%
	Other variable[b]	0.16	0.16	5%	5%
	Total CSD cost	*1.88*	*1.88*	*54%*	*54%*
Total delivered hydrogen cost		**3.50**	**3.47**	**100%**	**100%**

[a]Total might not add to 100% owing to rounding.
[b]Including utilities.

Table 61 summarizes the process energy results. The energy inputs are in the form of natural gas feedstock and electricity used for hydrogen production and compression. The only energy output is hydrogen. The process energy efficiency (energy output divided by energy input) is 68.4% for the current case and 71.0% for the future case. These are process energy inputs only and do not include energy used upstream of the process or consumed in vehicles; thus they do not represent life-cycle or "well-to-wheel" energy use.

Table 61. Forecourt Natural Gas: Process Energy Results Summary

Energy Component	Energy Input (MJ per kg hydrogen)		Energy Output (MJ per kg hydrogen)	
	Current	Future	Current	Future
Industrial natural gas	164	154	—	—
Commercial electricity[a]	11	15	—	—
Hydrogen	—	—	120	120
Total	**175**	**169**	**120**	**120**

Process energy efficiency (LHV)[b]
Current = 68.4%
Future = 71.0%

[a]Electricity used for production and compression.
[b]Process energy efficiency = process energy output ÷ process energy input.

Table 62 (current) and Table 63 (future) show the values used in the sensitivity analyses. See *Sensitivity Analysis* (page 15) to learn how these values were chosen. Figure 41 (current) and Figure 42 (future) show the sensitivity analysis results. For both cases, forecourt hydrogen storage capacity has the largest effect on hydrogen price.

Table 62. Forecourt Current Natural Gas: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Feedstock conversion efficiency (%)	76	73	68
Total fixed operating costs (thousand $)	45	75	120
Storage system capital cost (thousand $)	1,000	1,665	2,200
Industrial natural gas feedstock price ($/Nm^3)	0.17	0.24	0.32
Production system total direct capital cost (thousand $)	600	957	2,800
Operating capacity factor	0.95	0.85	0.60
Forecourt hydrogen storage capacity (%)	70	120	200

