Final Report: Hydrogen Production Pathways Cost Analysis (2013 – 2016)

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STRATEGIC ANALYSIS

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Table of Abbreviations

ANL	Argonne National Laboratory
AFV	Alternative Fuel Vehicle
BOP	Balance of Plant
CH_4	Methane
CO	Carbon Monoxide
CO_2	Carbon Dioxide
CS	Corn Stover
DFMA	Design for Manufacturing and Assembly
DOE	US Department of Energy
FC	Fuel Cell
FCE	Fuel Cell Energy Inc.
FCS	Fuel Cell System
H_2	Hydrogen
H2A	Hydrogen Analysis Production Model
HHV	Higher Heating Value
HT	High Temperature
kWh	Kilowatt Hours
LHV	Lower Heating Value
LT	Low Temperature
MCFC	Molten Carbonate Fuel Cell
MEA	Membrane Electrode Assemblies
MPR	Monolithic Piston-type Reactor
MYPP	Multi-Year Production Plan
NG	Natural Gas
NREL	National Renewable Energy Laboratory
P&D	Production and Delivery
PEM	Proton Exchange Membrane
SMR	Steam Methane Reformer
SOEC	Solid Oxide Electrolysis Cell
SOFC	Solid Oxide Fuel Cell
SR	Steam Reforming
SA	Strategic Analysis, Inc.
TEA	Techno-Economic Analysis
TRL	Technology Readiness Level
WGS	Water Gas Shift
WWS	Waste Water Systems

Table of Contents

Table of Figures ... 6

Table of Tables ... 7

1. Executive Summary ... 8
  - 1.1. Overview ... 8
  - 1.2. Summary of the H_2 Production Technologies Analyzed ... 9
    - 1.2.1. Proton Exchange Membrane (PEM) Electrolysis ... 9
    - 1.2.2. Solid Oxide Electrolysis Cell (SOEC) ... 9
    - 1.2.3. Dark Fermentation of Biomass ... 10
    - 1.2.4. Reformation of Bio-oil in Monolithic Piston-Type Reactors ... 10
    - 1.2.5. Reformation of Natural Gas in a Reformer-Electrolyzer-Purifier (REP) ... 10
  - 1.3. Results Summary ... 10

2. Project Scope and Tasks ... 12
  - 2.1. Project Objectives and Accomplishments ... 12
  - 2.2. Project Scope and Background ... 12
  - 2.3. Project Tasks ... 12
    - 2.3.1. Task 1: Cost Model Validation Case Study ... 12
    - 2.3.2. Task 2: Define Pathways and Gather Information ... 13
    - 2.3.3. Task 3: Create a Draft H2A Case Study ... 13
    - 2.3.4. Task 4: Vet Case Study Assumptions and Results ... 13
    - 2.3.5. Task 5: Case Study Documentation and Reporting ... 13

3. Proton Exchange Membrane Electrolysis ... 13
  - 3.1. Case Study Summary ... 13
    - 3.1.1. PEM System Description ... 14
    - 3.1.2. Inputs ... 14
    - 3.1.3. Results & Sensitivity ... 16
    - 3.1.4. Discussion Points With Regards to PEM ... 20

4. High Temperature Solid Oxide Electrolysis ... 20
  - 4.1. Case Study Summary ... 20
  - 4.2. SOEC System Description ... 20
  - 4.3. Baseline Input Parameters ... 23
  - 4.4. Results and Sensitivity ... 25

4.5. Discussion Points for SOEC ... 27
5. H_2 Production via Fermentation of Biomass ... 27
5.1. Case Study Summary ... 27
5.2. Dark Fermentation System Description ... 28
5.3. Baseline Input Parameters ... 31
5.4. Results and Sensitivity ... 32
5.5. Discussion Points for Dark Fermentation ... 34
6. H_2 Production from Steam Reforming of Pyrolysis in Monolithic Piston-Type Reactors ... 34
6.1. Case Study Summary ... 34
7. H_2 Production via a Reformer-Electrolyzer-Purifier ... 35
7.1. Case Study Summary ... 35
7.2. Analytical Basis ... 36
7.3. Baseline Cost Projection Results ... 40
7.4. Results Summary: Cost Breakdown Plots and Tornado Sensitivity Charts ... 42
7.5. Monte Carlo Analysis Details ... 44
7.6. REP Alternate Case for Very Low-Cost H_2 for non-Fuel Cell Vehicle Applications ... 46
7.7. Projected Future Case ... 47
7.8. Special Considerations ... 47
8. H2A Production Pathways Future Work ... 47
8.1. Research Directions ... 47
8.2. n^{th} Unit Research ... 47
9. Appendix: Project Publications and Presentations ... 49

Table of Figures

Figure 1 - H_2 Production Cost Summary for H_2 generation methods analyzed in this project. 11
Figure 2 - Generalized PEM Electrolyzer System 14
Figure 3 - PEM electrolysis H_2 production cost contributions (2007$/kg) for four case studies 17
Figure 4 - Tornado chart for *Projected Current* Forecourt PEM Electrolysis case. 18
Figure 5 - Tornado chart for *Projected Future* Forecourt PEM Electrolysis case. 18
Figure 6 - Tornado chart for *Projected Current* Central PEM Electrolysis case.. 19
Figure 7 - Tornado chart for *Projected Future* Central PEM Electrolysis case. 19
Figure 8 - *Projected Current* SOEC Baseline Design 22
Figure 9 - *Projected Future* SOEC Baseline Design 22
Figure 10 - Tornado chart showing parameter sensitivities for the *Projected Current* SOEC case............ 26
Figure 11 - Tornado chart showing parameter sensitivities for the *Projected Future* SOEC case 26
Figure 12 – Process flow diagram for fermentation at a central production facility................................. 29
Figure 13 - Optimization Curves for Fermentation Cases 30
Figure 14 - Tornado chart for the *Projected Future* central fermentation case. 33
Figure 15 - Sample drawing of monolith with catalyst coating and sorbent filled channels...................... 35
Figure 16 - Simplified Block Flow Diagram for FCE's Standalone system 37
Figure 17 - Simplified Block Flow Diagram for FCE's Integrated system.. 37
Figure 18 - Model of anode and cathode reactions with $CO3^{=}$ transport in REP. 39
Figure 19 - Cost breakdown for both REP cases. 41
Figure 20 - Tornado chart for single parameter sensitivity study of REP Standalone case study 43
Figure 21 - Tornado chart for single parameter sensitivity study of REP Integrated case study................ 44
Figure 22 - Monte Carlo results for H_2 production in the Standalone REP system.................................... 45
Figure 23 - Monte Carlo results for H2 production in the Integrated REP system 45

Table of Tables

Table 1 - H_2 production high-volume cost projections for the PEM Electrolysis cases 14
Table 2 - Input parameters for H2A Production cases for PEM electrolysis.. 15
Table 3 - H_2 production cost breakdowns in 2007$/kg H_2 for PEM electrolysis baseline cases 16
Table 4 - H2 Production High-Volume Cost Projections for the SOEC Cases.. 20
Table 5 - Input Parameters for SOEC H2A Central Production Baseline Cases.. 24
Table 6 - H_2 Production Cost Breakdowns in 2007$/kg H_2 for SOEC Baseline Cases................................. 25
Table 7 - Cost projections for hydrogen production from a centralized dark fermentation facility. 28
Table 8 - Input and key parameters for production cases for fermentation of corn stover 32
Table 9 - H_2 Production cost breakdowns in 2007$/kg H_2 for Fermentation baseline cases 33
Table 10 - H_2 Production Low-Volume Cost Projections for the REP Cases... 36
Table 11 - Key system parameters for REP case studies.. 40
Table 12 - Cost results for both Projected Current and Projected Future cases 41
Table 13 - Parameters varied in sensitivity studies and Monte Carlo analysis for Standalone case study 42
Table 14 - Parameters varied in sensitivity studies and Monte Carlo analysis for Integrated case study . 42
Table 15 - Cost breakdown of H_2 produced via REP without PSA purification 46

1. Executive Summary

1.1. Overview

This report summarizes work conducted under a three year Department of Energy (DOE) funded project to Strategic Analysis, Inc. (SA) to analyze multiple hydrogen (H_2) production technologies and project their corresponding levelized production cost of H_2. The analysis was conducted using the H2A Hydrogen Analysis Tool developed by the DOE and National Renewable Energy Laboratory (NREL). The project was led by SA but conducted in close collaboration with the NREL and Argonne National Laboratory (ANL). In-depth techno-economic analysis (TEA) of five different H_2 production methods was conducted. These TEAs developed projections for capital costs, fuel/feedstock usage, energy usage, indirect capital costs, land usage, labor requirements, and other parameters, for each H_2 production pathway, and use the resulting cost and system parameters as inputs into the H2A discounted cash flow model to project the production cost of H_2 ($\$/kgH_2$).

Five technologies were analyzed as part of the project and are summarized in this report:

- Proton Exchange Membrane technology (PEM),
- High temperature solid oxide electrolysis cell technology (SOEC),
- Dark fermentation of biomass for H_2 production,
- H_2 production via Monolithic Piston-Type Reactors with rapid swing reforming and regeneration reactions, and
- Reformer-Electrolyzer-Purifier (REP) technology developed by Fuel Cell Energy, Inc. (FCE)

Each production technology is briefly summarized below in the Executive Summary and further details are provided in sections 3-7 of the report. Section 2 overviews the project scope and contractual tasks.

Two production scales, Forecourt and Central, were generally considered for each H_2 production pathway, with one or both being selected for cost analysis based on technology suitability. Forecourt production is used for facilities that produce approximately 1,500 kg H_2/day, while central production is used for facilities that produce 50,000 kg H_2/day.[1]

Each technology is assessed at one or more development stages: Existing, Current, and Future.

- **Existing Case**: Case studies of Existing H_2 production are based on a system (or key technology) that has been demonstrated using process parameters and material costs used today at their current levels of production (typically only a few a year, perhaps none at all). These may be lab-scale or prototype systems and typically represent an upper bound on the cost to produce H_2.
- **Current Case**: Case studies of Current H_2 production costs are based on systems (or key technology) using process parameters demonstrated and used in the Existing case, but with generally lower capital and material costs corresponding to projected high rate serial production (eg. hundreds of systems per year).
- **Future Case**: Future case studies forecast improved process parameters consistent with normal improvements in technology as time passes and reduced costs corresponding with nth plant

volumes. The Future case analyses take into account cost trends for various parameters (such as feedstock or energy costs) and are used to project H_2 costs at far future date (nominally 2025). Feedstock and energy costs use the Energy Information Administration's 2009 Annual Energy Outlook projections (out to 2070) and are further extrapolated into future years using Pacific Northwest National Laboratory's GCAM model for the discounted cash flow analysis.

One of two primary methods was used for each case study to comprehensively define the system and quantify the system parameters to be used within the H2A cash flow model. The first method, typically utilized for mid-to-high Technology Readiness Level (TRL) cases, began with solicitation of responses to a questionnaire from industrial and research experts on the given technology. These responses were then analyzed and discussed within the SA, NREL, and ANL teams. SA also developed robust process models to verify the values provided by the experts and to generate appropriate system parameters that were missing or in conflict with the provided data. The other method, used more commonly for low-TRL cases, was to gather any relevant information that was available from experts through less formalized processes (review of documents, teleconferences, and e-mail). This information then became the cornerstone on which detailed process models were developed. The process model results were used as H2A inputs to develop levelized H_2 costs. Regardless of chosen method, extensive literature searches were also used to determine or verify system parameters.

All system designs were optimized to the best knowledge of the SA/NREL/ANL team and the subject matter experts consulted. Optimization efforts specifically targeted meeting DOE's MYPP goal of H_2 production costs of less than \$2/kg H_2 (\$2/gge). However, H_2 production technology is rapidly advancing and periodic updates to the analyses will certainly be needed to capture the cost impact of improvements.

1.2. Summary of the H_2 Production Technologies Analyzed

1.2.1. Proton Exchange Membrane (PEM) Electrolysis

Low-temperature PEM electrolysis units catalytically convert water to hydrogen and oxygen under a current load. On the anode, water is split into protons and oxygen described by the half reaction $H_2O \rightarrow 2H^+ + ½O_2$. The protons cross through a selective membrane electrolyte and recombine on the cathode according to the cathode half reaction $2H^+ \rightarrow H_2$. Oxygen is generated on the anode side of the unit while hydrogen is generated on the cathode which can be used as fuel. PEM electrolysis units typically operate at moderate temperature (~60-80°C) and high current densities.

1.2.2. Solid Oxide Electrolysis Cell (SOEC)

Like the PEM electrolyzer, an SOEC also generates H_2 by electrolytic water splitting. However, SOEC units operate at much higher temperatures and split water (in the form of steam) initially into molecular H_2 and O^{2-} ions. The O^{2-} ions are transported across an electrolyte membrane (typically a solid, catalyst coated ceramic). H^+ cations reform into H_2 without the presence of oxygen and exit the unit to be used as fuel. The process requires the use of high temperature steam as a feed source and generally operates at temperatures in excess of 600°C. The high temperatures used in SOEC increases the electrical efficiency compared to other types of electrolysis units.

1.2.3. Dark Fermentation of Biomass

Dark fermentation is the process of producing H_2 as a fermentation byproduct using. Microbes have been genetically engineered to increase H_2 production. The complex process begins with thermal and chemical breakdown of biomass which converts cellulose and hemi-cellulose to sugars. The sugars can then be fermented by the aforementioned microbes to create H_2. The technology analyzed in this project used corn stover as the biomass for fermentation. The process requires large volumes of biomass and water to support the fermentation process and is combined with wastewater treatment facilities to handle effluent and recycle as much water as possible.

1.2.4. Reformation of Bio-oil in Monolithic Piston-Type Reactors

The Monolithic Piston Type Reactor process consists of a pair of reactors, each containing a catalyst coated TiO_2 monolith filled with a CO_2 sorbent. The reactors operate temporally out of sync, switching between steam reforming of pyrolysis oil and regeneration of the CO_2 sorbent. During reformation, steam and pyrolysis oil are fed to one of the reactors at ~600°C. Reforming occurs over the catalyzed surface to produce H_2, CO_2, and coke. The coke is a solid particle which coats the reactor interior. The CO_2 gas is adsorbed onto the composite sorbent during the reformation process, simultaneously increasing the H_2 concentration of the product gas and driving the reaction towards completion. The temperature of the reactor cools to ~500°C during the reforming process due to the endothermic nature of steam reforming. After approximately ten minutes, the reactor switches to “regeneration mode” wherein the fuel supply is cut off and the reactor is depressurized which causes the adsorbed CO_2 to be released. Air is then blown through the reactor to combust the coke, which has remained in the reactor. This air-coke combustion heats the reactor beyond the 600°C reforming temperature and the cycle is ready to be repeated.

1.2.5. Reformation of Natural Gas in a Reformer-Electrolyzer-Purifier (REP)

The Reformer-Electrolyzer-Purifier (REP) technology is a product of Fuel Cell Energy, Inc. and derives from the company’s existing Molten Carbonate Fuel Cell (MCFC) technology, which normally generates electricity for large applications. The REP is essentially a MCFC stack operated in reverse (i.e. electrolysis). In short, natural gas (NG) first undergoes steam methane reforming (SMR) in a separate SMR reactor and the reformed gas is then fed into the REP unit, where the CO_2 is effectively transported across the electrolyte, splitting a water molecule as part of the reaction. Thus H_2 is generated both from reforming of methane and also from water electrolysis. Because the carbon dioxide (CO_2) generated during SMR is pumped across the MCFC electrolyte, high concentration ($\geq$95%) H_2 exits the REP unit.

1.3. Results Summary

Results for four of the five H_2 production system studies are presented as a levelized cost of H_2 and appear as \$/kg H_2. The levelized costs are computed using DOE’s H2A Production tool, a discounted cash flow model, which uses feedstock and energy costs, thermal requirements, and capital and maintenance costs, as input parameters specific to each case. All costs are presented in 2007\$. **Figure 1** shows cost results for the four analysis projects for which cost results are available. The range of H_2 production costs for *Projected Current* case studies is \$2.58 - \$51.02/kg H_2. It is worth noting that the *Projected Current* fermentation hydrogen cost (\$51.02/kg H_2) represents a significant outlier within the data due to its low level of commercial readiness: without this particular data point, the range of costs for hydrogen production from the *Projected Current* cases narrows to \$2.58 - \$5.14/kg H_2.[2] The range of H_2 production costs for *Projected Future* case studies is \$3.82 - \$5.65/kg H_2.[3]

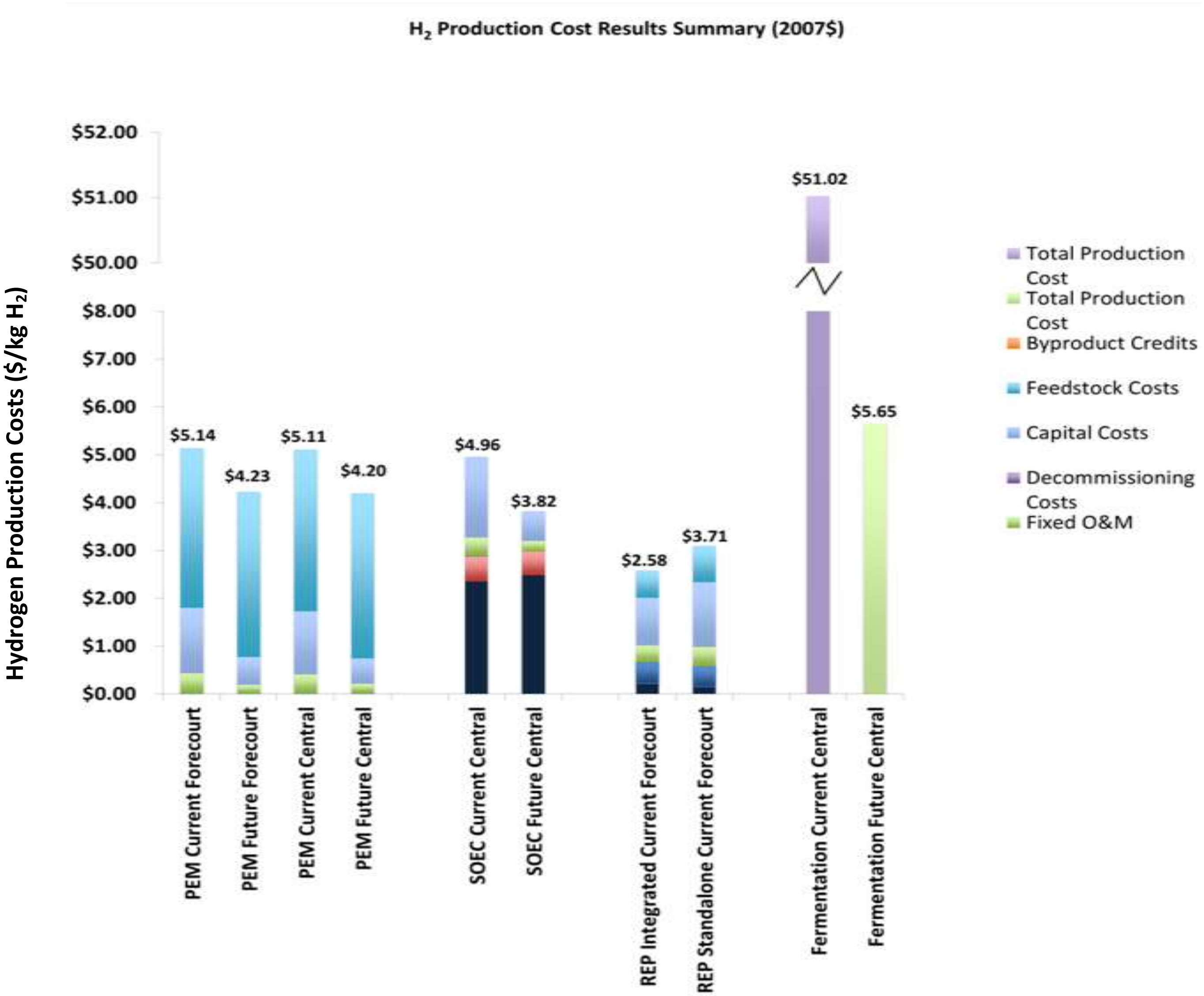

Figure 1 - H$_2$ Production Cost Summary for H$_2$ generation methods analyzed in this project. More details for each category are available below. (Fermentation costs are broken down in Section 5.)

