Final Report: H_2 Production Pathways Cost Analysis (2016 - 2021)

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Disclaimer

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Table of Abbreviations

AEM	anion exchange membrane
AGCC	Advanced Gaseous Combined Cycle
AMR	Annual Merit Review (meeting)
ANL	Argonne National Laboratory
BoM	bill of materials
BoP	balance of plant
BoS	balance of system
BPP	bipolar plate
CapEx	capital expense
CCM	catalyst coated membrane
CVM	cell voltage monitor
DFMA®	Design For Manufacture and Assembly
DOE	U.S. Department of Energy
EERE	DOE Office of Energy Efficiency and Renewable Energy
FCT	EERE Fuel Cell Technologies Program
GDL	gas diffusion layer
gge	gasoline gallon equivalent
H_2	hydrogen
H2A	Hydrogen Analysis
HDSAM	Hydrogen Delivery Scenario Analysis Model
HPTT	Hydrogen Production Technical Team
HT	high temperature
HVDC	high-voltage direct current
kW	kilowatts
LCA	life cycle assessment
LCOE	Levelized Cost Of Electricity
LT	low temperature
NETL	National Energy Technology Laboratory
NGCC	Natural Gas Combined Cycle
NGCC-CCS	NGCC with Carbon Capture and Sequestration
NREL	National Renewable Energy Laboratory
PEC	photoelectrochemical
PEM	proton exchange membrane
PGM	platinum group metal
psi	pounds per square inch
PTL	porous transport layer
P&D	production and delivery
R&D	research and development

SA	Strategic Analysis Inc.
SOE	solid oxide electrolysis
SOFC	solid oxide fuel cell
SOFC-CC	SOFC Combined Cycle
STCH	solar thermochemical hydrogen
STH	solar to hydrogen
TEA	technoeconomic analysis
tpd	tons per day
V	volt

Foreword

In support of the DOE's Fuel Cell Technologies (FCT) mission, Strategic Analysis Inc. (SA) conducted technoeconomic analysis (TEA) studies for various hydrogen (H_2) production technologies with a specific focus on electrolyzers. This project's goal is to improve cost analysis models and increase our understanding of technical areas demonstrating information deficiencies. Our modeling process involves defining a complete Production and Delivery (P&D) pathway, assessing technology status, identifying key cost-drivers to help guide R&D direction, and generating documentation made publicly available to the technical community for improved collaboration. We utilize several cost analysis methods for determining system or hydrogen cost, including Design For Manufacture and Assembly (DFMA®) and Hydrogen Analysis (H2A). Throughout the project, we collaborated with many subject matter experts at the DOE, National Renewable Energy Laboratory (NREL), Argonne National Laboratory (ANL), and other technical experts in industry to model current, state of the art systems as well as future systems. We analyzed multiple H_2-producing electrolysis technologies: Proton Exchange Membrane (PEM) Electrolysis, Solid Oxide Electrolysis (SOE), Anion Exchange Membrane (AEM) Electrolysis, Photoelectrochemical (PEC) Electrolysis, and Solar Thermochemical Hydrogen (STCH) Production. This project culminates with our recommendations of the most promising H_2 production technologies capable of meeting the DOE's goal of producing low-cost, clean H_2.

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1 Project Overview

System-level analyses of H_2 P&D technologies are needed to support selection of the DOE's portfolio priorities. Selection occurs after careful consideration of multiple inputs, including evaluations of technical progress and H_2 cost status, as well as projections of technology timelines and benefits. Equally important is evaluation of the potential of P&D pathways to meet the DOE's FCT threshold and target cost goals of <$4 and <$2 (respectively) per gasoline gallon equivalent (gge) delivered and dispensed H_2.[1]

This effort included annual cost analyses of key remaining challenges for technology pathways within the Hydrogen Production and Delivery sub-program portfolio. The effort primarily used the H2A model to determine status improvements resulting from technology advancements. (The H2A model is a discounted cash flow Excel model, developed by NREL, used to determine the levelized cost of hydrogen ($/kg$H_2$) based on a common analysis methodology and set of baseline input assumptions.) The effort also considered cost as a function of production volume, employed error bars to illustrate uncertainties in the cost estimates, and utilized sensitivity analyses to show the potential for cost reductions.