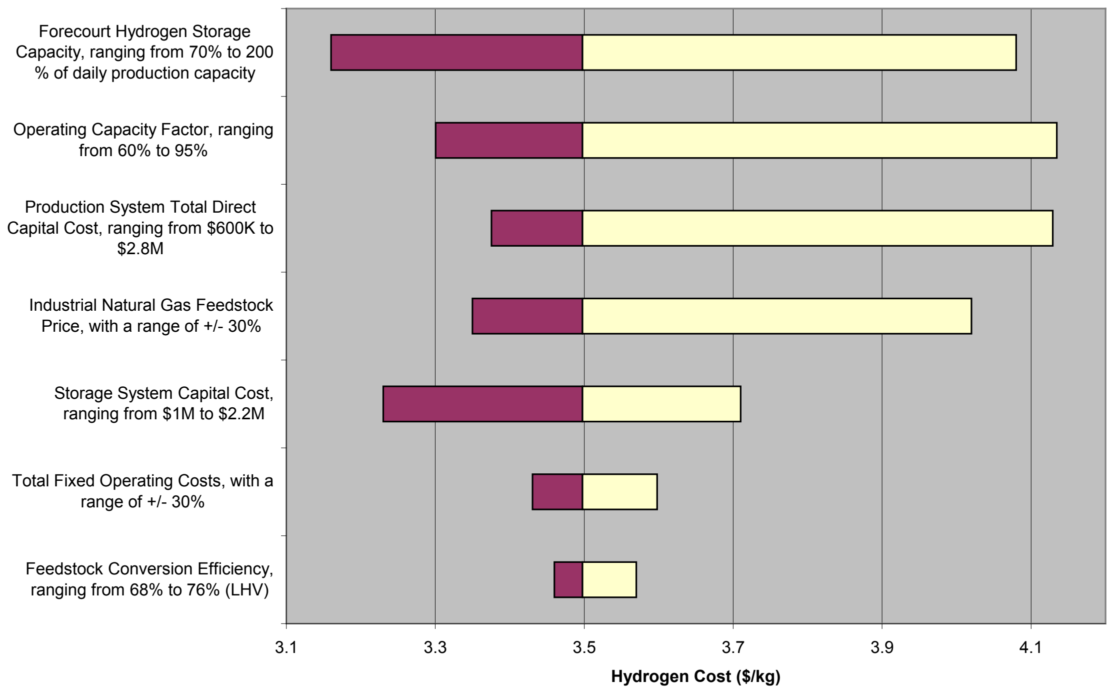


Figure 41. Forecourt Current Natural Gas: Sensitivity Analysis Results

Table 63. Forecourt Future Natural Gas: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Feedstock conversion efficiency (%)	80	78	70
Total fixed operating costs (thousand $)	35	58	93
Storage system capital cost (thousand $)	600	1,665	1,950
Industrial natural gas feedstock price ($/Nm^3)	0.15	0.22	0.29
Production system total direct capital cost (thousand $)	500	719	1,700
Operating capacity factor	0.95	0.85	0.60
Forecourt hydrogen storage capacity (%)	70	120	200