2. Project Scope and Tasks

2.1. Project Objectives and Accomplishments

The objective of this project was to analyze a series of Hydrogen (H_2) production and delivery (P&D) pathways specified by the U.S. Department of Energy (DOE) for cost and performance. Highlights of accomplishments from this project include:

- Completed H_2 pathway analyses to determine the most economical, environmentally benign, and socially feasible paths forward for the production and delivery of H_2 fuel for fuel cell vehicles.
- Identified key "bottlenecks" to the success of the pathways, primary cost drivers, and remaining Research and Development (R&D) challenges.
- Created or updated Hydrogen Analysis (H2A) Case studies on selected P&D pathways (as identified by DOE) which represents transparent, publicly vetted, projections of current and future hydrogen production cost which describe assumptions, methodology, and results.

2.2. Project Scope and Background

System-level analyses of H_2 P&D technologies is needed to support selection of portfolio priorities through evaluations of technical progress and H_2 cost status, as well as projection of technology timelines and benefits, and evaluation of the potential of P&D pathways to meet the DOE Fuel Cell Technologies Office (FCTO) cost goal of <$4/gge delivered and dispensed H_2 by 2020. This effort includes annual cost analyses of key remaining challenges for technology pathways within the H_2 P&D sub-program portfolio using, primarily, the H2A[4] model to determine status improvements resulting from technology advancements, cost as a function of production volume, illustrate uncertainties in the cost estimates with error bars, and show potential for cost reductions based on sensitivity analyses. The pathways identified by DOE for this project included Proton Exchange Membrane Electrolysis, High Temperature Solid Oxide Electrolysis, H_2 Production via Dark Fermentation of Biomass, H_2 Production by Steam Reforming of Pyrolysis oil in Monolithic Piston-Type Reactors, and H_2 Production in a Reformer-Electrolyzer-Purifier.

2.3. Project Tasks

The following tasks were performed in order to complete the project. The validation task (Task 1: Cost Model Validation Case Study) was completed for one H2A case and not repeated for others. All other tasks were repeated for each system analyzed:

2.3.1. Task 1: Cost Model Validation Case Study

Using the processes described below in Tasks 2, 3, and 4, a H_2 P&D pathway production case was created and vetted. The Validation Case Study demonstrated that the proposed analysis methodology is adequate and the following H2A cases are accurate and provide comprehensive results.

2.3.2. Task 2: Define Pathways and Gather Information

As directed by DOE, H_2 P&D pathways were defined and developed. Work was completed to update or create new H2A cases. Information was gathered from national laboratories, universities, and industrial sources in conjunction with literature reviews. The gathered information was augmented with engineering analysis, modeled using Excel® spreadsheets or ASPEN HYSYS®, or other system performance analysis necessary to define the P&D pathway option to sufficient detail for full capture of all significant cost parameters. Collected information included what was necessary to fully populate H2A case study models and includes technology factors and financial and economic factors.

2.3.3. Task 3: Create a Draft H2A Case Study

H2A P&D models were populated with data collected in Task 2. All efforts to ensure quality control have been made. At a minimum, each case study includes a text description of the process, a list of references, a process flow diagram, and tornado and waterfall charts, as appropriate, with sensitivity ranges.

2.3.4. Task 4: Vet Case Study Assumptions and Results

Assumptions and results from the draft cases created in Task 4 were provided to all team members. The sources consulted for information in Task 2 were specifically tasked with confirming the baseline assumptions. Only once all issues were resolved, and each team member agreed that the results were accurate, was the case study transmitted to DOE for publication.

2.3.5. Task 5: Case Study Documentation and Reporting

Each case study was documented within the actual H2A production model sheet. Further, a DOE Record was created for each case study. At the request of DOE, written reports or presentations were provided to DOE and USDRIVE Technical Teams. Annual Merit Review presentations were given each year and completed with annual reports.

3. Proton Exchange Membrane Electrolysis

3.1. Case Study Summary

The study of hydrogen production via proton exchange membrane (PEM) electrolysis is based on four cases, a *Projected Current* and *Projected Future* case for both the Forecourt (500-1,500 kg/day) and Central (50,000 kg/day) models. The projected high volume untaxed cost of hydrogen production from PEM electrolysis ranges from ~$4 to $5.80/kg H_2, based on Hydrogen Analysis version 3 (H2A v3) model case study results. The baseline projections shown in the **Table 1** incorporate averages of the manufacturer-supplied electrolyzer stack and balance of plant (BOP) costs; while the Low and High Values are included to reflect an expected spread in uninstalled capital costs (with all other technoeconomic inputs the same as in the Baseline cases), as vetted by the manufacturers. The Baseline, Low, and High values in the table for the Forecourt cases represent a standard 1,500 kg/day production capacity; as a variation on the standard *Projected Current* Forecourt case, an analysis of early market stations (500 kg/day capacity) was performed with inputs from the manufacturers. The result is also included in **Table 1**.

Table 1 - H_2 production high-volume cost projections for the PEM Electrolysis cases[5]

Case Study	Low Value ($/kg H_2)	Baseline ($/kg H_2)	High Value ($/kg H_2)	Early Market[6] ($/kg H_2)
Forecourt: *Projected Current* Case[7]	$4.79	$5.14	$5.49	$5.79
Projected Future Case[8]	$4.08	$4.23	$4.37	-
Central: *Projected Current* Case[9]	$4.80	$5.12	$5.45	-
Projected Future Case[10]	$4.07	$4.20	$4.33	-

3.1.1. PEM System Description

PEM electrolysis is the process of splitting water by method of supplying an electrical current to a cell in which the anode and cathode are separated by a solid polymer electrolyte. Water is passed over the anode while the anode is being supplied with an electrical current. The current splits the water into H^+ and O^{2-} ions. The H^+ cations pass through the polymer electrolyte to the cathode. At the cathode, the H^+ reforms into diatomic hydrogen, H_2, and leaves the system to be further purified, as needed. Once desired purity is reached, the H_2 can then be used as fuel. Oxygen (O_2) likewise leaves the electrolysis cell from the anode side of the system where it may have been diluted with air (if a sweep gas is used). As such, no practical use for the O_2 is considered at this time.

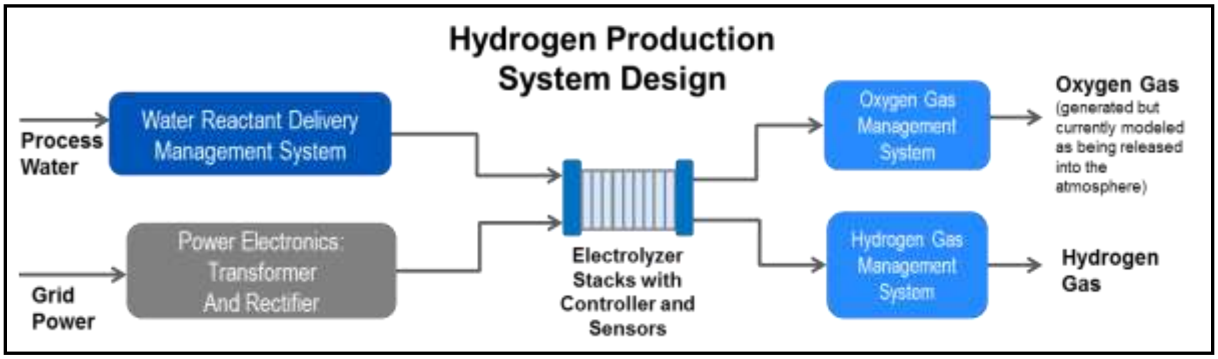

Figure 2 - Generalized PEM Electrolyzer System

3.1.2. Inputs

The major parameters used to develop the four H2A v3.0 baseline case studies are shown in Table 2 (all other H2A input parameters not cited in the table used standard H2A v3.0 default values[11]).

Table 2 - Input parameters for H2A Production cases for PEM electrolysis (costs in 2007\$[5] and in 2012\$[12])

Parameter	***Projected Current*** **Forecourt 1,500 kg/day**	***Projected Future*** **Forecourt 1,500 kg/day**	***Projected Current*** **Central 50,000 kg/day**	***Projected Future*** **Central 50,000 kg/day**
Plant Capacity (kg/day)	1,500	1,500	50,000	50,000
Total Uninstalled Capital (2012\$/kW)	\$940	\$450	\$900	\$400
Stack Capital Cost (2012\$/kW)	\$385[13]	\$171	\$423	\$148
Balance of Plant (BOP) Capital Cost (2012\$/kW)	\$555[14]	\$279	\$477	\$252
Total Electrical Usage kWh/kg Conversion Efficiency (LHV of H_2)	54.6[15] (61%)	50.3 (66%)	54.3 (61%)	50.2 (66%)
Stack Electrical Usage (kWh/kg) Conversion Efficiency (LHV of H_2)	49.2 (68%)	46.7 (71%)	49.2 (68%)	46.7 (71%)
BOP Electrical Usage (kWh/kg)	5.4	3.6	5.1	3.5
Electrolyzer Power Consumption at full power (MW)	3.4	3.1	113	105
Average Electricity Price over Life of Plant[16] (2007¢/kWh)	6.12	6.88	6.22	6.89
Electricity Price in Startup Year[17] (2007¢/kWh)	5.74	6.59	5.74	6.59
Outlet Pressure from Electrolyzer (psi)	450	1,000	450	1,000
Installation Cost (% of uninstalled capital cost)	12[18]	10	12	10
Replacement Interval (years)	7	10	7	10
Replacement Cost of Major Components (% of installed capital cost)	15	12	15	12
Plant Life (years)	20	20	40	40
Stack Current Density (mA/cm^2)	1,500	1,600	1,500	1,600
Capacity Factor (%)	86	86	97	97

3.1.3. Results & Sensitivity

The hydrogen production cost breakdown for the four H2A v3.0 PEM electrolysis baseline cases is shown in **Table 3**. These cases used inputs from the OEMs to determine the most likely parametric values for a given scale of the technology within a given timeframe. This is in contrast to the sensitivity analysis, which looks at the effects of deviations from those baseline inputs. As shown in the table, the primary cost driver for production is the electricity feedstock cost for the electrolysis. Although the electrolyzer electrical efficiency increases between the *Projected Current* and *Projected Future* cases (as seen in **Table 3**), the electricity price also rises (also shown in **Table 3**), and, as a result of this combined effect and other factors, electricity feedstock costs are slightly higher for the *Projected Future* versus *Projected Current* cases in **Table 3**.

Table 3 - H_2 production cost breakdowns in 2007$/kg H_2 for PEM electrolysis baseline cases

Component	*Projected Current* Forecourt 1,500 kg/day	*Projected Future* Forecourt 1,500 kg/day	*Projected Current* Central 50,000 kg/day	*Projected Future* Central 50,000 kg/day
Stack Capital Cost	$0.42	$0.16	$0.48	$0.17
BOP Capital Cost	$0.61	$0.25	$0.53	$0.26
Indirect Capital Cost and Replacement Cost	$0.32	$0.16	$0.32	$0.10
Decommissioning	$0.02	$0.01	$0.00	$0.00
Fixed operations and maintenance (O&M)	$0.42	$0.18	$0.40	$0.20
Electricity Feedstock	$3.34	$3.46	$3.38	$3.46
Variable O&M	$0.01	$0.01	$0.01	$0.01
Total H_2 Production Cost (2007$/kg H_2)	$5.14[19]	$4.23	$5.12	$4.20

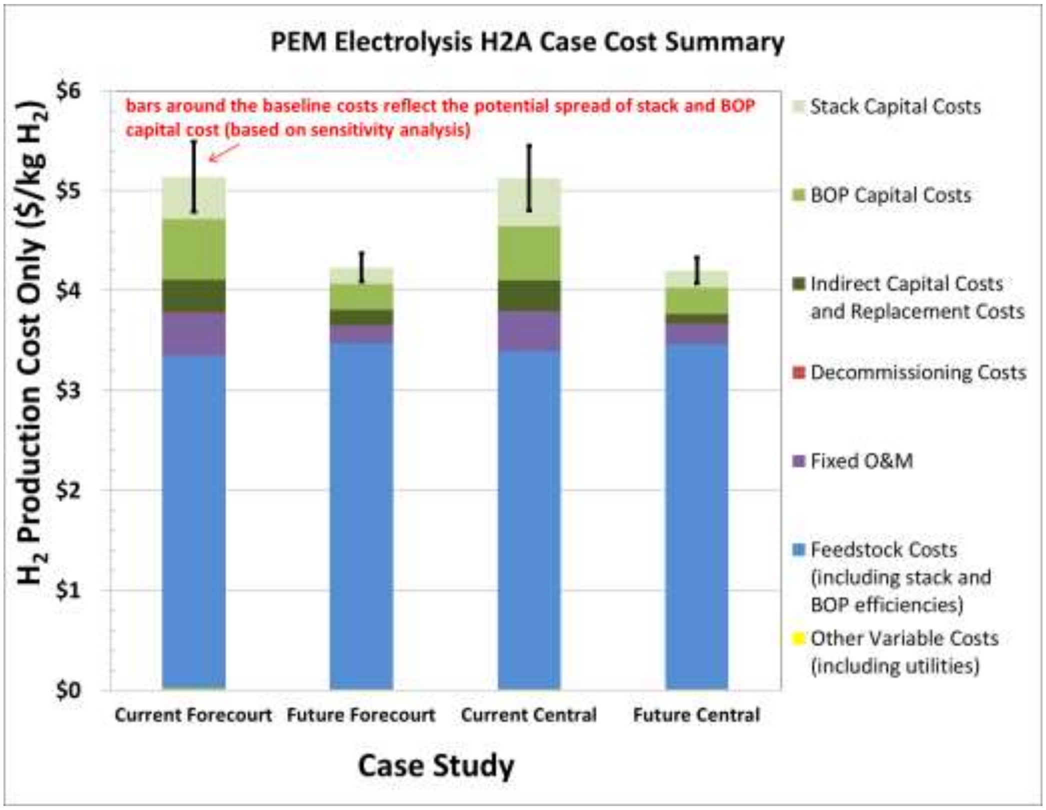

Figure 3 - PEM electrolysis H_2 production cost contributions (2007$/kg) for four case studies[20]

Tornado charts based on the parameter spreads provided by the electrolyzer companies were developed for the four cases to examine the impact of individual parameters on hydrogen cost in a single variable sensitivity analysis. These tornado charts, shown in **Figure 4** - **Figure 7**, plot the projected hydrogen cost variations on the x-axis against different single input parameters arranged along the y-axis. Specifically, the plots illustrate the H_2 production cost sensitivities to variations in: (1) average electricity price over life of plant; (2) electricity usage; (3) uninstalled capital cost; (4) site preparation cost; (5) stack replacement interval; and (6) stack replacement cost. Each tornado chart is organized from top to bottom to represent the most to least sensitive of the analyzed input parameters, respectively. The colored shading indicates either an increase (red) or a decrease (blue) in the baseline hydrogen cost from the change in input parameter. The data labels list the low and high values for the input parameters.

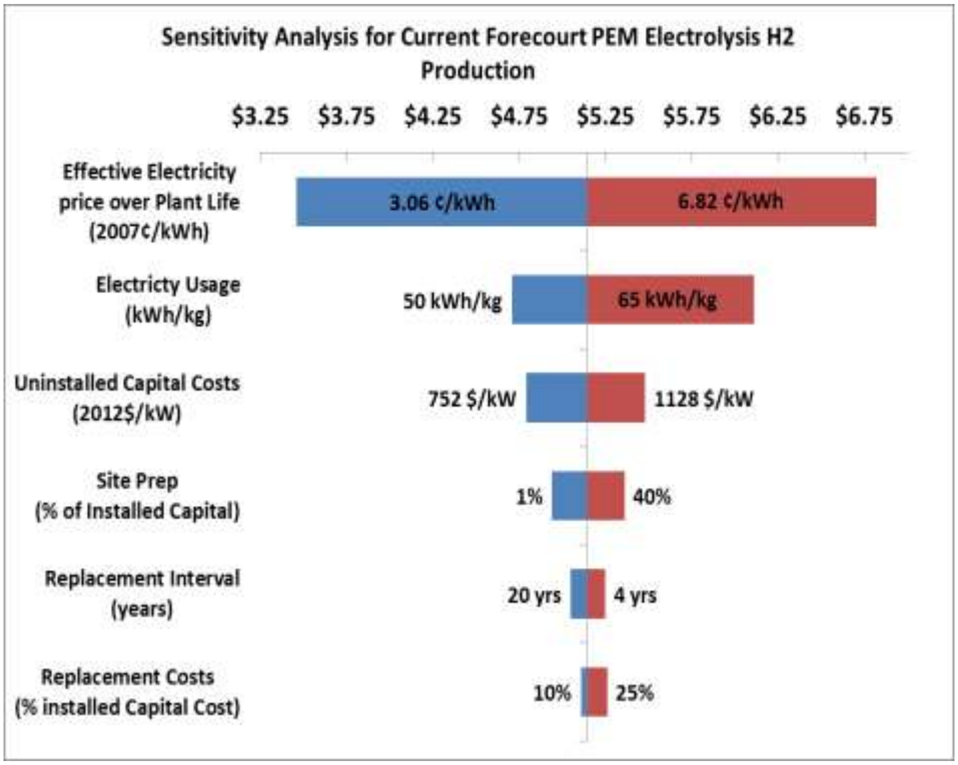

Figure 4 - Tornado chart showing parameter sensitivities for *Projected Current* Forecourt PEM Electrolysis case.