The project was conducted from 1 October 2016 to 31 December 2021 and was composed of five budget periods of annually recuring tasks.

1.1 Project Objectives

The project entailed H_2 pathway analysis on a series of H_2 P&D pathways specified by the DOE. Specific objectives included:

1. Performing cost analysis of multiple H_2 production pathways to evaluate the potential of the projected untaxed cost of producing H_2 at the DOE's FCT goal of <$2/kg H_2.
2. Identifying key cost and performance bottlenecks of these pathways to provide support to the DOE in identifying remaining Research and Development (R&D) challenges and to evaluate progress of the DOE R&D portfolio towards meeting the DOE's H_2 production targets.
3. Conducting deep-dive analyses and optimization studies on H_2 delivery scenarios encompassing transport, storage, and conditioning to evaluate projected costs against the target untaxed delivery cost of <$2/kg H_2.
4. Supplying information developed through TEA studies of H_2 P&D pathways in support of the DOE's further life cycle assessment (LCA) studies involving these pathways.
5. Responding to the scope and topic areas as defined in the statement of work and by the DOE.

[1] During the last year of the project, DOE proposed a "Hydrogen Shot", the first of the Earthshot initiatives, with a goal of reducing the cost of clean hydrogen to $1/kg$H_2$ by 2030. https://www.energy.gov/eere/fuelcells/hydrogen-shot

2 Project Tasks

The project objectives were accomplished in the five tasks detailed below. Task 1 was conducted once at the beginning of the project. Tasks 2-5 were repeated for each P&D pathway examined, which generally were assigned annually, although overlap occurred between years.

2.1 Task 1: Low TRL Pathway Methodology

The Team established a standardized and accepted method for modeling emerging technologies, i.e., technologies with a low technology readiness level (TRL), specified by the DOE. The standard methodology provided confidence that the analyses of the lower TRL pathways are adequate to return the best possible estimates of H_2 production and/or delivery cost with appropriate statistical spreads. Under previous contract, the Team conducted a cost model validation case study in 2014 based on H_2 generation via high TRL PEM electrolysis (Central and Forecourt). That model was used this as a starting point to make appropriate modifications for lower TRL cases. The 2014 validation case used the process described in Tasks 3, 4, and 5 to model PEM electrolysis and findings were presented to five commercial suppliers who confirmed the accuracy and reasonableness of the analysis. The revised methodology developed under this task was applied to the analyses of lower TRL P&D technologies approved by the DOE and for which adequate studies and data are available to compare performance parameters and H_2 cost.

2.2 Task 2: Select Pathway, Gather Information, & Define Preliminary System Process

The goal of this task was to provide a full technical description of the P&D pathway to allow H2A production models and Hydrogen Delivery Scenario Analysis Model (HDSAM) cost models in the next task. The Team gathered information from DOE-funded projects, industry, researchers, journal articles, and other literature sources to help technically define the production or delivery pathway. The Team conducted meetings with a core group of technology experts, which in most cases included DOE-funded project participants, to develop and validate the P&D input parameters. The gathered information was augmented with engineering analysis and system performance analysis necessary to refine the P&D pathway option into a system design with sufficient detail for full capture of all significant cost parameters.

The Team applied a formal review process for each P&D system process design. Our prior experience indicated that a formal design review was well-suited to identify and correct errors or inconsistencies in the system design and the H2A/HDSAM case study. The multi-tiered review process included at a minimum the following steps:

1. Internal review by lead Team technical experts with lead Team signoffs.
2. Internal review by non-lead Team members with non-lead Team signoffs.
3. When possible, external review by outside experts and collaborators (e.g., specialized DOE staff, corporate specialists, principal investigators for P&D pathways, experts in related research areas).