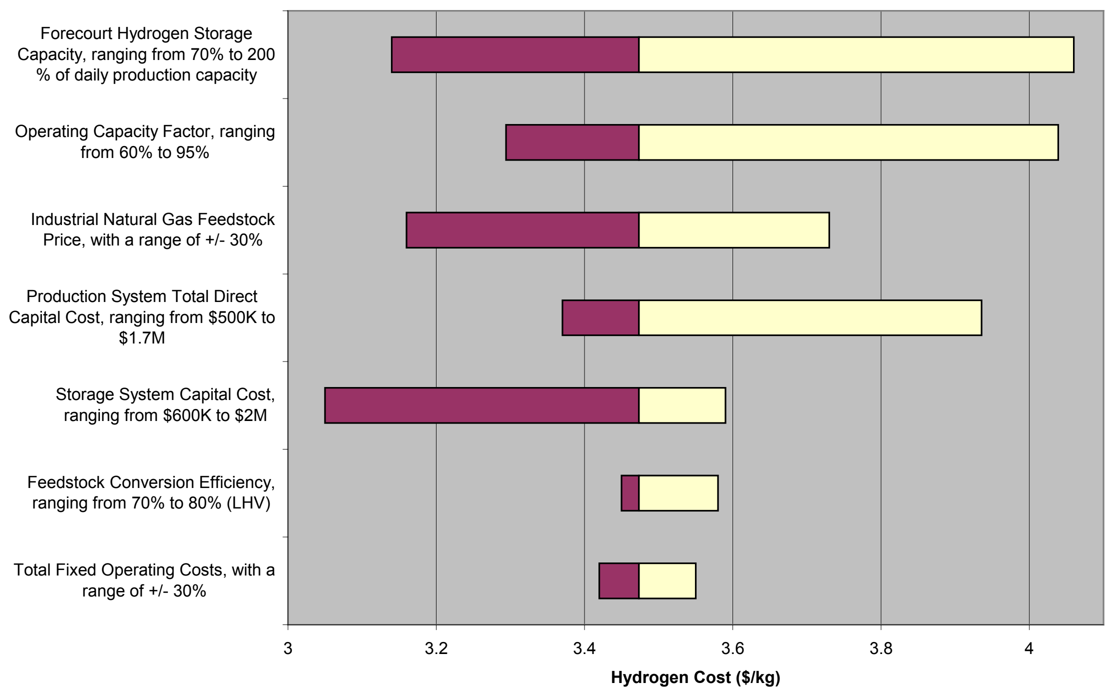


Figure 42. Forecourt Future Natural Gas: Sensitivity Analysis Results

Ethanol

Process Description

The forecourt current ethanol technology case is based on a 20-atm conventional tube-in-shell steam reactor (SR) with PSA gas cleanup (Figure 43). Precious metal catalyst is assumed. The catalyzed conversion of ethanol to methane is judged to occur rapidly to near-full ethanol conversion in a compact adiabatic reformer. Because methane is the primary component of the pre-reformer, the remainder of the system is nearly identical to that of a natural gas reformer system. Ethanol is the sole feedstock and is also used as a supplemental fuel to the burner. A 950°C burner adiabatic flame temperature and an 850°C reformer temperature are assumed. Flue gas is exhausted at 110°C. The PSA is based on a 4-bed Batta cycle achieving 75% hydrogen recovery.

The forecourt future ethanol technology case is based on a 20-atm integrated membrane stream reformer (reforming catalyst, water gas shift catalyst, and Pd-alloy membrane tubes integrated into a single vessel) (Figure 44). A 1:6.3 (by volume) admixture of reforming catalyst (Ni-Al-Ru at $150/kg, 2g/cc) and water gas shift catalyst (Fe/Cr Ox at $7/kg, 1 g/cc) is assumed. Gas hourly specific space velocity (GHSV) of the reactor catalyst system is 1,374 per hour (and is based on a 50% reduction of combined reformer/WGS catalyst volume compared with a non-integrated ethanol stream reforming configuration). The reactor vessel is based on a 4-pass annular heat exchange reformer configuration. Maximum process gas temperature is 500°C.

Ethanol and water are fed directly to the reactor at a 3:1 steam/C ratio without use of a pre-reformer. The membrane separator tubes are modeled as thin Pd-alloy layer supported on 1.27-cm diameter porous stainless steel support tubes. Hydrogen permeance is 527 scf/(hr•ft^2•atm$^{0.5}$). Overall membrane surface area is 46.8 ft^2 with a 90% hydrogen recovery. The required membrane surface area is calculated by 1-D differential model based on a single pass, non-reacting chemistry and membrane separator configuration wherein reformate enters the membrane tubes and has hydrogen removed according to the permeance and differential hydrogen pressures across the membrane. Cost of the Pd-coated porous stainless-steel membrane tubes is estimated at $700/ft^2 to be consistent with the DOE target of $1,000/ft^2 for a complete standalone membrane system (tubes plus housing, manifolds, etc.). Ethanol is the sole feedstock and is also used as a supplemental fuel to the burner. Flue gas is exhausted at 110°C.

For the current and future cases, the unit is assumed to be factory built (as opposed to onsite construction) and is skid mounted for easy and rapid installation. A single 1,500-kg/day unit is assumed (as opposed to the previous H2A assumption of parallel 750-kg/day units.) The system is assumed to be air cooled (and thus requires no cooling water flow). For the current case, the product hydrogen exits the PSA at 300 psi and is compressed for storage in metal cylinder storage tanks (2,500 psi maximum pressures). For the future case, the product hydrogen exits the reactor at 15 psi and, after cooling, is compressed to 300 psi before transfer to the refueling subsystem. The hydrogen is next compressed to 6,250 psi (maximum) for transfer into a 4-bed high-pressure cascade system to allow rapid filling of 5,000-psi onboard vehicular hydrogen tanks.