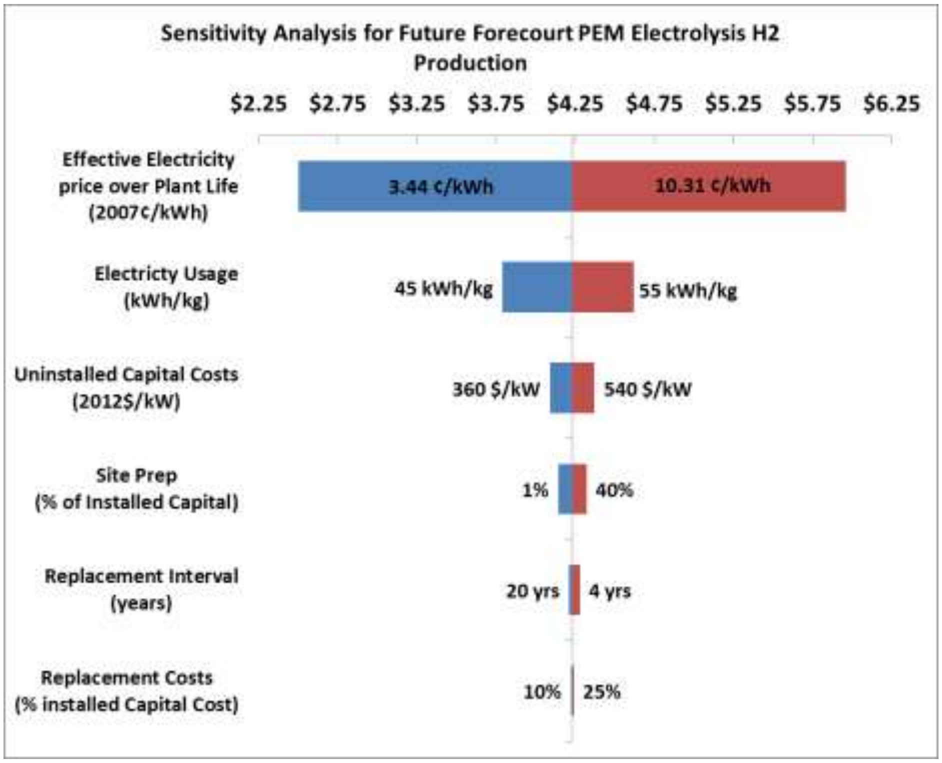

Figure 5 - Tornado chart showing parameter sensitivities for *Projected Future* Forecourt PEM Electrolysis case.

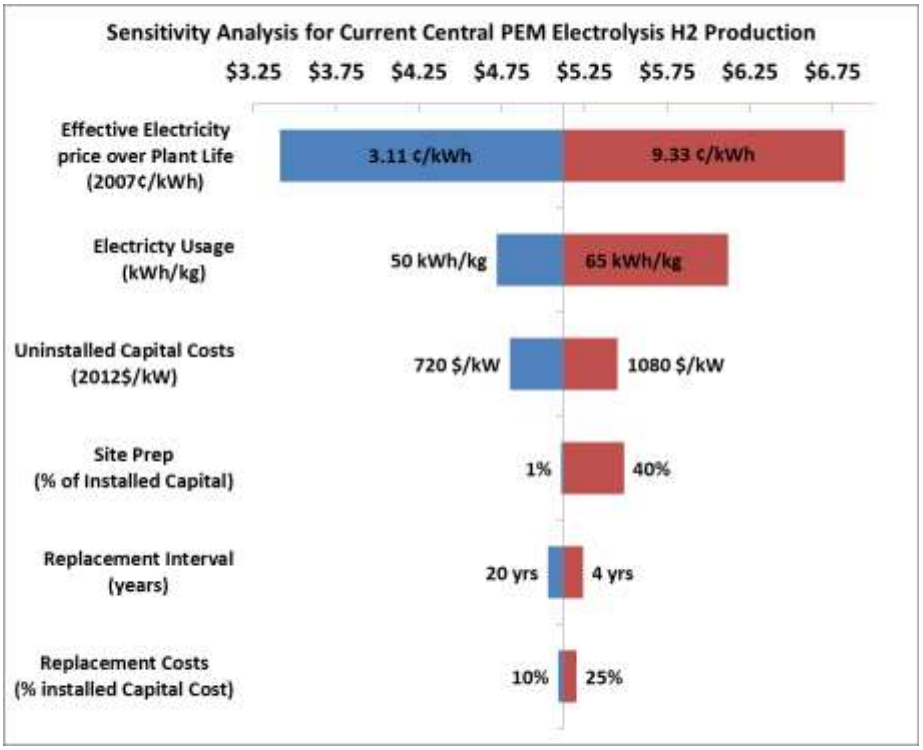

Figure 6 - Tornado chart showing parameter sensitivities for *Projected Current* Central PEM Electrolysis case.

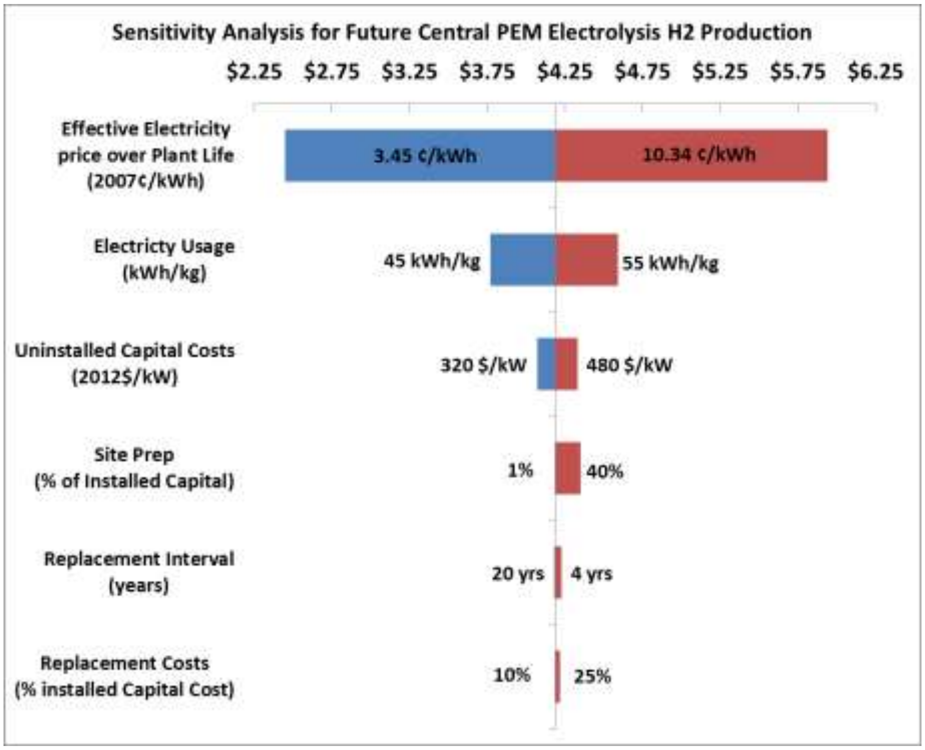

Figure 7 - Tornado chart showing parameter sensitivities for *Projected Future* Central PEM Electrolysis case.

3.1.4. Discussion Points With Regards to PEM

Electricity is the main parameter driving the cost of the PEM electrolysis process. In both Forecourt and Central cases, the *Projected Current* stack and BOP capital costs are also high contributors to the overall cost structure. Despite having a less significant effect on the overall production cost, electrical usage is a more interesting sensitivity parameter for further research.

4. High Temperature Solid Oxide Electrolysis

4.1. Case Study Summary

Two H2A v3.1 cases were developed: a *Projected Current* case based on 2014 lab scale technology; and a *Projected Future* case based on expected technology advancements by 2025. Given that there are no commercial SOEC stacks or systems available, and that only limited long-term durability data exists for cells/stacks at relevant operating conditions, the *Projected Current* case was extrapolated from technology demonstrated at the laboratory scale. Both cases were based on input from, and were subsequently reviewed by six solid oxide electrolysis cell research organizations (laboratories and companies, four in the United States and two international) to ensure the study parameters and results were relevant and accurate. Based on consultations with the study participants, a central production capacity of 50,000 kg H_2/day was modeled (i.e., forecourt/distributed production scales ≤1,500 kg/day were not analyzed).

The modeled costs to produce hydrogen (untaxed, excluding delivery and dispensing) are summarized in **Table 4** for the two cases studied. The baseline cost projections in the table were derived using inputs from study participants on the electrolyzer stack and balance of plant costs. The lower and upper bounds in Table 4 were calculated based on a Monte Carlo analysis in which multiple input parameters were simultaneously varied to estimate the lower and upper bounds on hydrogen cost.

Table 4 - H2 Production High-Volume Cost Projections for the SOEC Cases[5]

Central H_2 Production SOEC Case Study	Low Value ($/kg H_2)	Baseline ($/kg H_2)	High Value ($/kg H_2)
Projected Current Case[21]	$3.73	$4.95	$5.84
Projected Future Case[22]	$2.80	$3.83	$4.67

4.2. SOEC System Description

Generalized system designs were developed for both the *Projected Current* and *Projected Future* baseline cases using inputs and guidance from the study participants. Both cases envision the electrolysis cells operating very close to the thermo-neutral operating point.[23] The system flow schematic for the *Projected Current* baseline shown in **Figure 8** is based on a stack temperature of 800°C with an outlet gas pressure of 300 psi. Byproduct oxygen is not captured. Heat to warm the reactants to the stack inlet temperature is provided from a generic heat source, without judgment as to the heating source, (See **Table 5** for the heating cost[24] for each case). Steam is used as a sweep-gas on the oxygen-generating side of the cells (cathodes) to lower the oxygen partial pressure and thereby enhance performance and limit corrosion. (Alternatively, air may be used as a sweep gas and to reduce potential chromium migration.) The generalized system design developed for the *Projected Future* baseline case is shown in **Figure 9**. While similar to the *Projected Current* baseline case, it represents a more technologically-advanced version with the following differences:

- 700 psi product gas pressure
- Inclusion of an exhaust gas expander to generate electrical power (resulting in a nominally higher system electrical efficiency)
- Reduced thermal losses (due to tighter thermal integration)
- Improvement in electrical rectification efficiency (nominally 95% for the *Projected Current* case and 97% for the *Projected Future* case).

Cell current density (at the operating point) and area specific resistance (ASR) were not primary inputs into the performance or cost analysis. Rather these parameters were used by the participants to estimate stack capital cost, which was then used within the H2A model. While electrical efficiency does not change much between the *Projected Current* and *Projected Future* cases, a large increase in current density is expected (at the same operating voltage) which is expected to reduce the stack footprint, thereby decreasing the stack cost per kW.

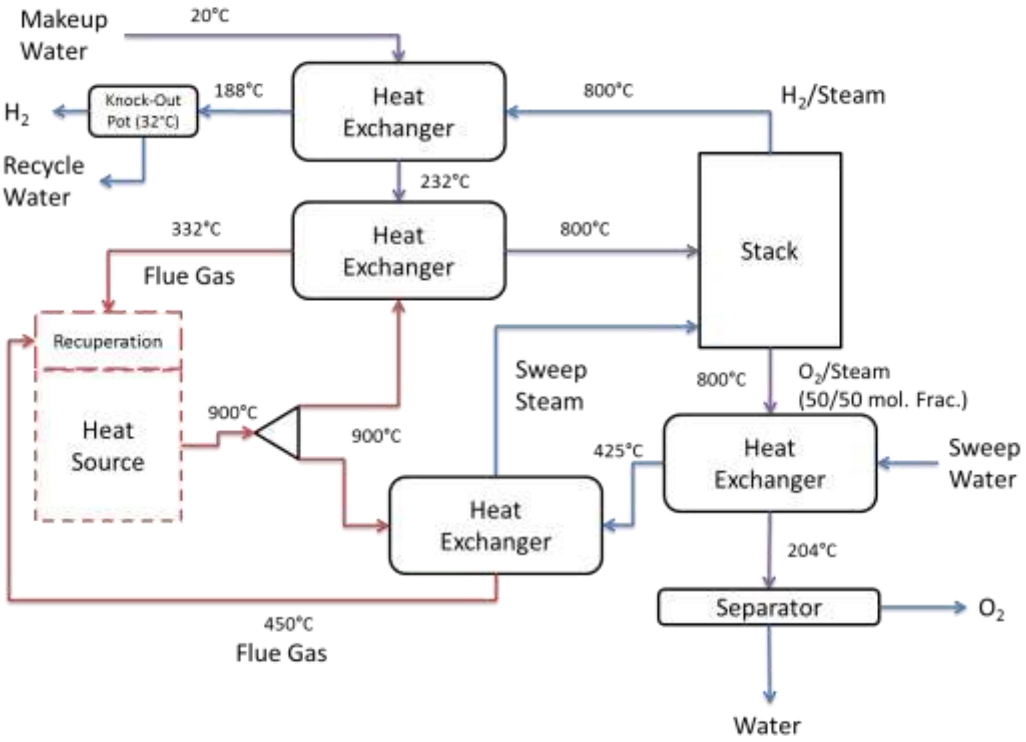

Figure 8 - *Projected Current* SOEC Baseline Design

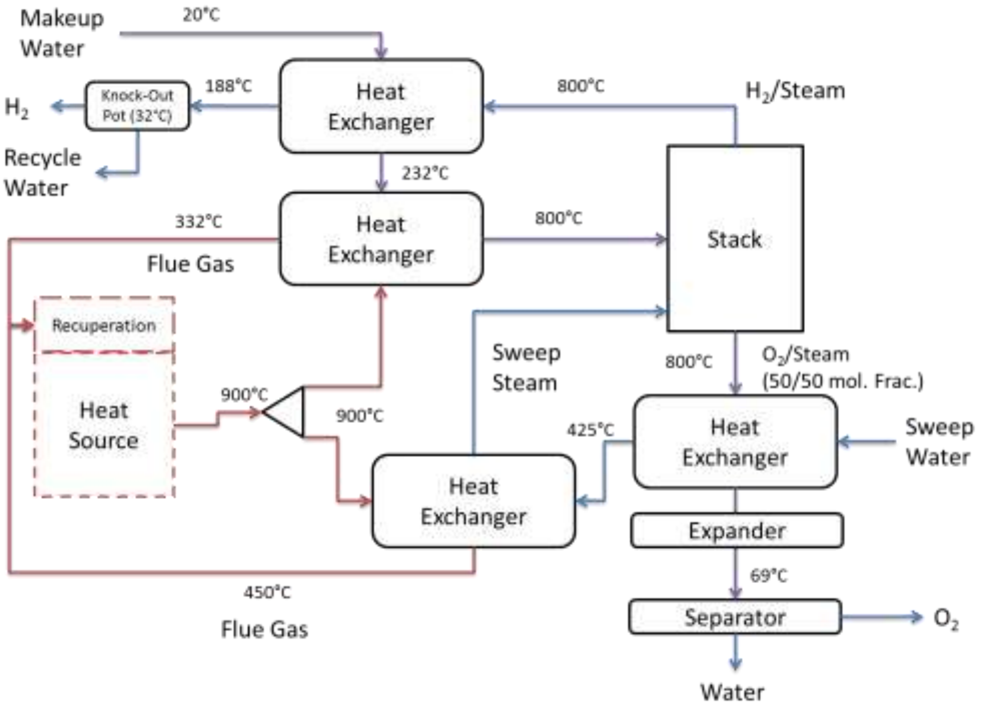

Figure 9 - *Projected Future* SOEC Baseline Design

The source of the high-grade heat required for SOEC operation is an important consideration for SOEC technology. (Some system configurations do not require high-grade waste heat.) For modeling and cost estimation purposes, a natural gas combustion system was selected as a representative generic heat source for SOEC systems as it provides a well-established and convenient analysis baseline. The capital and maintenance costs estimated for the burner system are dramatically less than the natural gas fuel costs by roughly three orders of magnitude. Consequently, changes in burner system capital cost are unlikely to have an appreciable impact on the effective overall cost of heat supplied by the system.

4.3. Baseline Input Parameters

The key parameters used to develop the two H2A v3.1 baseline case studies are shown in **Table 5**. Parameter values were drawn chiefly from responses to the questionnaire, but also were supported by engineering judgment/calculations and by utility pricing information from the Annual Energy Outlook (AEO).[17] Additional parameter values were drawn from standard H2A v3.1 default values[11] so as to create an overall assessment that is consistent with past H2A studies, but which is tailored to the unique attributes of the SOEC system.

To account for stack performance degradation, the analysis modeled stack operation at a constant voltage of 1.28V with the H_2 production rate (i.e., stack current density) decreasing over the course of a year. Based on study participant input, stack degradation rates of 0.9%/1,000h and 0.25%/1,000h and stack service lifetimes of 4 and 7 years were used for the *Projected Current* and *Projected Future* cases, respectively. A stack replacement schedule was developed where stacks reaching end of service life are removed and where stack capacity is added to bring the total H_2 production of the plant back to 100% at the beginning of each year. This process is repeated for the 40 year life span of the plant. The overall effective plant capacity (i.e. actual annual H_2 production divided by plant design annual H_2 production) represents the combined effect of diminishing H_2 production due to stack performance degradation during the year and the plant capacity factor due to planned/unplanned shutdowns.

Table 5 - Input Parameters for SOEC H2A Central Production Baseline Cases (costs in 2007$ and 2012$)

Parameter	Projected Current	Projected Future	Cost Basis
Plant Capacity (kg/day)	50,000	50,000	H2A
Total Uninstalled Capital (2012$/kW)	$820	$430	Ind. Questionnaire
Stack Capital Cost (2012$/kW)	$287	$99	Ind. Questionnaire
Balance of Plant (BOP) Capital Cost (2012$/kW)	$533	$331	Ind. Questionnaire
Total Energy Usage (kWh/kg)	50.9	46.6	Ind. Questionnaire
Net System Energy Efficiency[25]	66%	72%	Ind. Questionnaire
Stack Electrical Usage (kWh/kg) Stack Conversion Efficiency (% LHV H_2)	34.0 (98.0%)	34.0 (98.0%)	Ind. Questionnaire
System Electrical Usage (kWh/kg) System Conversion Efficiency (% LHV H_2)	36.8 (90.5%)	35.1 (94.9%)	Ind. Questionnaire
System Heat Usage (kWh/kg)	14.1	11.5	Ind. Questionnaire
Cell Voltage (V)	1.28	1.28	Ind. Questionnaire
Current Density[26] (mA/cm^2)	1,000	1,500	Ind. Questionnaire
Electrolyzer Power Consumption (MW)	76.6	73.1	Eng. Calculation
Effective Elec. Price over Life of Plant (2007¢/kWh)	6.24	6.89	AEO/Eng. Calc.
Electricity Price in Start-up Year[17] (2007¢/kWh)	5.74	6.59	AEO/Eng. Calc.
Thermal Energy Cost ($2007/GJ)[27] (2007¢/kWh)	10.1 (3.64)	11.5 (4.13)	AEO/Eng. Calc.
Hydrogen Outlet Pressure (MPa)	2.1 (300 psi)	4.8 (700 psi)	Ind. Questionnaire
Installation Cost (% of uninstalled capital cost)	12%	10%	H2A
Stack Service Life[28] (years)	4	7	Ind. Questionnaire
H2A Plant Capacity Factor	90%	90%	H2A
Percent Stack H_2 Production Rate due to degradation at end of first service year	83.2%	94.5%	H2A Calculation
Overall Effective Plant Capacity[29]	82.4%	87.5%	Eng. Calc
Effective Annual Stack Service Replacement Cost[30] (% of Stack Capital/year)	27.3%	12.8%	Eng. Calculation
Balance of Plant (BOP) Lifetime (years)	20	20	Ind. Questionnaire
BOP Replacement Cost (% of BOP initial investment)	100%	100%	AEO/Eng. Calc

4.4. Results and Sensitivity

The hydrogen production cost breakdown for the two H2A v3.1 SOEC baseline cases is shown in **Table 6**. These cases used inputs from the study participants to determine the most likely parametric values at a central scale for the two different technology years. The effects of deviations from these baseline inputs are considered separately in the sensitivity analysis section which shows that the primary cost driver for H_2 production via SOEC is the electricity required to run the electrolysis process. Unlike other cost categories, the price of electricity (as projected by AEO) is seen to increase between the *Projected Current* and *Projected Future* cases. This electricity price increase is partially offset by the higher electrical efficiency projected for the *Projected Future* case.