Preparation of case studies was iterative with numerous reviews and re-considerations of parameters and configuration. Consequently, the output of this task was preliminary in nature as it was expected to undergo revision. However, upon completion of the internal review process and creation of a substantially complete draft of the system process design, the Team considered Task 2 closed and proceeded to Task 3.

2.3 Task 3: Create A Draft H2A Case Study With Relevant HDSAM Scenario Studies

The Team populated the H2A P&D models with the data collected in Task 2. All H2A/HDSAM cases were consistent with system designs created in Task 2. Team members worked collaboratively to ensure quality and consistency across the different H_2 P&D cases that were developed. Case quality control included a consistent and appropriate level of detail, complete documentation of performance and cost assumptions, consistent formatting, and accurate chemical engineering process flowsheets with mass and energy conserved. At a minimum, each case study included a text description of the process, list of references, process flow diagram, tornado and waterfall charts with appropriate sensitivity ranges for capital costs, process efficiency, feedstock costs, and other key cost contributors. The H2A/HDSAM case was also subjected to the design review as described in Task 2. Once all signoffs were given, the H2A/HDSAM case was deemed complete and ready for Task 4.

2.4 Task 4: Externally Vet Case Study Assumptions And Results

While the system process and H2A/HDSAM case were both reviewed internally (and possibly externally) in Tasks 2 and 3, Task 4 entailed an external review by non-Team members. After the H2A case passed its internal review, the case was transmitted to a select group of experts for review. Reviewer comments were collected and incorporated as needed by the Team, and a final Team member signoff was obtained for each case. At this point, the case was formally submitted to the DOE for review. The DOE participated in internal case study reviews. Consequently, by the time the final case study was formally transmitted to the DOE, it should have already addressed the main issues.

2.5 Task 5: Case Study Documentation

Each case study was documented within the actual H2A and HDSAM P&D model spreadsheets to provide concise sourcing and contextual information. Additionally, as requested by the DOE, a document was prepared for each case to provide a brief (~5 page) description of the case study assumptions and results. This summary document was submitted to the DOE for record-keeping purposes and may be suitable for public release, but in some cases was also accompanied by an addendum of sensitive information resulting from case study development that was kept internal to the DOE. As requested by the DOE, PowerPoint presentations or written reports summarizing the P&D pathways, key assumptions, results, and sensitivity analyses were prepared for each case and presented to the DOE, USDRIVE Technical Teams, at the FCT Annual Merit Review (AMR), and as project summary documentation for use in the DOE annual report.

SA expected 1 to 3 Case Studies to be completed each year with their start/end dates to be on a rolling basis (in response to DOE requests) rather than all commencing in month 1 and finishing in month 12 of each budget period. Consequently, the progression of tasks was applied to a progression of Case Studies with a result that the Cases were at different stages of development. Some Cases were not necessarily completed by the end of the annual reporting period. In general, annual deliverables focused on completed Cases and SA also provided a status of ongoing cases as well as part of the Go/No-Go decision review.

3 Low TRL Pathway Methodology

The Team previously evaluated numerous case studies and established a clear analysis methodology for high TRL cases. However, these validated methodologies for predicting H_2 production costs for high TRL cases are not perfectly transferable to low TRL cases. As such, under this task, the Team developed new methodologies to be used when low TRL cases are being analyzed. These methods build off and closely track the validated methods developed for high TRL cases, with the main differences being an additional emphasis on information gathering methods and system design review.

3.1 High TRL Case Development Methodology

The process for developing an H2A case with a high TRL is shown in Figure 1. This method (except for a DOE Kickoff meeting) was utilized and validated in previous years.[2] High TRL cases are, by definition, relatively mature with an existing body of experimental work already conducted. Consequently, the evaluation process begins by collecting existing information required to design the system. To facilitate the information gathering process, a DOE Kickoff meeting is initiated at the beginning of each case study. The Kickoff meeting, which has not been utilized in the past, allows further identification of experts and the opportunity to accomplish several important tasks. After identifying experts of a technology, the experts can be contacted and the project and its goals can be explained. A request for information can be provided to the experts at the meeting, while any questions or reservations from the Tech Team can be addressed. A call with DOE members in attendance can lead to an increase in expert participation. Following the kickoff meeting, a questionnaire is sent to each of the consulting experts, requesting information regarding the selected H_2 production technology. The experts then supply the Team with relevant process, performance, and cost parameters. The Team reviews the information provided, identifies and isolates any proprietary information, and melds the data into relevant, but non-proprietary, performance parameters. Identifying the sensitive data allows SA to remove sensitive data from any published works yet still use the data to achieve a realistic system model and cost projection.