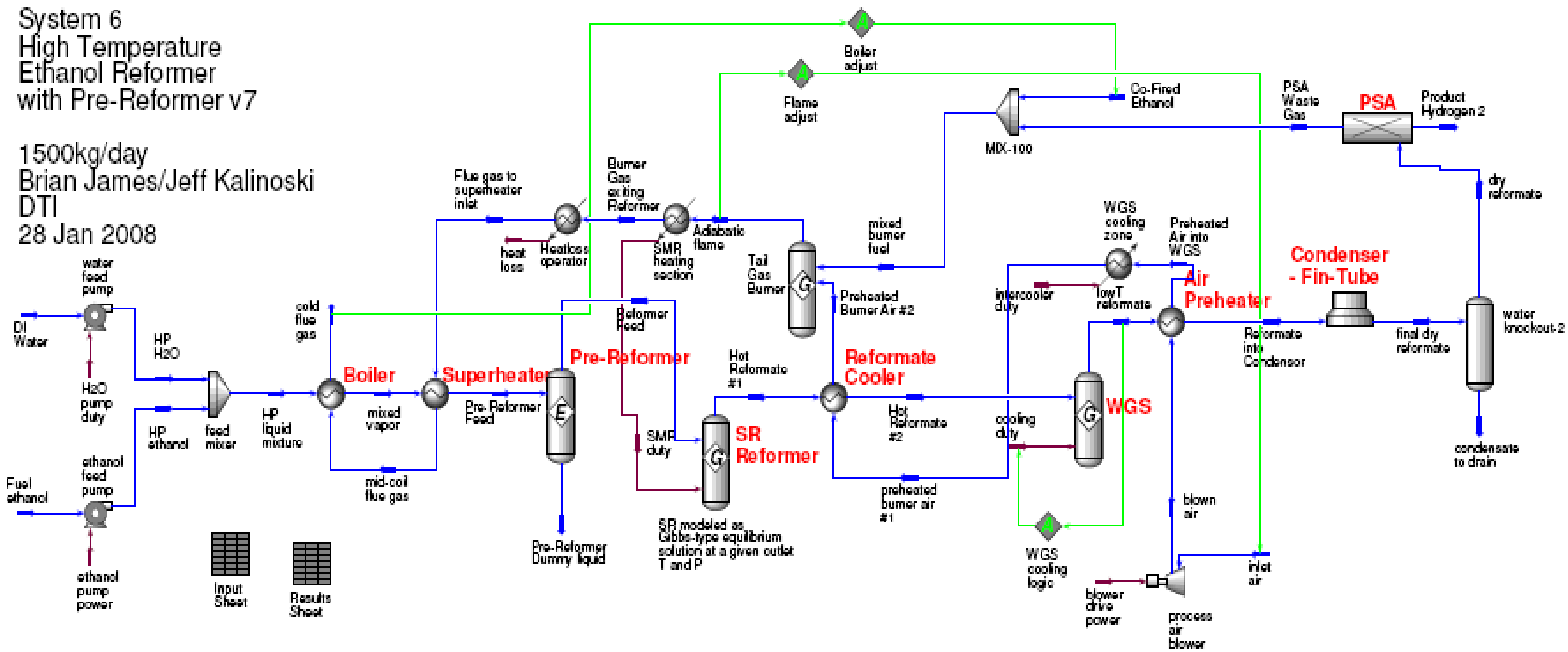


Figure 43. Process Flow Diagram—Current Ethanol

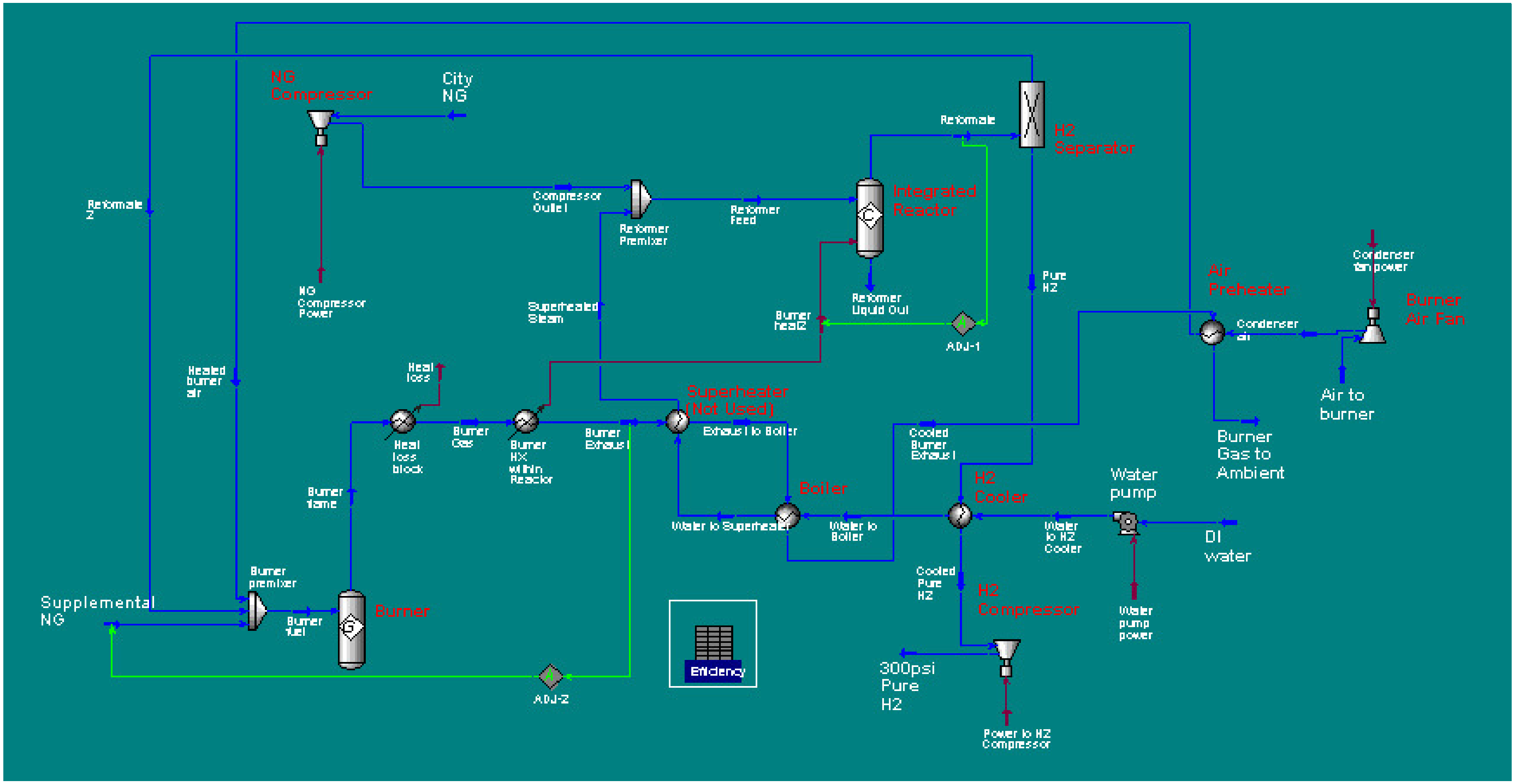


Figure 44. Process Flow Diagram—Future Ethanol

H2A Model Inputs

The following tables show the H2A Model input values for the forecourt current and future ethanol technology cases. Default values that are common to all central cases are not shown in the tables below; see *Common Input Parameters and Processes* (page 16) for those values.

Table 64 summarizes the cost inputs, including capital, fixed operating, and variable operating costs; note that the values in this table are in units of thousand 2005$, unlike the central case cost input tables, which have units of million 2005$. Table 65 summarizes the energy inputs.

Table 64. Forecourt Ethanol: Cost Input Summary

Capital Costs (thousand 2005$)		**Current**	**Future**
Initial direct production	Water feed system	0	0
	Primary feed system	0	0
	Burner feed system	0	0
	Boiler	40	50
	Superheater	10	10
	Pre-reformer	20	—
	Burner	0	0
	Annular Ref-WGS-MS	—	190
	Reformer	240	—
	Water gas shift reactor	190	—
	Air preheater	20	20
	Primary air feed system	0	0
	Reformate cooler	40	—
	Condenser	40	—
	Air feed system for condenser	0	—
	PSA unit	70	—
	Water purification	30	30
	Structural supports	20	20
	Controls system	40	40
	System assembly	210	210
	Miscellaneous	100	60
	Hydrogen compressor	—	230
	Ethanol underground storage tank	140	140
	Total initial direct production[a]	*1,210*	*1,010*
Initial direct CSD	Compressors (3)	940	940
	Dispensers (2)	50	50
	Cascade storage (325 kg)	390	390
	Low-pressure storage (16 × 89 kg)	1,670	1,670
	Electrical upgrading	80	80
	Trenching for higher-voltage lines	120	120
	Total initial direct CSD[a]	*3,250*	*3,250*
	Total initial direct[a]	***4,460***	***4,260***
Indirect depreciable (production + CSD)	Site preparation	240	240
	Engineering and design	350	350
	Project contingency	220	210
	Up-front permitting	130	130
	Total indirect[a]	*930*	*920*