Table 6 - H_2 Production Cost Breakdowns in 2007$/kg H_2 for SOEC Baseline Cases

Component	***Projected Current*** **50,000 kg/day**	***Projected Future*** **50,000 kg/day**
Stack Capital Cost	$0.24	$0.09
BOP Capital Cost	$0.45	$0.29
Indirect Capital Cost and Replacement Cost	$1.00	$0.24
Decommissioning	$0.00	$0.00
Fixed Operations and Maintenance (O&M)	$0.38	$0.23
Thermal Energy Feedstock	$0.53	$0.49
Electricity Utility	$2.34	$2.49
Variable O&M	$0.01	$0.00
Total H_2 Production Cost (2007$/kg H_2)	$4.95	$3.83

Tornado charts were developed for the *Projected Current* and *Projected Future* cases for centralized SOEC hydrogen production to examine the impact of individual parameters on hydrogen cost in a single variable sensitivity analysis. These tornado charts, shown in **Figure 10** and **Figure 11**, plot the projected hydrogen cost variations on the x-axis against different single input parameters arranged along the y-axis. Specifically, the plots illustrate the H_2 production cost sensitivities to variations in (1) effective electricity price over life of plant; (2) uninstalled capital cost; (3) stack service lifetime; (4) average price of heat over life of plant; (5) thermal usage; (6) plant capacity due to operational downtime; and (7) electrical usage. Each tornado chart is organized from top to bottom to represent the most to least sensitive of the analyzed input parameters, respectively. The colored shading indicates either an increase (red) or a decrease (blue) in the baseline hydrogen cost from the change in input parameter. The data labels list the low and high values for the input parameters.

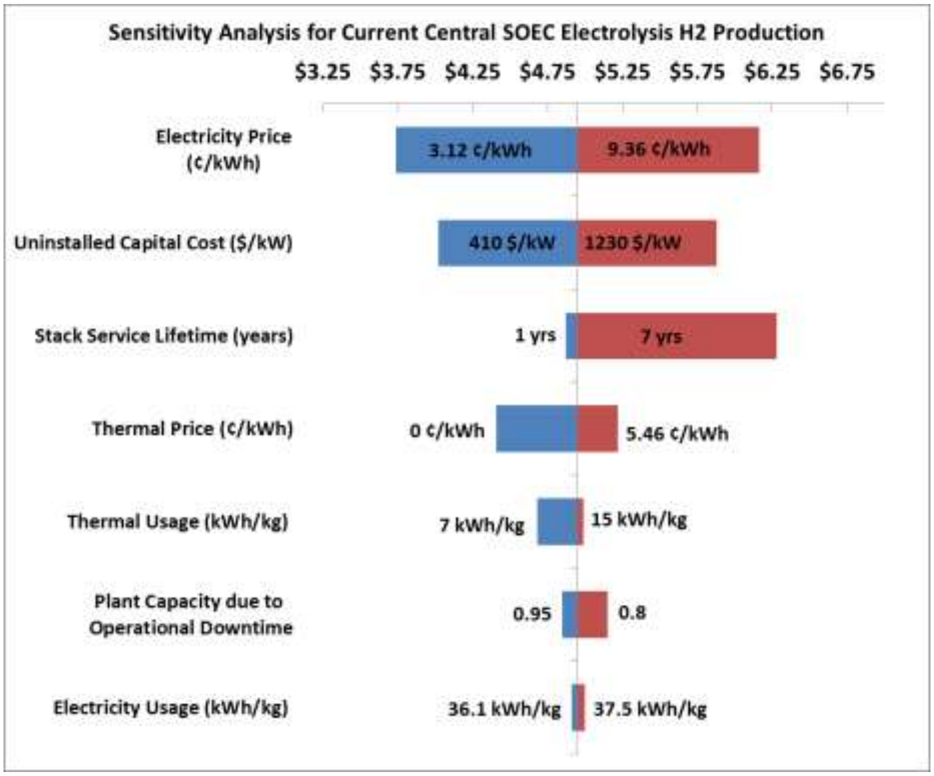

Figure 10 - Tornado chart showing parameter sensitivities for the *Projected Current* SOEC case

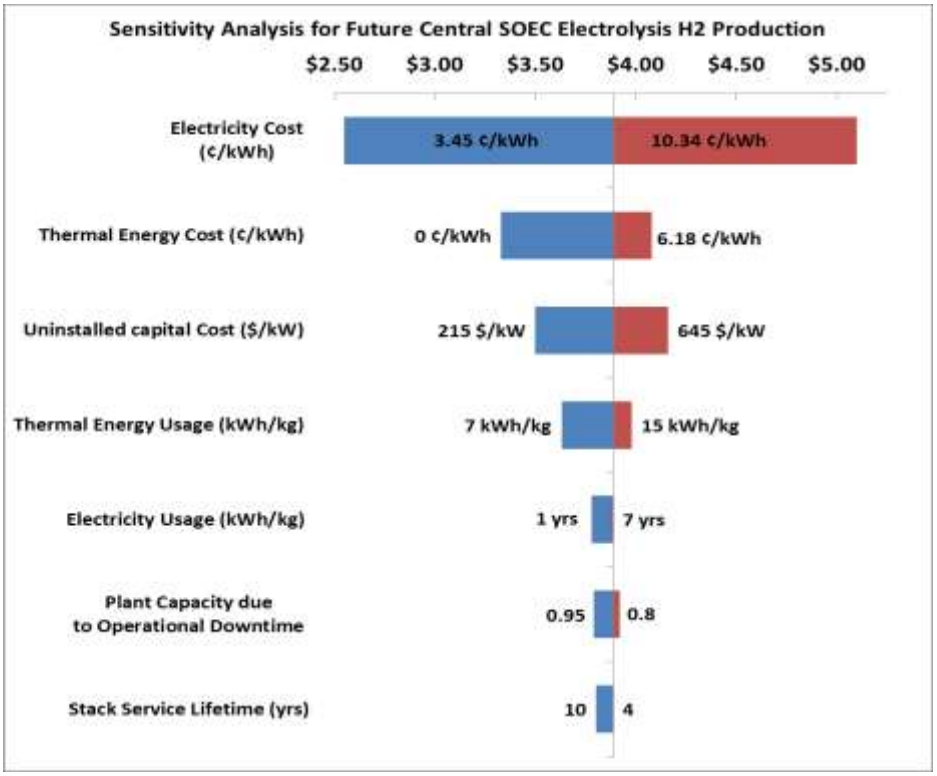

Figure 11 - Tornado chart showing parameter sensitivities for the *Projected Future* SOEC case

4.5. Discussion Points for SOEC

The tornado charts show that for the SOEC process investigated, for both *Projected Current* and *Projected Future* cases, hydrogen production cost is primarily dependent on and most sensitive to changes in the price of electricity. This result is consistent with both alkaline electrolysis and PEM electrolysis, where electricity price is also the main cost driver. Also, especially for the *Projected Current* case, the hydrogen production cost is sensitive to changes in the uninstalled capital cost and stack service lifetime. Note that electricity usage is comparatively far down the tornado graph as a direct result of the narrow range of usage values used within the sensitivity analysis. That narrow range is attributable to the narrow band of responses received for both the *Projected Current* and *Projected Future* cases from the industry respondents. Finally, the lower bound on heat price is set at zero to reflect the scenario where heat is available to the electrolyzer system at no cost.

5. H_2 Production via Fermentation of Biomass

5.1. Case Study Summary

The projected cost to produce hydrogen (H_2) from dark fermentation of biomass (corn stover) using techniques and strains currently in development at the laboratory scale is greater than \$50/kg[31] (untaxed, high system production rates). However, it is expected to drop dramatically in the future to \$5.65/kg by 2025, if assumed improvements in the technology and high volumes are realized. Two cases were considered, a *Projected Current* year case based on 2015 technology using performance and design parameters that have been simultaneously demonstrated in the lab at low reactor volumes, and a *projected Future* case based on projected technological advancements by 2025. The cost analysis was performed using the Hydrogen Analysis version 3.101 (H2A Production v3.101) model and its associated assumptions[32] for a centralized production facility with a production capacity of 50,000 kg H_2/day.[33] The analysis utilizes a system design based on lab-demonstrated hydrogen production procedures[34] and using capital costs derived from a 2013 NREL report[35] on the production of hydrocarbons from lignocellulosic compounds.

The modeled costs (untaxed, delivery and dispensing not included) to produce hydrogen are summarized in **Table 7** for the two cases studied. The baseline costs are the projected costs to produce hydrogen for the *Projected Current* and *Projected Future* cases. The low and high values are included to reflect a range of uncertainty (±25%) in installed capital costs (with all other techno-economic inputs the same as in the baseline cases).

Table 7 - High-volume cost projections for hydrogen production from a centralized dark fermentation facility with 50,000 kg H_2 /day production capacity using 2014 (*Projected Current*) and 2025 (*Projected Future*) technologies.

Case Study	Low Value ($/kg H_2)	Baseline ($/kg H_2)	High Value ($/kg H_2)
Projected Current Case (2014)	$40.88	$51.02	$75.67
Projected Future Case[36](2025)	$3.39	$5.65	$7.90

5.2. Dark Fermentation System Description

There are no commercial dark fermentation hydrogen production facilities on which to base the system designs. Consequently, relevant techno-economic analysis inputs were derived for a hypothesized system. The process design for this projected hydrogen fermentation plant draws from two main sources: a hydrogen production fermentation plant previously conceptualized in 2009,[37] and a design and cost report for the production of lignocellulosic ethanol.[35] A 2013 National Renewable Energy Laboratory (NREL) report was supported and supplemented by data from previous versions of the report.[38,39] Data from these reports were adjusted to reflect recent technological progress and thinking (See **Figure 12**) and to adapt for hydrogen production. The alterations primarily consist of elimination of the distillation columns and scaling of the waste water system. The distillation columns are not required in the hypothesized system design as the system is not producing and purifying ethanol. The waste water treatment was scaled according to the size of the system and the content of the organic components in the waste stream. These organic compounds were modeled as being converted to biogas in the waste water treatment center. Consistent with the 2009 analysis and the laboratory data, hydrolysis pretreatment of cellulose and hemicellulose have been combined into one reactor, with combined saccharification and fermentation assumed to occur in a subsequent single reactor. In accordance with the 2009 analysis, *Clostridium thermocellum* converts cellulose to hydrogen and other byproducts in the *Projected Current* case analysis. *Clostridium thermocellum* can also be combined with other microbes to create a microbial consortium that is capable of converting both cellulose and hemicellulose derivatives to hydrogen. This microbe consortium is modeled for the *projected Future* case as a technological improvement that will increase corn stover to hydrogen yield. Reaction parameters such as reaction rates, compound concentration, and product yields were provided by NREL.[40] Capital equipment design, cost, and performance data were gathered from the literature,[35] modified as appropriate to meet the hydrogen fermentation plant needs, and used to populate a set of baseline cases. Inputs to the H2A model fell into five primary categories:

1. Engineering system definition,
2. Capital costs,
3. Operating costs,
4. Variable and fixed expenses, and
5. Replacement costs

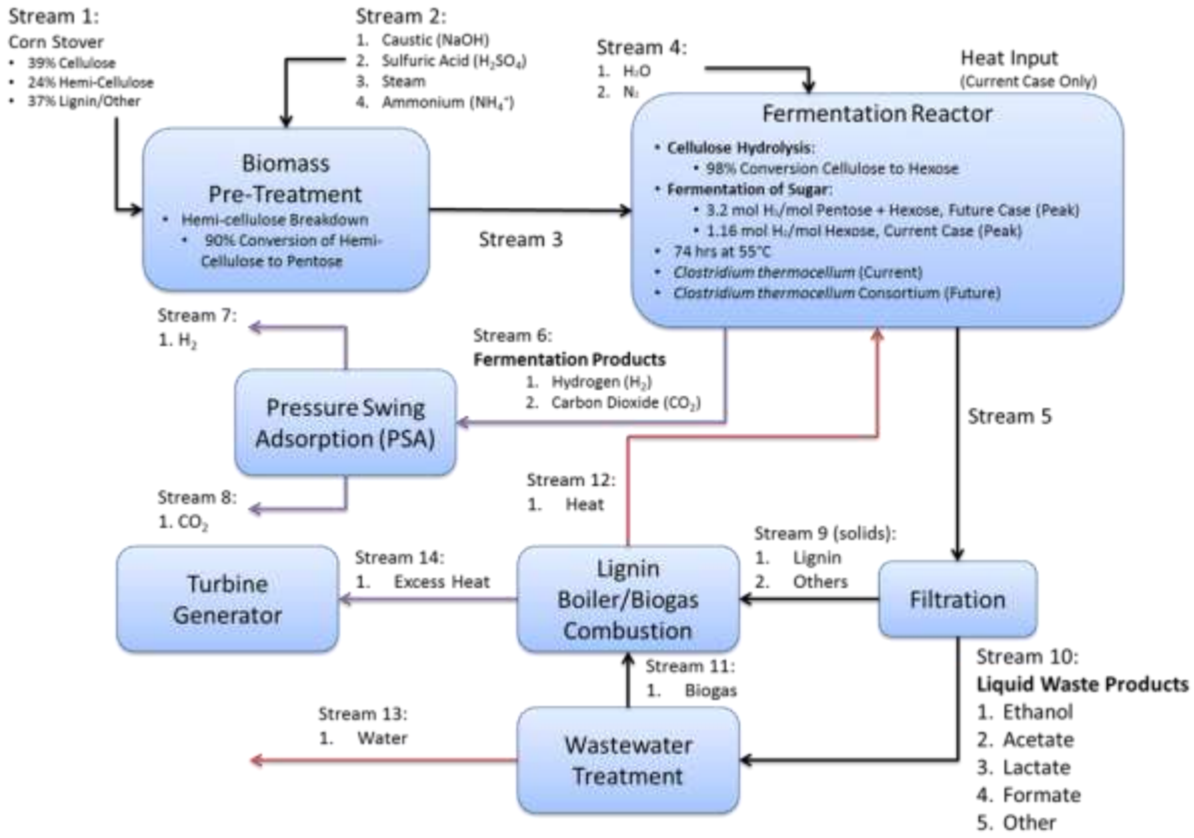

Figure 12 – Process flow diagram used as the model system to project the cost to produce hydrogen via fermentation at a central production facility with a 50,000 kg/day capacity.

For each technology year considered, a model was created to determine the capital cost based on production volume, molar conversion of sugars to hydrogen, heat and energy requirements, and energy byproducts. Both *Projected Current* and *Projected Future* cases envision a hydrogen fermentation plant (**Figure 12**) in which feed material (modeled as corn stover, with cost of preparation for processing included) is delivered to the plant. Feedstock is first broken down via the pretreatment process.[41] The partially converted feedstock is then sent to a fermentation reactor in which two main reactions occur: (1) cellulose is hydrolyzed to hexose sugars, and (2) sugars are fermented into hydrogen and other products. For the *Projected Current* case, *C. thermocellum* resides within the fermentation reactor and ferments only hexose sugars. For the *projected Future* case, a consortium of microbes, based on *C. thermocellum*, resides in the fermentation reactor and ferments both hexose and pentose sugars. The fermentation reactors are modeled as operating at 55°C for a given batch time. The fermentation batch time was determined through a cost optimization study, by plotting projected hydrogen cost as a function of fermentation time (see **Figure 13**). Maximum fermentation time was limited to 74 hours, as NREL data showed maximum conversion at that limit. The optimization curves, based on 2015 lab results, suggest the minimum cost of hydrogen corresponds to a fermentation time of 74 hours for both the *Projected Current* and *Projected Future* cases.[42]

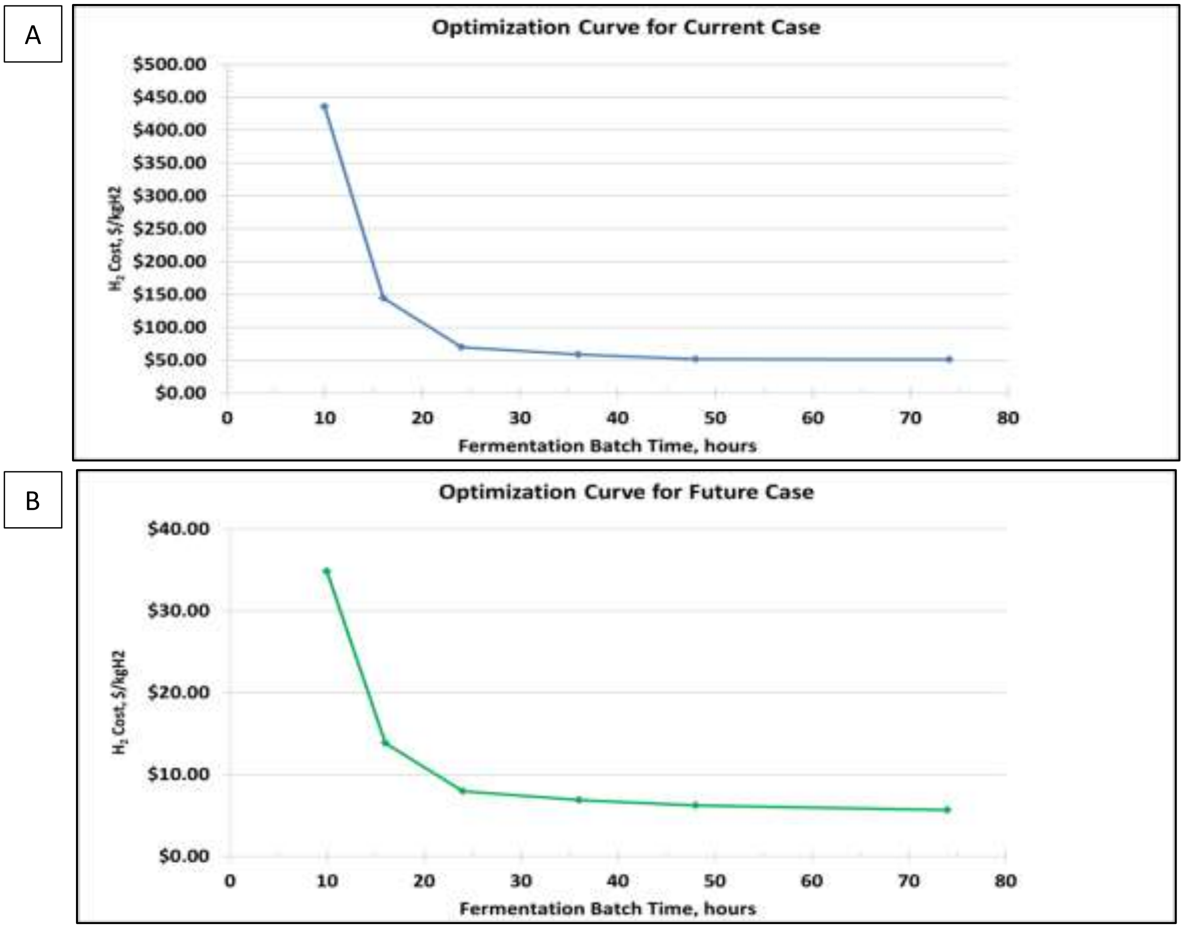

Figure 13 - Optimization Curves for Fermentation Cases A) *Projected Current* case and B) *Projected Future* case

The H_2 and CO_2 gaseous products are vented from the fermentation reactor and separated from one another via Pressure Swing Adsorption (PSA). After fermentation, the broth is filtered, with the solids fraction (mostly lignin) used for energy recovery, and the liquid fraction (a dilute mixture of ethanol, acetate, and other organic acids) sent to waste water treatment. The waste water treatment plant is based on anaerobic digestion to create a byproduct gas of mostly methane. The byproduct gases, as well as the lignin, are combusted to generate thermal energy to heat the system.[43] The excess thermal energy is converted to electricity in a gas turbine electrical generator and sold to the grid for byproduct credit equivalent to $11.93 and $8.19 per kg of hydrogen produced for the *Projected Current* and *projected Future* cases, respectively.[44] In the cases without byproduct credit, the hydrogen cost is higher since there is no revenue collected from electricity sales, but the cost increase is partially offset by a reduction in system capital cost as there is no need for the gas turbine. Fuel cell conversion of the byproduct gases into electricity was considered but ultimately rejected due to a desire to focus the cases on fermentation technology and to leave them unencumbered by uncertainty of fuel cell capital cost projections.