[2] This validated methodology, as described here and shown in Figure 1, will be used again (including the additional DOE Kickoff meeting) if requests are made by the DOE to analyze high TRL cases.

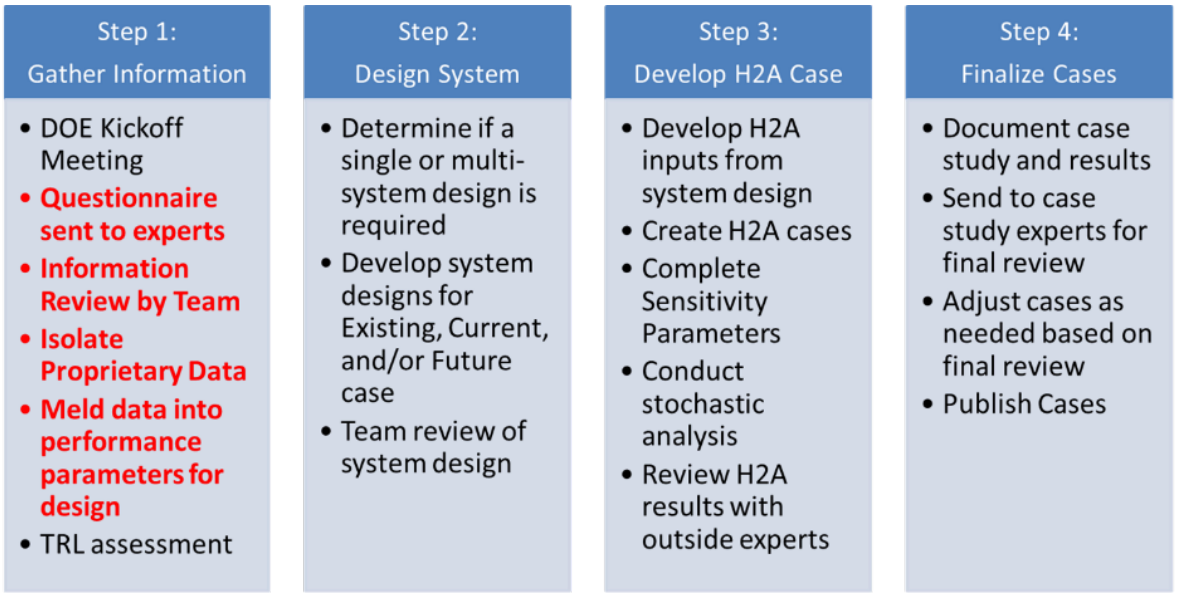


Figure 1 - High TRL H2A process workflow. Steps in red are conducted for high TRL cases but not for low TRL cases. All other steps are conducted in both high and low TRL cases.

3.2 Low TRL Case Development Methodology

In comparison, the low TRL case methodology follows the same outline as that of the high TRL methodology, maintaining the four major steps used in the high TRL analysis. However, the low TRL analysis incorporates several extra sub-steps to ensure valid and accurate results. Additionally, the low TRL cases are expected to have larger input parameter uncertainty bands, possibly more sensitivity parameters, and wider expected H_2 output price ranges within both the baseline and stochastic models and sensitivity analysis. Table 1 shows the expected differences (excluding process workflow) between low and high TRL cases.