		Current	Future
	Total initial (direct + indirect)[a]	***5,390***	***5,180***
Expected replacement (production + CSD)[b]		740	610
	Total[a]	**6,130**	**5,790**

Fixed Operating Costs (thousand 2005$/year)		**Current**	**Future**
Production + CSD	Licensing, permits, and fees	0	0
	Property taxes and insurance	100	90
	Rent	50	50
	Operating, maintenance, and repairs	100	90
	Labor	40	40
	Overhead and G&A	10	10
	Total[a]	**300**	**280**

Variable Operating Costs (thousand 2005$/year)[c]		**Current**	**Future**
Energy feedstocks, utilities, byproducts	Ethanol	1,090	980
	Commercial electricity (production + CSD)	90	150
Other materials and byproducts	Process water	10	10
Other variable operating costs	Other variable operating[d]	0	0
	Total[a]	**1,190**	**1,140**

[a]Components might not add to total owing to rounding (all numbers rounded to the nearest $10,000; all values less than $10,000 are entered as zero).
[b]Sum of expected replacement capital costs over the 20-year plant life, adjusted to the year in which they are incurred using an NPV calculation.
[c]These costs vary over time in accordance with H2A Model price tables; startup year costs are shown here.
[d]Waste disposal costs, non-feedstock fuels, environmental surcharges, etc.

Table 65. Forecourt Ethanol: Energy Input Summary

Feed	Lower Heating Value	Material Use (per kg hydrogen)	
		Current	Future
Ethanol	0.076 mmBtu/gal	2.2 gal	2.0 gal
Commercial electricity[a]	3.6 MJ/kWh	2.5 kWh	3.9 kWh

[a]Electricity used for production and compression.

H2A Model Results

The following tables and figures show the H2A Model results for the forecourt current and future ethanol technology cases. Table 66 summarizes the cost results. The forecourt current ethanol case produces hydrogen for $3.18/kg, with a total delivered hydrogen cost of $5.07/kg. The future case produces hydrogen for $2.91/kg, with a total delivered hydrogen cost of $4.79/kg. Ethanol feedstock is the largest expense for both cases. CSD costs account for more than one-third of the delivered hydrogen cost for both cases.

Table 66. Forecourt Ethanol: Cost Results Summary

	Cost Component	Cost Contribution ($/kg hydrogen)		Percentage of Hydrogen Cost[a]	
		Current	Future	Current	Future
Production	Capital	0.58	0.46	11%	10%
	Fixed O&M	0.20	0.17	4%	4%
	Feedstock	2.34	2.11	46%	44%
	Other variable[b]	0.05	0.17	1%	4%
	Total production cost	*3.18*	*2.91*	*63%*	*61%*
CSD	Capital	1.26	1.26	25%	26%
	Fixed O&M	0.46	0.46	9%	10%
	Other variable[b]	0.16	0.16	3%	3%
	Total CSD cost	*1.88*	*1.88*	*37%*	*39%*
Total delivered hydrogen cost		**5.07**	**4.79**	**100%**	**100%**

[a]Total might not add to 100% owing to rounding.
[b]Including utilities.

Table 67 summarizes the process energy results. The energy inputs are in the form of ethanol feedstock and electricity used for hydrogen production and compression. The only energy output is hydrogen. The process energy efficiency (energy output divided by energy input) is 64.9% for the current case and 69.5% for the future case. These are process energy inputs only and do not include energy used upstream of the process or consumed in vehicles; thus they do not represent lifecycle or "well-to-wheel" energy use.