Key differences between the *Projected Current* and *Projected Future* case are:

1) A change in fermentation broth concentration[45] from 12.8 g/L (*Projected Current*) to 175 g/L (*Projected Future*).[46,47]
2) A change from *C. thermocellum* (capable of converting only hexose sugars) in the *Projected Current* case to a microbial consortium (capable of converting both hexose and pentose sugars) in the *projected Future* case.
3) An increase in peak molar conversion of sugars to H_2 from 1.16 mol H_2/mol sugar (*Projected Current*) to 3.2 mol H_2/mol sugar (*Projected Future*). [48]

The yields were determined based on experimental data from NREL. NREL ran fermentation studies with acid hydrolysis pre-treated corn stover (PCS) and Avicel®[49] feedstocks, creating fermentation broth with cellulose concentrations of 1 g, 2.5 g, and 5 g cellulose/L.[50] The peak molar yields ranged from approximately equivalent to 1.16 mol H_2/mol hexose to approximately 3.2 mol H_2/mol hexose, at 74 hours.[48] The selected operating points for lowest system cost are 1.16 mol H_2/mol hexose at 74 hours for a 12.8 gram corn stover/L[51] broth concentration for the *Projected Current* case, and, for the *Projected Future* Case, 3.2 mol H_2/mol sugar (pentose and hexose) at 74 hours for a 175 g corn stover/L broth concentration. The *Projected Future* case model used the highest yield demonstrated by NREL. Note, however, that this yield occurred at the lowest fermentation broth concentration tested (1 g cellulose/L) while the *Projected Future* case is based on achievement of this high molar yield at the *Projected Future* broth concentration of 175 g/L. This is a substantial projected performance improvement, but one deemed reasonable and appropriate for the 2025 timeframe of the *Projected Future* case.

Byproduct sales of the ethanol and acetate produced during fermentation were considered, but they were ultimately not included due to the unfavorable economics associated with concentrating and isolating the products to levels required for the marketplace.

5.3. Baseline Input Parameters

The parameters used in the two H2A v3.1 baseline case studies are summarized in **Table 8**. Parameter values are based on the hypothetical plant shown in **Figure 12** and conversations with industry researchers. They are supported by standard H2A v3.1 default values,[52] engineering judgment and calculations, and utility pricing information from the Annual Energy Outlook (AEO). [53]

While broth concentration is not an input parameter for H2A, it is included in **Table 8** because of the large impact it has on fermenter heating requirements and the overall system capital cost. The concentration of the broth directly adjusts the volume of the fermentation broth, which in turn adjusts the quantity of reactors required, total reactor capital cost, and the total heating requirements. In the *Projected Current* case, 2.7 billion liters of broth per total batch (requiring 728 individual reactors) are required to produce 50,000 kg H_2/day while only 43.5 million liters of broth (12 reactors) are needed per batch in the *Projected Future* case, due to the higher broth concentration and higher molar conversion. This directly affects the system capital cost and energy balance that are critical to hydrogen cost. Broth concentration contributes to an electrical energy surplus of 179kWh/kg H_2 and 116 kWh/kg H_2 in the *Projected Current* and *Projected Future* cases, respectively.

Table 8 - Input parameters and other key parameters for H2A Production cases for fermentation of corn stover (and in 2010$[54], as marked)

Parameter	*Projected Current* Central	*Projected Future* Central
Plant Capacity (kg/day)	50,000	50,000
Fermentation Broth Concentration[55] (g/L)	12.8	175
Broth volume per batch (L)	2.7 billion	43.5 million
Number of reactors required	728	12
Total Uninstalled Capital (2010$)	$1,773M	$386M
Total Feedstock Required (kg/kg H_2)	229.2	49.47
Hemi-Cellulose to Pentose Conversion (%) **(in pretreatment reactors)**	90	90
Pentose Conversion (%) **(in fermentation reactor)**	0	100
Cellulose to Hexose Conversion (%)[35,56] **(in fermentation reactor)**	98	98
Hexose Conversion (%) **(in fermentation)**	100	100
Molar Conversion (mol H_2/mol Sugar) **(in fermentation reactor)**	1.16 molH_2/mol Hexose 74h batch time	3.2 molH_2/mol (Pentose & Hexose) for 74h batch time
Energy Byproduct Recovery	Energy Excess	Energy Excess
Electrical Energy Purchased (kWh/kg H_2)	5.4	2.6
Electrical Energy Byproduct (kWh/kg H_2)[57,58]	179	116
Repair And Maintenance Costs **(% of capital cost/year)**	0.5	0.5

5.4. Results and Sensitivity

The hydrogen production cost breakdown for the two H2A v3.1 fermentation cases is shown in **Table 9**. Large differences in capital cost are observed between the *Projected Current* and *Projected Future* cases and result primarily from the low concentration of the fermentation broth in the *Projected Current* case, which leads to a large number of high volume reactors and supplemental equipment. Electrical energy costs required to run the fermentation plant appear on the “Variable O&M” line of the cost breakdown.

The excess electrical energy generated in both cases can be sold back to the grid. The revenue generated by this byproduct appears on the “Byproduct Credits” line of the cost breakdown; the effective electricity byproduct selling price is 6.40¢/kWh and 6.60¢/kWh (levelized over the 40 year analysis period) for the *Projected Current* and *Projected Future* cases, respectively.

Table 9 - H_2 Production cost breakdowns in 2007$/kg H_2 for Fermentation baseline cases

Component	**Projected Current Central 50,000 kg/day**	**Projected Future Central 50,000 kg/day**
Installed Capital Cost	$36.07	$7.86
Decommissioning	$0.05	$0.01
Fixed operations and maintenance (O&M)	$5.67	$1.49
Feedstock Costs[5]	$18.01	$3.82
Byproduct Credits	-$11.93	-$8.19
Variable O&M (incl. electrical utilities)	$3.15	$0.65
Total H_2 Production Cost (2007$/kg H_2)	$51.02	$5.65[59]

Results from a sensitivity study are plotted in a tornado chart, displayed in **Figure 14**, and show projected hydrogen cost variations on the x-axis against single input parameters along the y-axis. Specifically, the plots illustrate the H_2 production cost sensitivities to variations in: (1) feedstock cost; (2) total installed capital cost; (3) broth concentration; (4) electrical turbine generator efficiency; and (5) reduced fermentation time due to an increased reaction rate. The tornado chart is organized from top to bottom to represent the most to least sensitive of the analyzed input parameters. The colored shading indicates either an increase (red) or a decrease (blue) from the baseline hydrogen cost due to the change in input parameter value. The y-axis labels list the low, baseline, and high values for the input parameter.

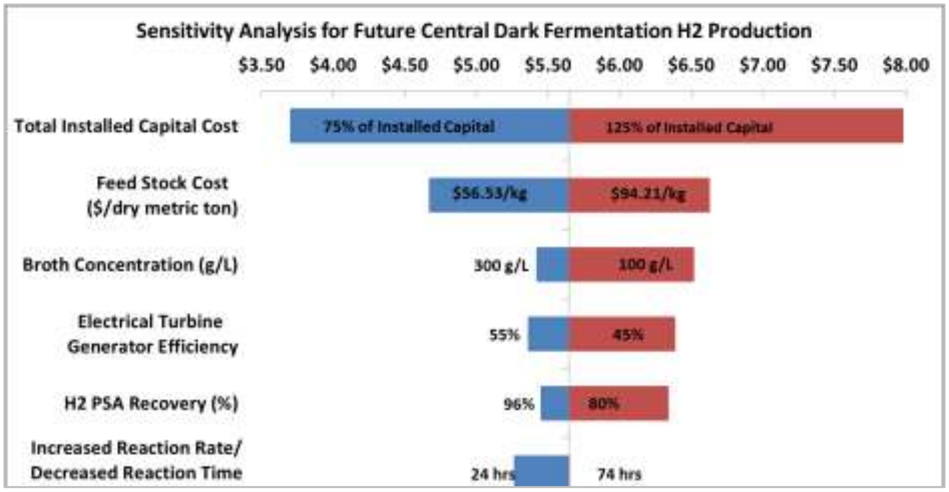

Figure 14 - Tornado chart showing parameter sensitivities for the *Projected Future* central fermentation case.

The tornado chart shows that the fermentation process investigated is highly dependent on the system capital cost which was varied +/-25% based on the perceived accuracy of the capital cost estimates. Hydrogen cost is also sensitive to changes in the price of the feedstock. Broth concentration appears to be only a modestly sensitive parameter but this result is only valid because the broth concentration parameter is limited to the lower bound of 100g/L. Were broth concentration evaluated at the *Projected Current* case value of 12.8g/L, it would be the dominant parameter in a sensitivity analysis due to its impact on the energy balance and capital equipment costs (see above for complete description.

5.5. Discussion Points for Dark Fermentation

Hydrogen produced via fermentation will require advances in technology to become a feasible production technology. The molar yield of the conversion of biomass to hydrogen will need to be improved and the operating fermentation broth concentration will need to be substantially higher than currently demonstrated in the lab. Even with these advances, the cost of hydrogen is projected to be approximately $13.19/kg prior to consideration of any byproduct credits. Byproduct credits offer a significant opportunity to reduce the cost of hydrogen. However, the production of the byproducts (electrical or chemical) must not reduce hydrogen production. Further cost reduction to approximately $3/kg H_2 may be achievable by lowering equipment capital cost, further increasing the molar yield, increasing electrical byproduct generation, increasing PSA recovery, and raising the fermentation broth concentration above 175 g/L. Further research is needed to explore the potential, feasibility, and extent of these improvements.

6. H_2 Production from Steam Reforming of Pyrolysis in Monolithic Piston-Type Reactors

6.1. Case Study Summary

Two H2A v3.101 cases were developed: a *Projected Current* and a *Projected Future* case based on expected technology advancements by 2025.[60] Fundamentally, the (Monolithic Piston-type Reactor) MPR reforming process analyzed is similar to Steam-Methane Reforming (SMR) except for adjustments required due to the fuel in question being pyrolysis oil instead of methane. At the beginning of the MPR process, oil is pumped to pressure, heated, mixed with steam, and fed to the MPR. The vaporized oil and steam are passed across a custom catalyst-coated TiO_2 monolith at 600-650°C and an operating pressure of 24 bar. The catalyst coated monolith promotes oil:steam ratio reforming. Every other channel of the monolith is packed with a powder composite sorbent (primarily comprised of dolomite) (See Figure 15). CO_2 adsorbs to the composite sorbent, allowing H_2 to exit the MPR at a relatively high purity when compared to traditional steam reforming processes (both oil:steam reforming and SMR). Other gas components are formed during reforming, primarily CO and CH_4, and do not adsorb to the composite sorbent.[61] The product H_2, along with CO, CH_4, and residual CO_2, exit in the reactor product stream at approximately 500°C. The other major compound formed during the reforming reaction is coke, which will deposit on the monolith and the sorbent.[62] After reforming is complete, the reactor switches to regeneration mode. The fuel feed to the reactor is stopped, the reactor pressure reduced, and air is blown through the reactor. The air flow ignites the coke, heating the monolith to approximately 700°C, and releasing CO_2 from the composite sorbent.

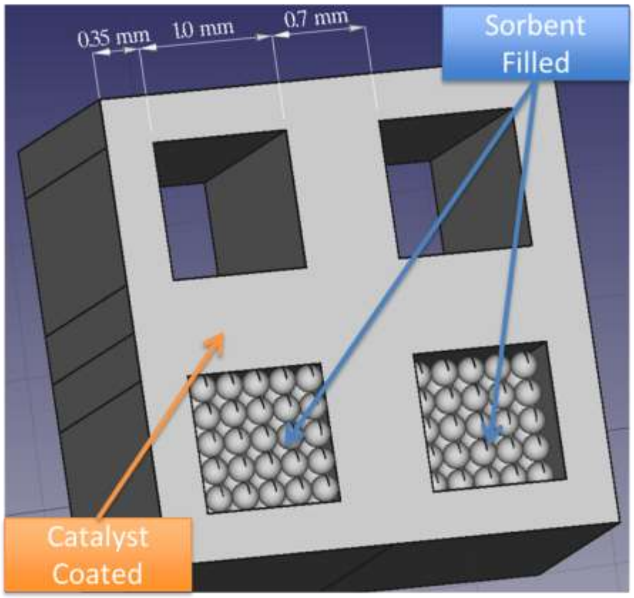

Figure 15 - Sample drawing of monolith with catalyst coating and sorbent filled channels

After review, it was determined that insufficient experimental data existed to complete H2A cases with the required degree of accuracy. As such, the cases were placed on hold until such time as sufficient data is available to complete H2A cases with the highest standard of quality.

7. H_2 Production via a Reformer-Electrolyzer-Purifier

7.1. Case Study Summary

H2A case studies were developed for Fuel Cell Energy’s Reformer-Electrolyzer-Purifier (REP) technology. The REP is a derivative of FCE’s Molten Carbonate Fuel Cell technology and is extremely similar in design and materials to FCE’s line of Direct Fuel Cell (DFC®) products. The core of the REP system is essentially a smaller DFC® stack operated in reverse (i.e. in electrolysis mode to produce H_2 rather fuel cell mode to produce electricity). Because the electrolyte passes carbon ions, the REP also acts to purify the fuel/H_2 stream and can achieve >95% H_2 purity on the exit stream.

FCE has identified two system designs for study. The first is a “Standalone” system that solely produces H_2 from natural gas, water, and electricity inputs. The second system design is an “Integrated” system that combines an REP (electrolysis) stack with an existing 5-7MW DFC® power plant. A small slip stream of partially reformed gas is removed from the MCFC and fed to the REP, allowing the integrated unit to produce H_2 and power from inputs of natural gas and water. Efficiency of the integrated unit is enhanced by “sharing” the reformer within the fuel cell stack and by making use of system waste heat.

Design parameters, capital costs, fuel and energy usages, and system operating conditions were provided by FCE. SA confirmed these values through Hysys modeling and used the information, combined with other analyses, to develop complete H2A studies for both the Standalone and Integrated cases. The *baseline* cost projections are presented in **Table 10** below. The baseline case was predicted on capital costs for a system production rate of 500 per year. The *low* and *high* values in the table were calculated based on a Monte Carlo analysis in which multiple input parameters were simultaneously varied to estimate the lower and upper bounds on hydrogen cost with 90% confidence (i.e. there is a 90% probability that H_2 cost will be between the low and high values).

Table 10 - H_2 Production Low-Volume Cost Projections for the REP Cases[63]

Projected Current H_2 Production REP Case Study	Low Value ($/kg H_2)	Baseline ($/kg H_2)	High Value ($/kg H_2)
Integrated	$2.30	$2.58	$3.06
Standalone	$2.75	$3.10	$3.59

7.2. Analytical Basis

Analyses to project the high volume cost of producing hydrogen at a forecourt facility[64] by use of the REP with a plant capacity of 1,500 kg/day were performed by Strategic Analysis, Inc. (SA) in conjunction with the National Renewable Energy Laboratory (NREL) using the H2A v3.101 model. Case studies were developed for two different systems. Both cases are *Projected Current*[65] (2015) cases. The cases are described as Standalone and Integrated. While both cases are based on REP technology and have target H_2 production rates of 1,500 kg/yr, the systems are extremely different. The Standalone system conducts steam methane reforming followed by electrolysis and purification in a REP unit (See **Figure 16**). The Integrated system combines Direct Fuel Cell (DFC®) technology for power production with a REP unit. The DFC® in the Integrated system is a 5-7MW power generator (a current commercial product developed by FCE with operating units in use) which pulls a small slip-stream of reformed gas from the DFC® and feeds it into a REP for further H_2 production and H_2 purification (See **Figure 17**). Central site production (eg. ~50,000 kgH_2/day at a city edge or rural location) was not analyzed for this technology.