Table 1 - Expected differences (excluding process workflow) between low and high TRL cases

Parameters	Low TRL	High TRL
Reviews by Topic Experts	3-4	1-2
H2A Project Contingency[3]	5%-25%	10-20%
H2A Process Contingency[4]	0-10%	0%
Sensitivity Range: Capital Cost (High TRL Components)	+/-25%	+/-25%
Sensitivity Range: Capital Cost (Low TRL Components)	Varies	N/A
Sensitivity Range: Operational Parameters (i.e., Power Density)	Wide	Narrow
Expected Difference in Costs between Current and Future	Wide	Narrow
Expected Monte Carlo Range	Wide	Narrow

The process workflow for a low TRL case (shown in Figure 2) is similar to the high TRL process workflow. The process begins with a kickoff meeting involving any available technology experts in conjunction with the Team and the DOE. While technology experts are consulted in both high TRL and low TRL cases, information sources will likely be more varied for low TRL cases as the systems are less likely to be integrated and specific components of the technology may need to be researched independently of one another. Literature searches and patent reviews for similar or proxy technologies that are similar to the case study technology may also need to be investigated for specific parameters or specific details regarding integration of technological components. All gathered information is reviewed and a TRL level is identified by the Team.

[3] Project Contingency is a percentage of the total initial capital investment. The project contingency inherently reflects a level of confidence for the H_2 production or delivery project as a whole (i.e., consideration is given to BoP, raw materials, waste, etc. as well as H_2 production or delivery).

[4] Process Contingency is a percentage of the total initial capital investment. The process contingency should provide a contingency budget for the actual H_2 production or delivery method itself. This contingency does not support BoP or other components of the total project.