Table 67. Forecourt Ethanol: Process Energy Results Summary

Energy Component	Energy Input (MJ per kg hydrogen)		Energy Output (MJ per kg hydrogen)	
	Current	Future	Current	Future
Ethanol	176	159	—	—
Commercial electricity[a]	9	14	—	—
Hydrogen	—	—	120	120
Total	**185**	**173**	**120**	**120**

Process energy efficiency (LHV)[b]
Current = 64.9%
Future = 69.5%

[a]Electricity used for production and compression.
[b]Process energy efficiency = process energy output ÷ process energy input.

Table 68 (current) and Table 69 (future) show the values used in the sensitivity analyses. See *Sensitivity Analysis* (page 15) to learn how these values were chosen. Figure 45 (current) and Figure 46 (future) show the sensitivity analysis results. For both cases, ethanol price has the largest effect on hydrogen price.

Table 68. Forecourt Current Ethanol: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Total fixed operating cost (thousand $)	50	93	130
Ethanol use (gal/kg hydrogen)	2.0	2.2	2.3
Storage system capital cost (thousand $)	1,000	1,665	2,200
Operating capacity factor	0.95	0.85	0.6
Low-pressure storage excess capacity (% of design capacity)	11	58	150
Production system total direct capital cost (thousand $)	700	1,207	3,000
Ethanol price ($/gal)	0.75	1.07	1.39