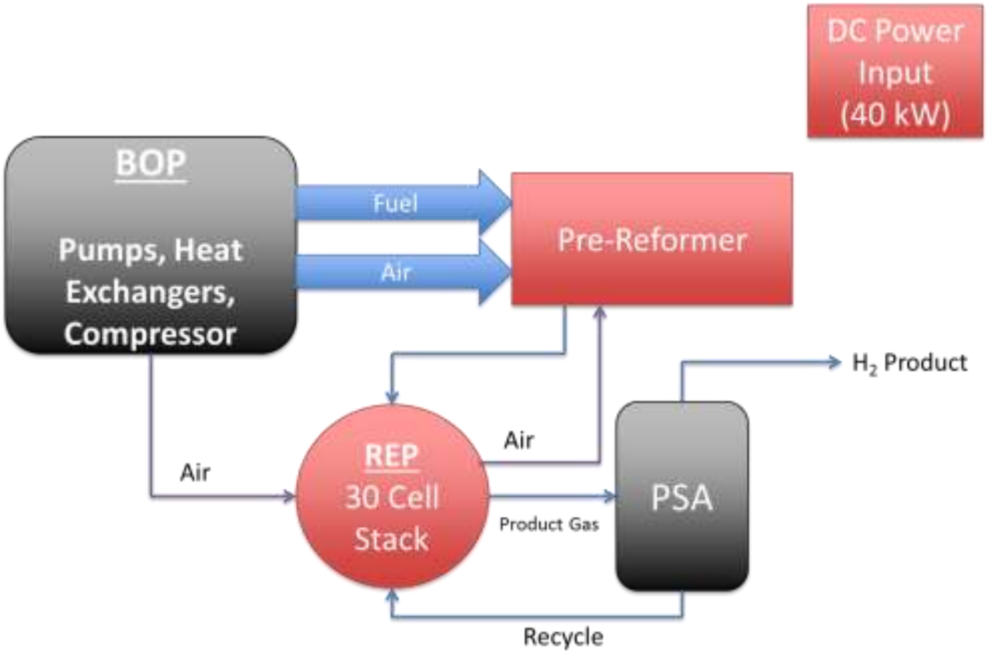

Figure 16 - Simplified Block Flow Diagram for FCE's *Projected Current* Standalone system with PSA purification to achieve "five nine's" purity

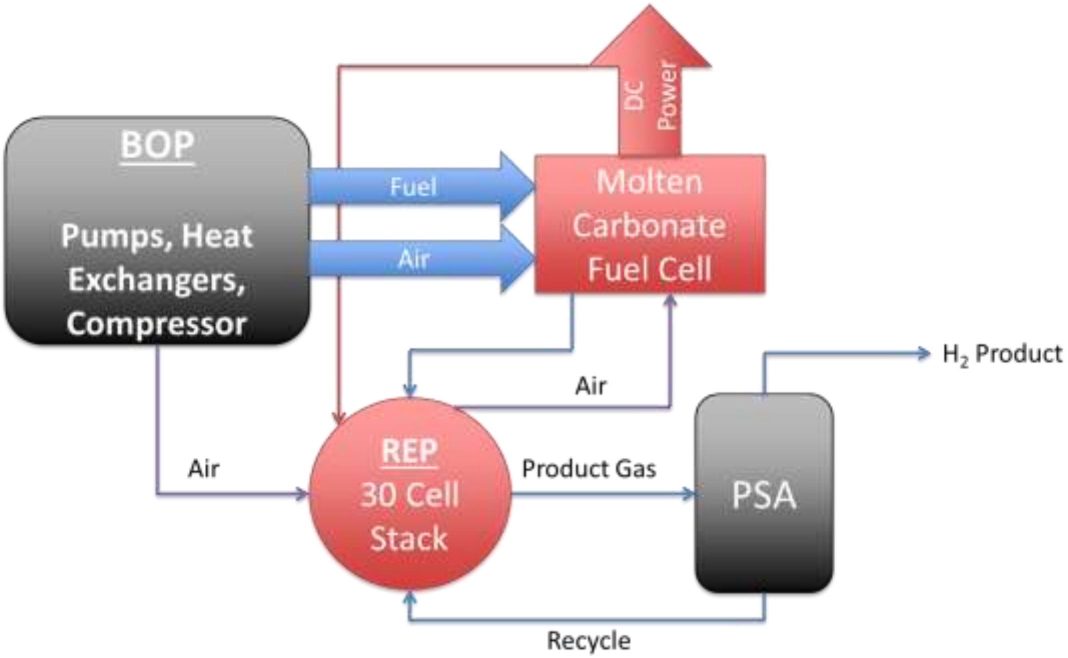

Figure 17 - Simplified Block Flow Diagram for FCE's *Projected Current* Integrated system with PSA purification to achieve "five nine's" purity

Relevant techno-economic data for the two cases was provided by researchers at FCE. The requested data included system operating parameters, reaction data, and current laboratory status and results. FCE provided SA with H2A cases and stream tables for both the integrated and standalone cases. With established system parameters, Aspen HYSYS® models were created to model the complete systems. Feed and operating parameters were matched to FCE data and the target production was set to 1,500 kg/day H_2. Variations in any parameters between the FCE data and the model were recorded.

The Reformer-Electrolyzer-Purifier technology is based on FCE's molten carbonate DFC® but is run in electrolyzer mode. DFC® is a commercial technology sold exclusively by FCE. The REP process begins with Steam Methane Reforming (SMR), which is a well-known process (Eq. 1), converting natural gas into H_2 and CO. CO can be further reacted with H_2O to produce H_2 and CO_2 in a Water Gas Shift (WGS) reaction (Eq. 2). In a traditional SMR plant, the WGS reaction is carried out in a separate catalyzed reactor at a reduced temperature to produce a syngas composition of approximately 72 mol% H_2.

SMR Reaction: $$CH_4 + H_2O \rightarrow H_2 + CO \qquad (1)$$

WGS Reaction: $$H_2O + CO \rightarrow H_2 + CO_2 \qquad (2)$$

However, in the REP system design, no separate WGS reactor is needed. After SMR, the product gas passes through the REP (essentially a Molten Carbonate Electrolyzer (MCE)). Not only does the REP electrolyze and split water to produce more H_2, but the REP transports CO_2 across the molten carbonate electrolyte as CO_3^{2-}. Once across the membrane, the CO_3^{2-} decomposes to form CO_2 and O_2. The transport of CO_3^{2-} across the membrane allows for a high purity H_2 ($\geq$ 95%) stream to leave the REP (the remaining ~5% gas is CO_2, CO, and CH_4). However, the REP exit gas next flows to a Pressure Swing Adsorption (PSA) unit for further purification to >99.999% ("five nines H_2 purity") suitable for fuel cell use. The off-gas from the PSA can be recycled back to the inlet of the REP for additional passes at reformation/electrolysis. In this manner, the system is able to achieve near 100% conversion of fuel into H_2. (An occasional purge of the system would be required to exhaust inert gases (N_2, Ar) from the system.)

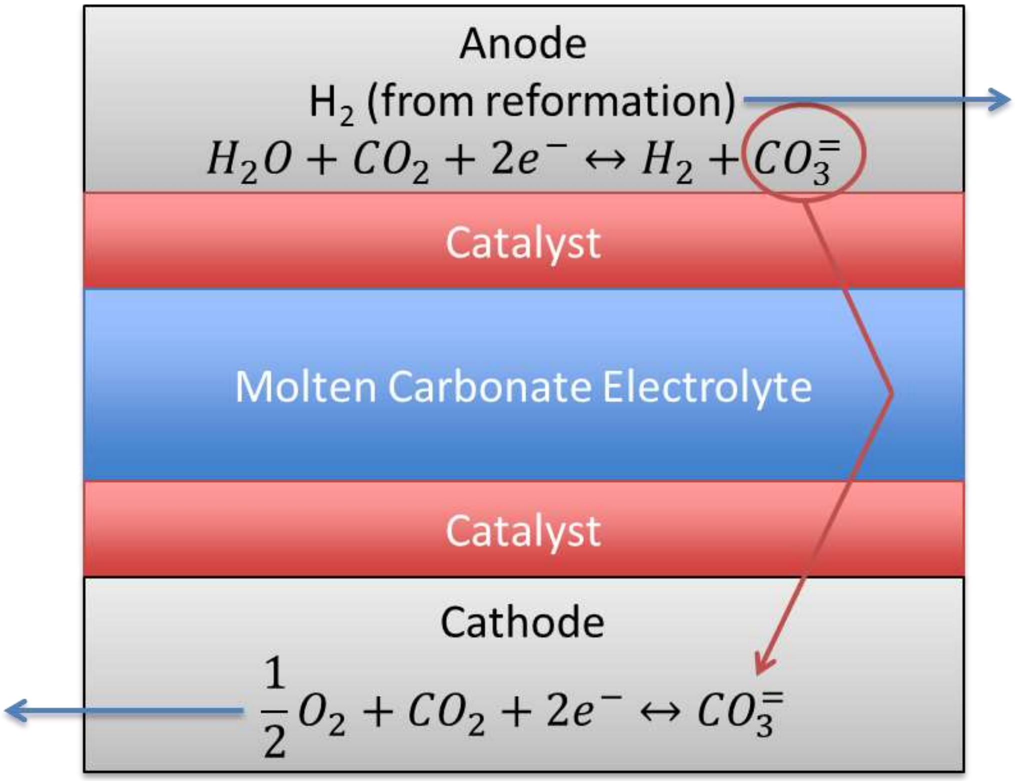

Figure 18 - Model of anode and cathode reactions with CO3$^=$ transport in REP.

While Pressure Swing Adsorption (PSA) was selected for purification in the modeled system, other purification systems, such as electrochemical separation, are also feasible. An H_2 compressor was added to raise the REP exit gas to a pressure suitable for PSA separation: 300psi was selected to match the system outlet pressure specified by DOE for H2A analysis. PSA gas recovery and sizing calculations were performed by Argonne National Laboratory (ANL). PSA cost models developed between SA and ANL were used to provide a cost for the PSA. For applications not needing fuel cell grade H_2 purity, the PSA and compressor can be eliminated and the >95% H_2 purity REP exhaust gas can exit the system. Such delivered H_2 would be less expensive due to the reduction in system capital cost.

Parameter values were drawn chiefly from FCE information and system modeling analysis, but also were supported by engineering judgment regarding appropriate process values for a functioning system. All parameter values were supported with internal calculations. Additional parameter values were drawn from standard H2A v3.101 default values[66] so as to create an overall assessment that is consistent with past H2A studies.

Table 11 - Key system parameters for REP case studies

Parameter	Units	Integrated Case	Standalone Case
Natural Gas Feed to REP	mmBTU/hr (kmol/hr)	0.107 (7.95)	0.143 (10.57)
Water:NG Feed	kmol H_2O/kmol NG	1.8	9.93
Electricity Utility	kWh/kg H_2	10.34	9.449
Capital Cost	2010$	1.58M	2.37M
Stack Replacement Interval	Years	5	5
Yearly maintenance cost	% of direct capital	5.81%	4.51%
REP Operating Temperature	°C	600	600
REP Operating Pressure	Bar	1.13	1.13

Along with the other major inputs to H2A, maintenance and replacement costs are of particular importance. Data for annual maintenance costs was provided by FCE. Replacement was modeled as REP stack replacement once every five years. In past H2A analyses of similar fuel cell systems, certain high-temperature BOP components were replaced every 10 years.[67] However, FCE currently operates its DFC® systems with a 20 year lifetime for BOP components and there is no evidence to suggest that the BOP components in either REP system would require different replacement parameters. Consequently, the FCE replacement schedule is used within the baseline H2A model but should the more conservative replacement schedule (shorter lifetime of select high-temperature components) be adopted, the system cost would only increase by approximately \$0.05/kg H_2 and \$0.07/kg H_2 for the integrated and standalone cases, respectively.

7.3. Baseline Cost Projection Results

The hydrogen production cost breakdowns for the two H2A v3.101 cases are shown in **Table 12**. The baseline case parameters, defined in **Table 11** above, were used to generate the costs within the context of the H2A model. Note that the capital cost of the integrated system only covers the equipment added to the fuel cell system: it does not include the cost of the fuel cell itself.

The cost differences between the two cases are significant. The integrated system has a lower capital cost due to the “free” reforming functionality provided by the integrated fuel cell. Additionally, the integrated system has lower effective natural gas usage since the fuel cell provides heat (for reformation) that otherwise would be supplied by burning natural gas. These two differences are major advantages for the integrated system and result in a \$0.63/kg H_2 cost differential.

Table 12 - Cost results for both Projected Current and Projected Future cases

	Hydrogen Production Cost Contribution	
Cost Component	Integrated REP System ($/kg)	Standalone REP System ($/kg)
Capital Costs	$0.99 (38.37%)	$1.34 (43.23%)
Decommissioning Costs	$0.01 (0.3876%)	$0.01 (0.3226%)
Fixed O&M	$0.34 (13.18%)	$0.41 (13.26%)
Feedstock Costs	$0.57 (22.09%)	$0.76 (24.52%)
Other Raw Material Costs	$0.00 (0.00%)	$0.00 (0.00%)
Byproduct Credits	$0.00 (0.00%)	$0.00 (0.00%)
Other Variable Costs (including electric utilities)	$0.67 (25.97%)	$0.58 (18.71%)
Total	**$2.58**	**$3.10**

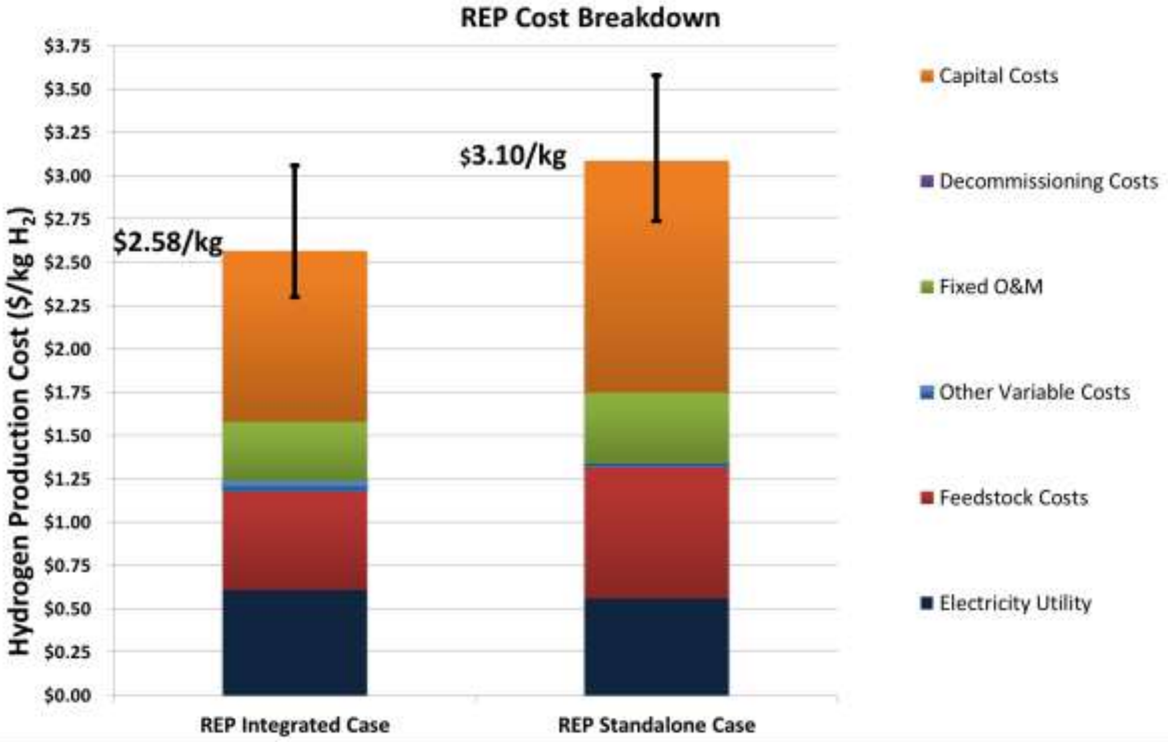

Figure 19 - Cost breakdown for both REP cases. Error bars are the result of a Monte Carlo analysis and represent a 95% confidence interval for the most likely value of H_2 production costs via REP

7.4. Results Summary: Cost Breakdown Plots and Tornado Sensitivity Charts

Figure 19 plots the H_2 production cost breakdown results for the two baseline cases shown in **Table 11**. Uncertainty analysis was performed to determine the most likely range of hydrogen costs using the Monte Carlo method. The H2A cost computation was repeated 75,000 times with each parameter independently and simultaneously varied between a low and high value. Parameters varied in the Monte Carlo and sensitivity analysis are shown in **Table 13** and **Table 14**. The resulting set of hydrogen cost projections allows assessment of the hydrogen cost range and the probability of occurrence. The ranges of potential hydrogen costs corresponding to the middle 90% of projections are displayed as "error bars" in **Figure 19**.[68]

Table 13 - Parameters varied in sensitivity studies and Monte Carlo analysis for Standalone case study

Parameter	Lower Bound	Projected Cost (2007$/kg H_2)	Baseline Value	Projected Cost (2007$/kg H_2)	Upper Bound	Projected Cost (2007$/kg H_2)
Natural Gas Cost	$4.01/ mmBTU	2.90	$5.34/ mmBTU	$3.10	$6.68/ mmBTU	$3.29
Electricity Cost	4.88¢/kWh	$2.98	6.10¢/kWh	$3.10	7.32¢/kWh	$3.21
Energy Usage	9.44 kWh/ kg H_2	$2.99	9.45 kWh/ kg H_2	$3.10	9.78 kWh/ kg H_2	$3.20
Capital Cost	$1,739,000	$2.74	$2,508,000	$3.10	$2,901,500	$3.46
Stack Replacement Interval	10 yrs	$3.00	5 yrs	$3.10	2 yrs	$3.43

Table 14 - Parameters varied in sensitivity studies and Monte Carlo analysis for Integrated case study

Parameter	Lower Bound	Projected Cost (2007$/kg H_2)	Baseline Value	Projected Cost (2007$/kg H_2)	Upper Bound	Projected Cost (2007$/kg H_2)
Natural Gas Cost	$4.01/ mmBTU	$2.43	$5.34/ mmBTU	$2.58	$6.68/ mmBTU	$2.73
Electricity Cost	4.88¢/kWh	$2.45	6.10¢/kWh	$2.58	7.32¢/kWh	$2.71
Energy Usage	9.44 kWh/ kg H_2	$2.52	10.34 kWh/ kg H_2	$2.58	10.52 kWh/ kg H_2	$2.59
Capital Cost	$1,268,200	$2.33	$1,713,000	$2.58	$1,901,900	$2.83
Stack Replacement Interval	10 yrs	$2.46	5 yrs	$2.58	2 yrs	$2.95

Tornado charts based on the parameter spreads summarized in **Table 13** and **Table 14** were developed for the integrated and standalone cases for forecourt hydrogen production to examine the impact of individual parameters on hydrogen cost in a single variable sensitivity analysis. These tornado charts, shown in **Figure 20** and **Figure 21**, plot the projected hydrogen cost variations on the x-axis against different single input parameters arranged along the y-axis. Specifically, the plots illustrate the H_2 production cost sensitivities to variations in (1) Natural Gas cost; (2) Electricity cost; (3) Energy Usage; (4) Stack Cost; (5) BoP Cost; and (6) Stack Replacement Interval. Each tornado chart is organized from top to bottom to represent the most to least sensitive of the analyzed input parameters, respectively. The colored shading indicates either an increase (red) or a decrease (blue) in the baseline hydrogen cost from the change in input parameter. The data labels list the low and high values for the input parameters.

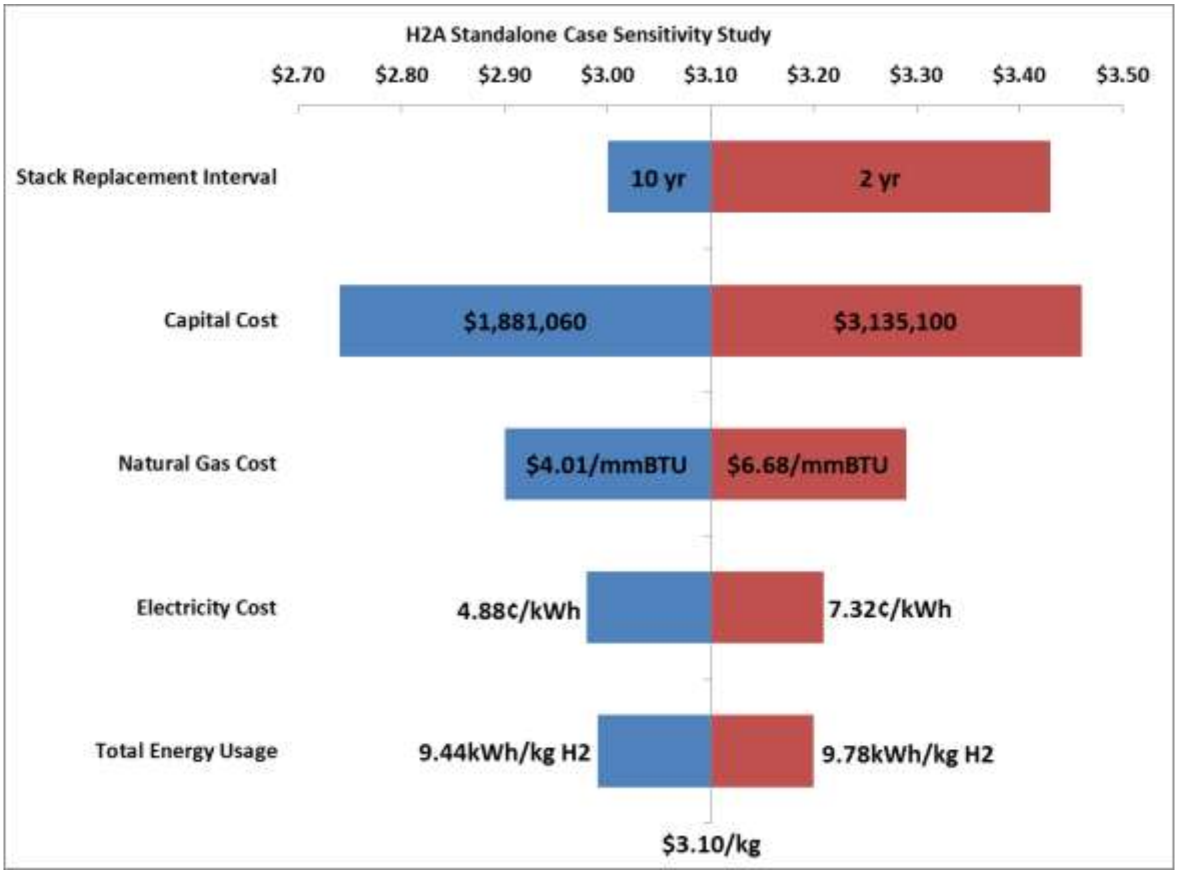

Figure 20 - Tornado chart for single parameter sensitivity study of REP Standalone case study

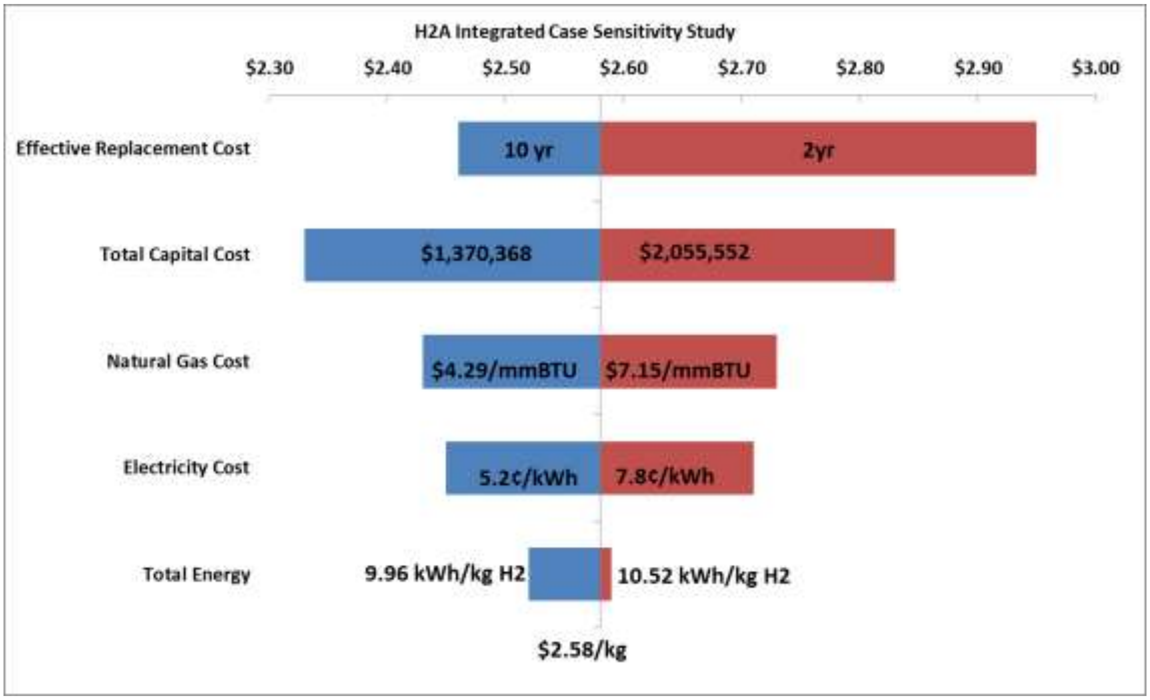

Figure 21 - Tornado chart for single parameter sensitivity study of REP Integrated case study

The tornado charts for both the Integrated and Standalone cases show similar results. Both show short (two year) stack life leading to significantly increased cost but limited H_2 cost reductions resulting from increases (above 5 years) in stack lifetime. The tornado charts also show that H_2 cost from REP technology is most sensitive to the feedstock cost and electricity cost. The cost of feedstock and utilities being primary cost drivers is consistent with both PEM and SOEC electrolysis.

7.5. Monte Carlo Analysis Details

The parameters in **Table 13** and **Table 14** were also used to conduct Monte Carlo analysis. A triangular distribution was used with the upper and lower bounds from **Table 13** and **Table 14**. The upper and lower bounds represent the 5% and 95% points along the triangular distribution. The results of the Monte Carlo analysis provide mean, median, and mode values within 5% of the calculated baseline price. The range of production costs from the Monte Carlo analysis represents the middle 90% confidence interval of the analysis.

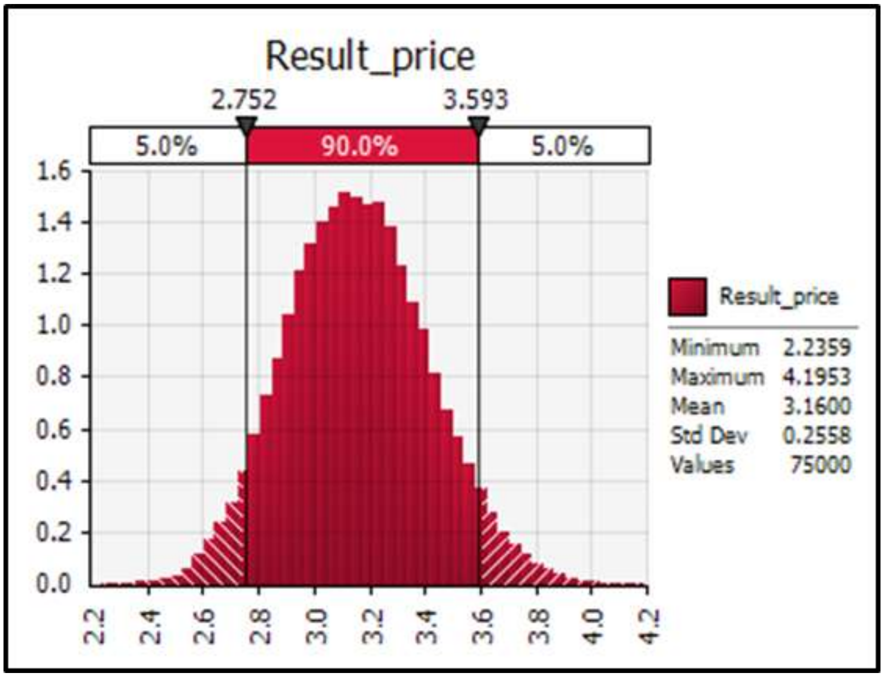

Figure 22 - Monte Carlo results for H_2 production in the Standalone REP system with a 90% confidence interval

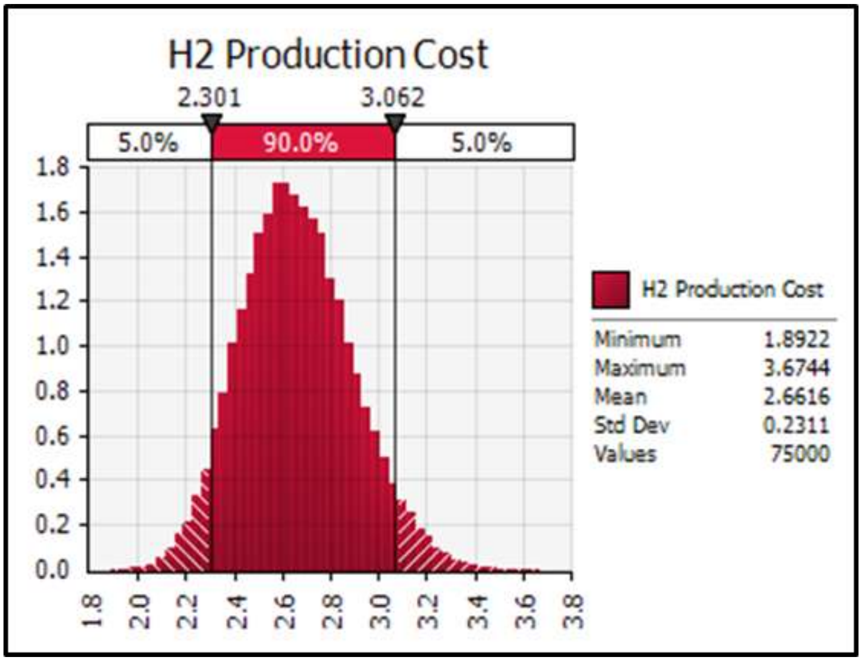

Figure 23 - Monte Carlo results for H_2 production in the Integrated REP system with a 90% confidence interval

7.6. REP Alternate Case for Very Low-Cost H_2 with 95%+ H_2 Purity suitable for non-Fuel Cell Vehicle Applications

As previously stated, the baseline case analyzed imagines an H_2 production system for fuel cell vehicles and thus (by DOE and H2A standards) must deliver H_2 at "five nines" purity and 300 psi pressure. However, this purity and pressure requirement adds considerably to cost and not all H_2 applications have those requirements. For example, the metals annealing market is well suited for low pressure, 90%+ H_2 gases mixed with methane/CO_2. Since the REP system can produce H_2 at greater than 95% purity before the use of a PSA subsystem, it is logical to analyze a system without the use of a PSA for alternative markets such as the metal annealing.

Several changes were made to convert the baseline (fuel cell vehicle based) REP systems to the alternate (non-fuel cell vehicle based) configurations:

1) The PSA was removed.
2) The reformate compressor was removed.
3) The recycle of PSA exhaust gas to the REP inlet was removed (as there is no longer a PSA). This slightly reduces system efficiency since 100% of the fuel in no longer converted to H_2.
4) A methanator was added to the REP exhaust stream to convert CO to CH_4. This consumes a small portion of the product H_2 and thus very slightly reduces efficiency.
5) The system flow rates and efficiency were recomputed to adjust for H_2 lost in the methanator and removal of the exhaust gas recycle.

Models were developed for both the Integrated and Standalone system configuration with results shown in Table 15. Approximately \$0.70/kgH_2 savings results from this alternate configuration making it one of the lowest cost production systems analyzed to date.

Table 15 - Cost breakdown of H_2 produced via REP without PSA purification

	Alternate REP Hydrogen Production Cost Contribution	
Cost Component	Integrated REP System ($/kg)	Standalone REP System ($/kg)
Capital Costs	$0.59	$0.96
Decommissioning Costs	$0.01	$0.01
Fixed O&M	$0.20	$0.30
Feedstock Costs	$0.53	$0.72
Other Raw Material Costs	$0.00	$0.00
Byproduct Credits	$0.00	$0.00
Other Variable Costs (including electric utilities)	$0.55	$0.47
Total	$1.88	$2.46
Savings compared to Baseline REP system	**Δ -$0.70**	**Δ -$0.64**

FCE has developed several other alternate configurations for the REP system. For instance, FCE has proposed a system utilizing an Electrochemical Hydrogen Compressor (EHC) to compress the H_2 and remove impurities instead of using the PSA purification analyzed in the baseline case. The methane-rich off gas can then be recycled back through the REP system to improve efficiency and recovery while incurring only the capital cost of the EHC and a low cost for the power to run the EHC. This system would be expected to have similar cost savings to what is shown in Table 6.

7.7. Projected Future Case

Normally, both *Projected Current* and *Projected Future* baseline cases are created for H2A analyses. However, given the high TRL associated with REP technology and the commercial nature of the DFC® technology it is based upon, no *Projected Future* case has been created for this analysis. It is expected that if REP technology were to be produced commercially, then the cost of production would drop in the future due to 1) typical trends of technological improvements, 2) increased production rates of the REP stacks, and 3) incorporation of easy-to-install modular design features.

7.8. Special Considerations

Data provided by Fuel Cell Energy Inc. has been provided under a Non-Disclosure Agreement to Strategic Analysis Inc. and the National Renewable Energy Laboratory. Pursuant to the agreement, some data provided by FCE is considered proprietary and does not appear in this document (for instance, a detailed process flow diagram is considered proprietary). However, the values contained in this Record are considered non-proprietary approximations that capture the function and performance of the modeled system but without revealing exact values.

8. H2A Production Pathways Future Work

8.1. Research Directions

A follow-on award has been issued to SA (supported by NREL and ANL) to continue investigation of H_2 production systems. The production pathways that will be analyzed under the new project have not yet been selected. However, it is expected that the future pathways to be analyzed will be low on the Technology Readiness Scale, similar to the TRL of the Monolithic Piston Project (**Section 6**) described above. It is also expected that delivery pathways will be analyzed.

8.2. nth Unit Research

To service H_2 vehicles across the country, clearly more than a single large production plant will be needed. Consequently the question is raised as to what the appropriate annual plant production rate to assume for the cost analyses. It is currently assumed that forecourt cases, which produce only enough H_2 to supply a small refueling station, will require approximately 300-500 systems to be built annually. Selection of the 300-500 units production rate attempts to reflect a realistic annual production rate based on station demand. Comparatively, central cases are assumed to require fewer plants to be built annually, due to the larger quantities of H_2 produced at each site. The annual production rate of central plants may range between 10 and 100+ plants, depending on the specific technology's H_2 production capacity and the level of anticipated H_2 demand growth.

The “nth unit” concept represents a sufficiently high production rate such that unit cost is no longer declining rapidly with increased rate. This is sometimes called being “at or beyond the knee” of the cost curve. In both forecourt and central cases, the n^{th} plant assumption primarily affects the projected capital cost of the system but also influences the degree of design customization.

In the previously mentioned analyses, SA and NREL have incorporated learning curves to account for the price reduction achievable by buying or fabricating various system components at the n^{th} unit rate. When possible, price quotes for multiple quantities of a given Bill of Material (BOM) component were obtained. If multiple price quotes were obtained, curve fits were used to determine the parameters to build a learning curve, as shown in Eq. 4.

$$C_n = C_o * n^a \tag{4}$$

Where:

- C_n is the cost of the n^{th} unit produced
- C_o is the cost of the first unit produced
- n is the cumulative volume of production
- a is the slope of the function when plotted on a log-log scale. Can also be expressed as the log of the learning rate over log(2).

If multiple price quotes were not obtained, a generic learning factor was applied.

The projected cost reduction from this methodology is conservative, resulting in only a ~ 1% price reduction for every doubling of production (and only a ~10% reduction between the first unit and the 500th unit). In conjunction with future contract studies, SA hopes to investigate improved methodologies to project cost specifically for system production rates in the low 100’s per year.