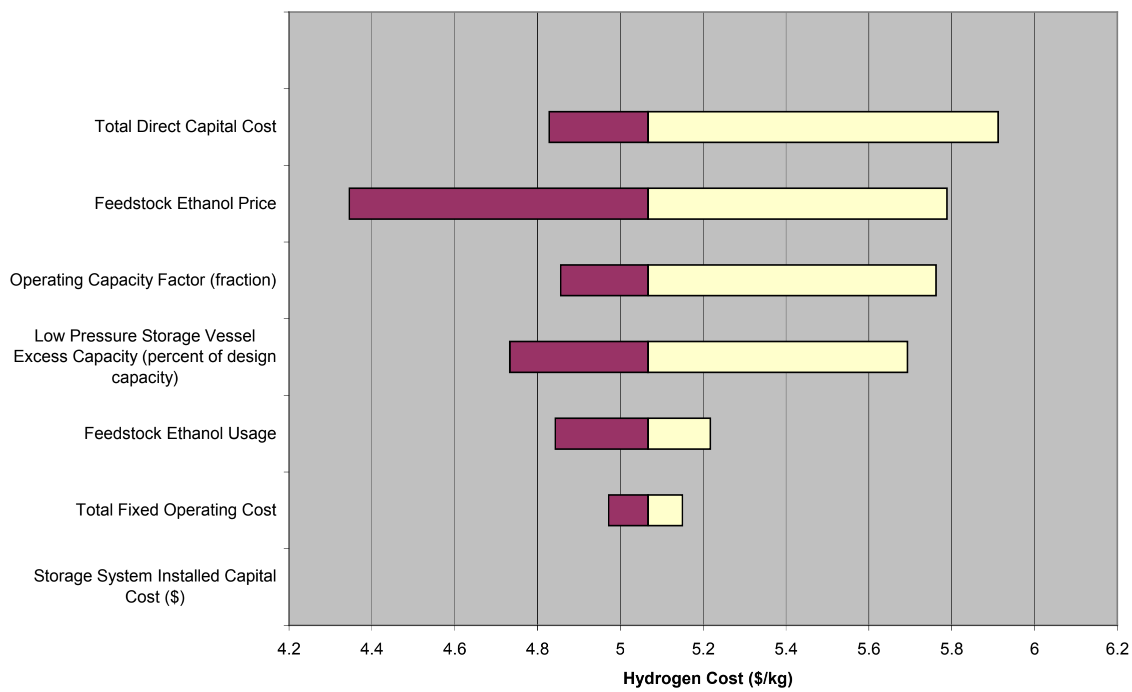


Figure 45. Forecourt Current Ethanol: Sensitivity Analysis Results

Table 69. Forecourt Future Ethanol: Sensitivity Analysis Values

Variable	Lower Value	Nominal Value	Upper Value
Total fixed operating cost (thousand $)	40	79	110
Ethanol use (gal/kg hydrogen)	1.9	2.0	2.1
Storage system capital cost (thousand $)	600	1,012	1,950
Operating capacity factor	0.95	0.85	0.6
Low-pressure storage excess capacity (% of design capacity)	11	58	150
Production system total direct capital cost (thousand $)	500	1,012	1,800
Ethanol price ($/gal)	0.75	1.07	1.39

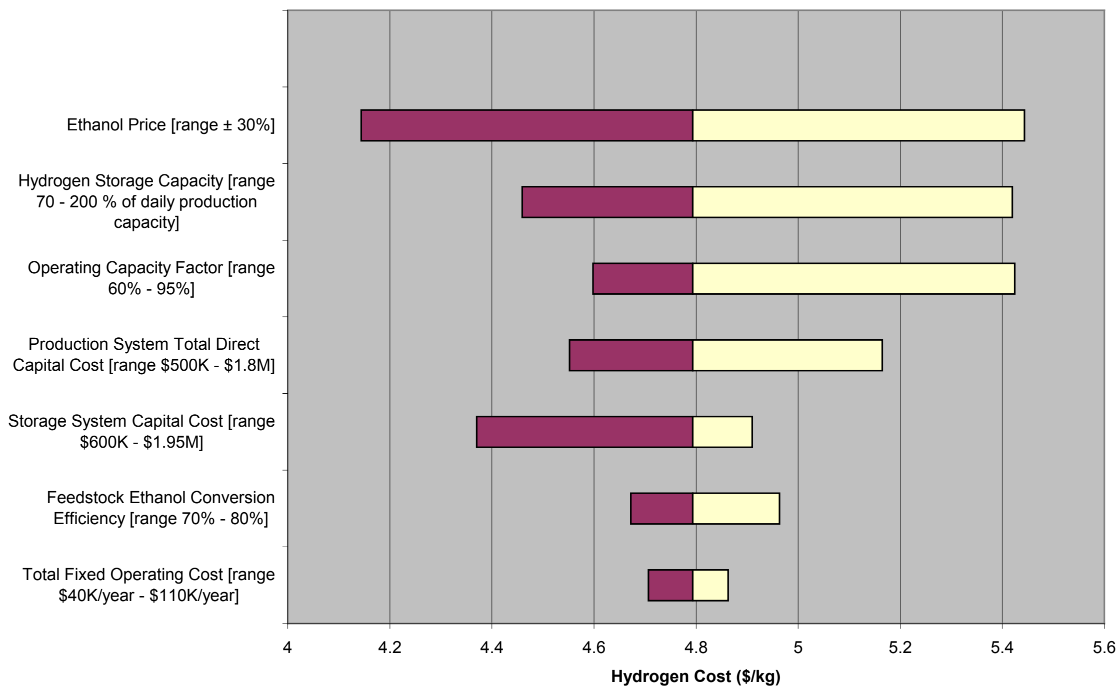


Figure 46. Forecourt Future Ethanol: Sensitivity Analysis Results

Sources

This report draws from the following two sources. Each H2A technology case contains additional references specific to the technology; download the technology cases for detailed information.

Steward, D.; Ramsden, T.; Zuboy, J. *H2A Production Model, Version 2 User Guide*. NREL/TP-560-43983. Golden, CO: National Renewable Energy Laboratory, September 2008. www.nrel.gov/docs/fy08osti/43983.pdf.

U.S. Department of Energy. H2A Production Analysis Web Site and Case Studies. Washington, DC: U.S. Department of Energy Hydrogen Program, 2008. www.hydrogen.energy.gov/h2a_production.html.

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