9. Appendix: Project Publications and Presentations

1. Moton, J. M., James, B. D., Colella, W. G., Baum, G. N., “Detailed H2A Case Study for an Emerging Technology.” *American Chemical Society and the American Institute of Chemical Engineers Techno–economic Assessments: Making Sound Investments in Alternative Energy R&D Meeting*, Washington, D.C., May 20th, 2013.
2. James, B. D., Moton, J. M., Colella, W. G., “Guidance for Filling Out a Detailed H2A Production Case Study.” Webinar Presentation to the General Public organized by the U.S. Department of Energy Office of Energy Efficiency & Renewable Energy Fuel Cell Technologies Program, Washington, D.C., July 9th, 2013.
3. James, B. D., Moton, J. M., Colella, W. G., “Description of Detailed H2A Production Case Study.” Presentation to the Hydrogen Pathways Technical Team (HPTT) and Fuel Pathway Integration Technical Team (FPITT), Pacific Northwest National Laboratory (PNNL), July 31st, 2013.
4. James, B. D., “Techno-economic Boundary Analysis of Biological Pathways to Hydrogen Production (2009).” Presentation to Biological Hydrogen Production Workshop, National Renewable Energy Laboratory, September 24th, 2013.
5. James, B. D., Colella, W. G., Moton, J. M., Saur, G., Ramsden, T., “Analysis of Hydrogen Costs from Proton Exchange Membrane (PEM) Electrolyzers.” Presentation to the U.S. Department of Energy Office of Energy Efficiency & Renewable Energy Fuel Cell Technologies Program, Washington, D.C., September 27th, 2013.
6. James, B. D., Colella, W. G., Moton, J. M., Saur, G., Ramsden, T., *PEM Electrolysis H2A Production Case Study Documentation*, report for the U.S. DOE EERE FCT program, December 2013.
7. James, B. D., Colella, W. G., Moton, J. M., Saur, G., Ramsden, T., *ADDENDUM to the PEM Electrolysis H2A Production Case Study Documentation*, report for the U.S. DOE EERE FCT program, December 2013.
8. Colella, W.G., Moton, J.M., James, B.D. “Techno-Economic Analysis of Advanced Approaches for Generating Hydrogen Fuel for Vehicles,” *Fifth European Fuel Cell Technology & Applications Conference - Piero Lunghi Conference and Exhibition (EFC2013)*, Rome, Italy, Dec. 11th-13th, 2013 (EFC13-180).
9. Colella, W.G., “Reducing Energy, Environmental, and Economic Constraints in Global Transport Supply Chains with Novel Fuel Cell and Hydrogen Technologies,” *Fifth European Fuel Cell Technology & Applications Conference - Piero Lunghi Conference and Exhibition (EFC2013)*, Rome, Italy, Dec. 11th-13th, 2013 (EFC13-178).
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15. Colella, W.G., “Reducing Energy, Environmental, and Economic Constraints in Global Transport Supply Chains with Novel Fuel Cell and Hydrogen Technologies,” *Fifth European Fuel Cell Technology & Applications Conference - Piero Lunghi Conference and Exhibition (EFC2013)*, Rome, Italy, Dec. 11th-13th, 2013 (EFC13-178).
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18. Colella, W. G., James, B. D., Moton, J. M., Saur, G., Ramsden, T., “Next Generation Hydrogen Production Systems Using Proton Exchange Membrane Electrolysis,” *Proceedings of the ASME 2014 12th Fuel Cell Science, Engineering and Technology Conference*, June 30th-July 2nd, 2014, Boston, Massachusetts, USA, ESFuelCell2014-6649 (draft conference proceeding prepared).
19. James, B. D., Colella, W. G., Moton, J. M., Saur, G., Ramsden, T., *PEM Electrolysis H2A Production Case Study Documentation*, report for the U.S. DOE EERE FCT program, Revised and Publicly Released March 2014.
20. James, B. D., Colella, W. G., Moton, J. M., Saur, G., Ramsden, T., *ADDENDUM to the PEM Electrolysis H2A Production Case Study Documentation*, report for the U.S. DOE EERE FCT program, Revised and Publicly Released March 2014.
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22. Colella, W.G., James, B.D., Moton, J.M., Saur, G., Ramsden, T.G., “Thermo-economic Analysis of Producing Hydrogen with Proton Exchange Membrane Electrolyzers,” *International Energy Agency (IEA) Advanced Fuel Cells Annex 25 Meeting No 10*, SOFC-POWER Inc. premises in Trento, Italy, April 23rd-24th 2014 (delivered remotely via webinar.)
23. Colella, W. G., James, B. D., Moton, J. M., Saur, G., Ramsden, T., “Next Generation Hydrogen Production Systems Using Proton Exchange Membrane Electrolysis,” *ASME 2014 12th Fuel Cell Science, Engineering and Technology Conference*, Boston, MA, June 30th-July 2nd, 2014 (ESFuelCell2014-6649).
24. James, B. D., Colella, W. G., Moton, J. M., Saur, G., Ramsden, T., *PEM Electrolysis H2A Production Case Study Documentation*, report for the U.S. DOE EERE FCT program, Revised and Publicly Re-Released June 2014.
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26. Colella, W.G., James, B. D., Moton, J.M., “Hydrogen Pathways Analysis for Polymer Electrolyte Membrane (PEM) Electrolysis,” *2014 DOE Hydrogen and Fuel Cells Program and Vehicle Technologies Office Annual Merit Review and Peer Evaluation Meeting*, Washington, D.C., June 16th-20th, 2014.
27. Colella, W. G., James, B. D., Moton, J. M., Saur, G., Ramsden, T., “Thermoeconomic Analysis of Proton Exchange Membrane Electrolysis Systems for Vehicle Fuel Production,” *ASME 2014 8th Conference on Energy Sustainability*, Boston, MA, June 30th-July 2nd, 2014 (ESFuelCell2014-6655).
28. Colella, W. G., James, B. D., Moton, J. M., “Technical and Economic Performance of Next Generation Hydrogen Production Pathways,” *Fuel Cell Seminar*, Los Angeles, CA, Nov. 11th, 2014.
29. Brouwer, J., Blekhman, D., and Colella, W., “Fuel Cells 101,” *Fuel Cell Seminar*, Los Angeles, CA, Nov. 10th, 2014.
30. Colella, W. G., James, B. D., Saur, G., *Technical and Economic Performance of Next Generation Solid Oxide Electrolysis Cell (SOEC) Systems for Hydrogen Production*, presentation to the Hydrogen Production Technical Team (HPTT), to be presented remotely from SA Office in Arlington, VA, Jan. 6th, 2014.
31. Colella, W. G., James, B. D., Saur, G., Ramsden, T., *H2 Pathways Status: Draft SOEC Case Values*, Presentation to Department of Energy Fuel Cell Technologies Office, presented remotely from SA Office in Arlington, VA, Dec. 10th, 2014.
32. James, B. D., Colella, W. G., Moton, J. M., Saur, G., Ramsden, T., *H2A Model and Case Study Development Overview*, Presentation to Department of Energy Fuel Cell Technologies Office, presented remotely from SA Office in Arlington, VA, Oct. 1st, 2014.
33. Colella, W. G., James, B. D., Saur, G., *Technical and Economic Performance of Next Generation Solid Oxide Electrolysis Cell (SOEC) Systems for Hydrogen Production*, presentation to the Hydrogen Production Technical Team (HPTT), presented remotely from SA Office in Arlington, VA, Jan. 6th, 2015.
34. James, B.D., DeSantis, D. *Update to HPTT: Technical and Economic Analysis of Biofermentation of Corn Stover*, Presentation to Hydrogen Production Technical Team (HPTT), February 17th, 2015
35. James, B.D., Moton, J.M, DeSantis, D., Houchins, C., Moton, J.M,, *Benchmarking Transformational Energy Technologies*, Presentation at the 227th Electrochemical Society Meeting, Chicago, IL, 5/28/2015.

36. James, B.D., Moton, J.M., DeSantis, D., Saur, G., *2015 DOE Hydrogen and Fuel Cells Program and Vehicle Technologies Office Annual Merit Review and Peer Evaluation Meeting*, Presentation at AMR, 6/10/2015.
37. Houchins, C, James, B.D., DeSantis, D, *2016 Techno-Economic Analysis: Assessing progress of emerging technologies towards meeting cost and performance targets,* Solar Fuels Workshop, University of Delaware, 3/7/2016.
38. James, B.D., Houchins, C, DeSantis, D, *2016 Techno-Economic Analysis: Water Splitting technologies and metrics*, Advanced Water Splitting Materials Workshop, Stanford University 4/14/2016
39. James, B.D., DeSantis, D., Moton, J.M., Houchins, C Saur, G., *2016 DOE Hydrogen and Fuel Cells Program and Vehicle Technologies Office Annual Merit Review and Peer Evaluation Meeting*, Presentation at AMR, 6/8/2016.
40. James, B.D., DeSantis, D., Houchins, C., *Pathways to Solar Hydrogen Technologies Workshop*, Lorentz Center, Netherlands, 6/13/2016.

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[1] Forecourt H2A models include cost calculations for compression, storage, and distribution of H_2. However, only H_2 production costs are reported in this work.

[2] The *Projected Current* Fermentation case represents an outlier in the data. That this data point represents an outlier is predictable when considering that the *Projected Current* case is based off of an existing, low-TRL laboratory bench system that is not ready for commercial use or production. The other *Projected Current* case studies in the analysis are all based off of existing technologies that have been developed to a pilot plant test scale, if not a commercial production level. Clearly, the difference in the outlier data is correlated to this difference in commercial readiness of the underlying technology. See Section 5 for further details on the analysis and development of the fermentation case study.

[3] We note that it is counterintuitive that the lower-limit of the range for the *Projected Current* cases would be lower than the lower-limit of the range for the *Projected Future* cases. However, this is explained by there being no *Projected Future REP system* case study within the data set. Section 7.7 explains why only a *Projected Current* case is analyzed.

[4] The H2A Distributed and Central Production Models 3.101 can be found at www.hydrogen.energy.gov/h2a_production.html2 The H2A Production Case Studies can be found at http://www.hydrogen.energy.gov/h2a_production

[5] Hydrogen costs reported in 2007$/kg, consistent with H2A v3 methodology which utilizes data from the *Energy Information Administration (EIA) Annual Energy Outlook (AEO) 2009 Report* (where 2007$ is the standard cost basis).

[6] Current Forecourt analysis at the 500 kg/day production level, more representative of early market stations.

[7] Uses 6.12¢/kWh average electricity price (levelized over 20 yr. life); with 450psi outlet pressure. See Table 2.

[8] Uses 6.88¢/kWh average electricity price (levelized over 20 yr. life); with 1000psi outlet pressure. See Table 2.

[9] Uses 6.22¢/kWh average electricity price (levelized over 40 yr. life); with 450psi outlet pressure. See Table 2.

[10] Uses 6.89¢/kWh average electricity price (levelized over 40 yr. life); with 1000psi outlet pressure. See Table 2.

[11] Default values described at: http://www.hydrogen.energy.gov/h2a_analysis.html#assumptions.

[12] Electrolyzer capital costs are listed in U.S. 2012 dollars (2012$) because that is the reporting year for the four electrolyzer companies. However, hydrogen cost results ($/kg) are reported in 2007 dollars (2007$), according to the standard H2A v3 methodology approved by DOE.

[13] Corresponding value of $516 for the 500 kg/day analysis.

[14] Corresponding value of $618 for the 500 kg/day analysis.

[15] Value unchanged for the 500 kg/day analysis.

[16] Average electricity price over life of plant (20 years for Forecourt cases and 40 years for Central cases). Note that the average *Projected Current* Forecourt electricity price (6.12¢/kWh) is less than the average *Projected Current* Central electricity price (6.22¢/kWh) because of the different time horizons for the investment (20 years versus 40 years). The same pattern holds for the *Projected Future* cases.

[17] Based on Energy Information Administration (EIA) Annual Energy Outlook (AEO) data.

[18] Corresponding value of 17.5% for the 500 kg/day analysis.

[19] In the early market 500 kg/day analysis, the total projected H_2 production cost is increased to $5.79/kg including a total capital cost contribution increase to $1.75/kg, and an O&M cost contribution increase to $0.66/kg.

[20] Based on case-dependent electricity prices of 6.12¢/kWh, 6.88¢/kWh, 6.22¢/kWh and 6.89¢/kWh, respectively, as per Table 2.

[21] Levelized cost of hydrogen production in the *Projected current* case assumes a 40 year plant life, 300 psi outlet pressure, and 6.24¢/kWh average electricity price (based on AEO projections for current pricing). See Table 5.

[22] Levelized cost of hydrogen production in the *Projected future* case assuming a 40-yr. plant life, 700 psi outlet pressure, and 6.89¢/kWh average electricity price (based on AEO projections for future pricing). See Table 5.

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[23] The thermo-neutral operating point refers to a cell operating voltage where ohmic losses within the cell (which releases heat) are balanced by the water splitting heat of reaction (which consumes heat). Thus the cell operates without a large temperature gradient between inlet and outlet streams. The operating voltage is 1.28 V.

[24] Heat price is based on the 40 year average of industrial natural gas price as predicted by the *EIA AEO 2009 Report*, beginning in the start-up year (2015 for Current, 2025 for Future), and an 85.7% combustor efficiency.

[25] Efficiency is defined as H_2 Product Output Energy/Input Electrical and Heat Energy. H_2 Product Output Energy is based on the lower heating value (LHV) of H_2.

[26] Current density is not used directly within the H2A analysis but is included here as a representative value to allow comparison between the *Projected Current* and *Projected Future* cases.

[27] The thermal energy cost is based on the average EIA AEO 2009 reference case costs for natural gas over the plant life, a combustion efficiency of 85.7%, and burner capital costs over the plant lifetime of ~$0.01/GJ.

[28] Stack service life represents the duration of the stack's plant operational use producing hydrogen. It differs from stack lifetime in that the stack may still have H_2 production capacity at the end of its service life.

[29] Product of plant capacity factor and linear average of % stack H_2 production rate at beginning of service year (i.e. 100%) and end of service year.

[30] Effective annual stack service replacement cost represents the constant average (over 40 year plant life) annual cost incurred to replace H_2 production capacity lost to performance degradation and to stacks taken off-line at the end of their service life.

[31] 2007 dollars are used as the cost basis (i.e., reported as 2007$/kg H_2), consistent with H2A v3.1 methodology and assumptions.

[32] H2A is a discounted cash-flow model providing transparent reporting of process design assumptions and a consistent cost analysis methodology for hydrogen production at central and forecourt facilities. H2A addresses cost scenarios where sufficiently high annual and cumulative volumes have been reached so that economies of scale for capital and unit costs have been achieved. See also at: http://www.hydrogen.energy.gov/h2a_production.html.

[33] H2A Production v3.1 Dark Fermentation Cases are at http://www.hydrogen.energy.gov/h2a_prod_studies.html. See Table 2 for a summary of case input parameters.

[34] Personal communication with NREL researchers Pin-Ching Maness and Lauren Magnusson.

[35] Davis, R. *et al. Process Design and Economics for the Conversion of Lignocellulosic Biomass to Hydrocarbons : Dilute-Acid and Enzymatic Deconstruction of Biomass to Sugars and Biological Conversion of Sugars to Hydrocarbons Process Design and Economics for the Conversion*. (2013). doi:10.2172/1107470

[36] Uses 6.89¢/kWh effective electricity price (levelized over 40 yr. life) as electricity byproduct selling price.

[37] B.D. James, G. N. Baum, J. Perez and K. N. Baum. *Technoeconomic Boundary Analysis of Biological Pathways to Hydrogen Production*. (2009). http://energy.gov/eere/fuelcells/downloads/technoeconomic-boundary-analysis-biological-pathways-hydrogen-production.

[38] Aden, A. *et al.* Lignocellulosic Biomass to Ethanol Process Design and Economics Utilizing Co-Current Dilute Acid Prehydrolysis and Enzymatic Hydrolysis for Corn Stover. (2002). doi:NREL/TP-510-32438

[39] Wooley, Robert, Mark Ruth, John Sheehan, Kelly Ibsen, Henry Majdeski, and Adrian Galvez. *Lignocellulosic Biomass to Ethanol Process Design and Economics Utilizing Co-Current Dilute Acid Prehydrolysis and Enzymatic Hydrolysis Current and Futuristic Scenarios*. (1999).

[40] Maness, P., Logan, B. 'Fermentation And Electrohydrogenic Approaches To Hydrogen Production'. Annual Merit Review (2015). Presentation. https://www.hydrogen.energy.gov/pdfs/review15/pd038_maness_2015_o.pdf

[41] Alternate processing steps are feasible. For instance, the lignin can be filtered prior to entering the fermentation reactor. This would reduce the size of the fermentation reactors and preclude lignin inhibition of the reaction. However, it would also possibly lower yield by introducing reactable feedstock losses as part of the lignin filtration process.

[42] Modeled assuming the rates of hydrogen production are proportionally increased to reach the specified final molar yield (1.16 for *Projected Current*, 3.2 for *Projected Future*) at the cost optimum time for batch

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fermentation. There exists the possibility to improve the fermentation rate and reduce the fermentation time in future systems. Such a scenario is demonstrated in the sensitivity analysis shown in **Table 4.**

[43] The lignin is still wet after filtration (modeled as 30 wt% water) and is dried prior to burning. The energy for drying is included in the system energy balance.

[44] Electricity byproduct selling price is set at 6.89¢/kWh average electricity price (levelized over 40 year life). Similar byproduct credit systems have been used in previous analyses completed by NREL.

[45] Broth concentration is defined as grams of feedstock (corn stover) per liter of slurry within the fermentation reactor.

[46] 175 g/L was chosen to align with BETO's target value of 17.5% loading in the fermentation reactor as shown in: Davis, Ryan. 'DOE Bioenergy Technologies Office (BETO) 2015 Project Peer Review'. 2015. Presentation. http://energy.gov/sites/prod/files/2015/04/f21/biochemical_conversion_davis_0315.pdf

[47] The increased broth concentration poses several processing challenges. Handing of the broth may be difficult due to increased viscosity and may require extra capital equipment to transport the slurry. Additionally, the high broth concentration may inhibit the fermentation reaction or require the use of high levels of pre-treatment chemicals which could be costly or toxic if they are not neutralized before entering the fermentation reactor. Achieving the targeted yields with high broth concentration is a significant research challenge.

[48] Maness, P.C., Logan, B. Fermentation and Electrohydrogenic Approaches to Hydrogen Production. Annual Merit Review (2015). Presentation, Slide 6. https://www.hydrogen.energy.gov/pdfs/review15/pd038_maness_2015_o.pdf

[49] Avicel® is a pure cellulose product commercially available and typically used in laboratory testing.

[50] For the purpose of calculations, biomass, Avicel®, and other complex compounds were compared based on equivalent cellulose loadings.

[51] NREL studies were conducted with a loading based on cellulose content. In order to match the cellulosic loading used by NREL for the *Projected Current* case, 12.8 g Corn Stover/L are envisioned for the fermentation loading (Corn Stover is assumed to have 39% cellulosic content. Thus, 12.8 g Corn Stover contains 5 g Cellulose).

[52] Default values described at: http://www.hydrogen.energy.gov/h2a_analysis.html#assumptions.

[53] *Energy Information Administration (EIA) Annual Energy Outlook (AEO) 2009 Report.*

[54] Fermentation capital costs are listed in U.S. 2010 dollars (2010$), which is converted from the reporting year for the 2013 NREL report upon which many of the capital cases were based. However, hydrogen cost results ($/kg) are reported in 2007 dollars (2007$), according to the standard H_2A v3.1 methodology approved by DOE. H2A v3.1 adjusts for inflation automatically when the reference years are defined.

[55] Broth concentration is not an actual input to the H2A model but is listed here because it is a defining parameter in determining capital cost and energy use.

[56] Varanasi, S., Rao, K., Relu, P. A. & Yuan, D. Methods for Fermentation of Xylose and Hexose Sugars. (2013).

[57] Electrical purchases and byproducts are reported separately for clarity but in practice only a net electrical transaction would occur.

[58] Energy purchase and byproduct are book-kept separately to ensure clarity of energy distribution. In reality, most facilities would likely use the generated energy onsite to run the plant equipment.

[59] While the sum of the *Projected Future* Case subcategory costs in Table 9 is $5.64/kg H_2, this is due to rounding of the subcategory costs and the actual H2A projected total cost is $5.65/kg H_2

[60] A *Projected Current* case was considered and modeled, but PNNL is about to complete more rigorous testing of the reforming technology. The results of such experiments would make the *Projected Current* case results irrelevant. The new data may be provided by PNNL and analyzed in the future.

[61] Literature searches suggest that CH_4 does not adsorb to dolomite-based composite sorbent. Further, literature searches have not suggested CO will adsorb to dolomite-based sorbents.

[62] Coke can be generated in many different forms. For modeling and performance computations, coke is assumed to be elemental carbon.

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[63] Hydrogen costs are reported in 2007$/kg, consistent with H2A v3.101 methodology which uses data from the *Energy Information Administration (EIA) Annual Energy Outlook (AEO) 2009 Report* (where 2007$ is the cost basis).

[64] An integrated MCFC/REP system was analyzed and, while it is designed for 1,500 kg H_2/day, it requires a 7MW MCFC plant to be co-located with the REP unit. This system does not fit with the traditional forecourt model but for size. Given that H_2 production rates are the same for this system as other forecourt models, a forecourt model was used for this analysis.

[65] The *Projected Current* case is based on current state-of-the-art laboratory-demonstrated technology, with extrapolated scale-up to and industrial process that includes high-volume manufacturing.

[66] Default values described at www.hydrogen.energy.gov/h2a_analysis.html#assumptions.

[67] The Bill of Materials and capital costs provided by FCE are considered proprietary and will not be discussed in detail here.

[68] The range of hydrogen cost is based on simultaneous probabilistic variation of the parameters (and values) shown in **Table 13** and **Table 14**. A triangular probability distribution is assumed for each parameter. Results are shown for the middle 90% of cost predictions.