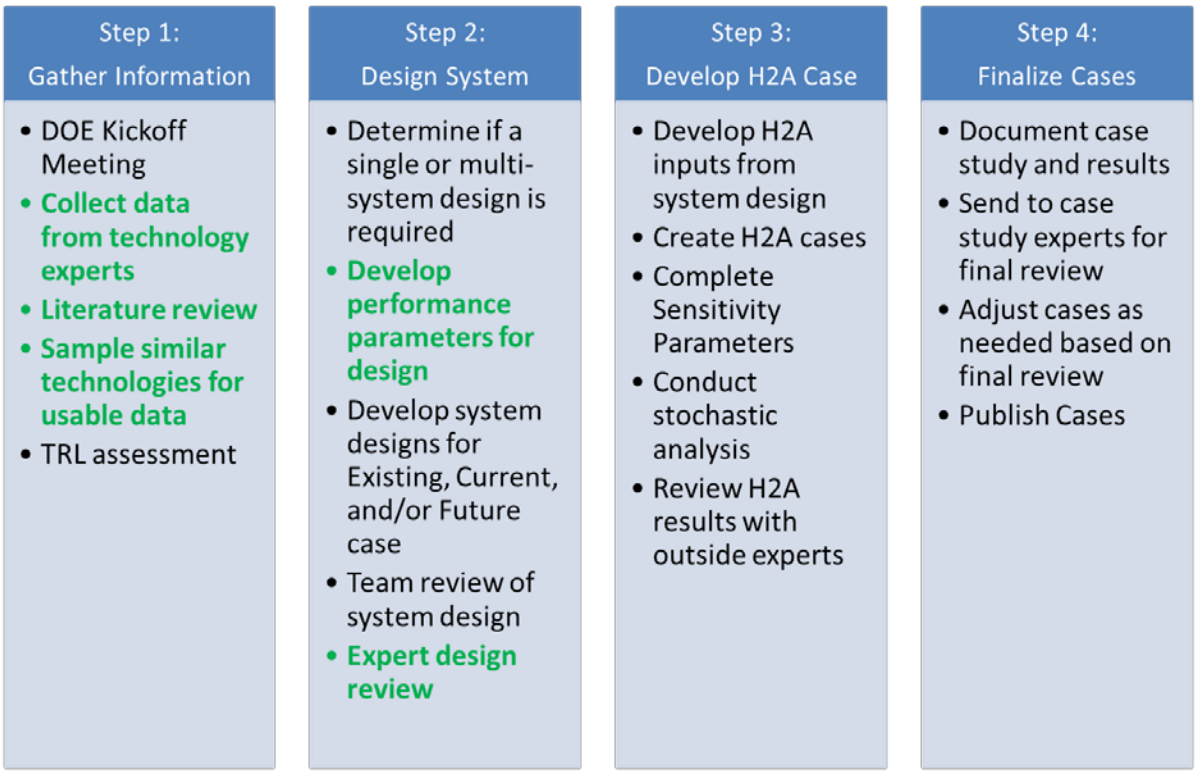


Figure 2 - Low TRL H2A process workflow. Steps in green are conducted for low TRL cases but not for high TRL cases. All other steps are conducted in both high and low TRL cases.

As with the high TRL cases, after all the required information is collected, a system design is developed. Once the system design is created and H2A input parameters are identified, the Team conducts an in-depth, thorough review of the system design and adjustments are made until the Team agrees on an appropriate system design. The design is submitted to the technology experts for their review and further adjustments are made based on any received feedback. Once the system design is finalized, the H2A cases are populated and analyzed for single parameter sensitivity studies. The sensitivity parameters used in the single parameter sensitivity studies are then used to conduct Monte Carlo analyses to identify the most probable price of production for H_2. The H2A results of the Monte Carlo analyses have an associated confidence interval, which indicates how likely are the results. The Monte Carlo results provide a range of possible H_2 prices with the upper and lower bounds of possible prices at a given confidence interval.

As defined, the process workflow is expected to provide valid H_2 production prices for the low TRL cases analyzed in H2A model. However, with any emerging process, there may be changes required to the process depending on case specifics. The Team maintains flexible practices and adjusts the process as necessary. All variations are noted as a part of the process documentation.

3.3 TRL Analysis

While the TRL definitions are clearly defined by the DOE, the assignment of a specific TRL level can be challenging. Furthermore, a singular number defining an entire system is not sufficient for a complete understanding of a technology's maturity. As such, instead of simply using a system TRL, a parameter TRL is utilized in all cases from this point forward. Assessing the TRL of individual system parameters provides a way to identify which parameters are potential bottlenecks for the system as a whole. It also allows for a prediction of which components are already integrated, which ones will integrate easily with other components, and which ones will be difficult to integrate with other components.

By separating various system parameters, such as described above, a parameter TRL can be created. To complete a parameter TRL analysis, several parameters for each technology study are identified and their TRLs are assessed. Selection of the relevant parameters varies from system to system and can take the form of a component or an attribute. Any technical parameter in the system that can be optimized or improved upon can potentially be used as a parameter. Several examples illustrating the parameter TRL concept are described below:

- The parameter of overall yield would be given a high parameter TRL if it has been already demonstrated, but a lower value if its constituent step yields have been demonstrated but never put together simultaneously to demonstrate overall yield. In this example, "integration" in the TRL definition refers to demonstration of the overall yield.
- A reaction yield may be un-optimized for the desired reaction, but a similar reaction may have a demonstrated higher (optimized) yield. In such a case, the higher yield may be included in the case study but assigned a low parameter TRL value to convey the lack of demonstration.
- Since virtually no production or delivery technology is standalone, balance of plant (BoP) will likely be a common parameter in the TRL assessment. In many cases, a high parameter TRL will be given to BoP to denote use of common, already demonstrated, BoP components.
- A specific integration parameter may be defined to isolate the exact shortcoming. For instance, an "Integration of truncated chlorophyll mutation into Organism A" parameter may be used to illustrate that, while truncation of the chlorophyll gene has been previously demonstrated, it has not yet been successfully done within the desired organism (a hypothetical Organism A, in this example).

Once identified, the parameters are assigned a TRL number (shown in Table 2). One key consideration for each physical parameter selected is the level of integration with the remaining system components. Should the component be unintegrated at the time of the study, due consideration will be given to whether similar components have been integrated into similar systems before.

Table 2 - Sample of TRL breakdown analysis. Predicted advancements are assigned to Current or Future Cases to help with system design and H2A.

Technology Element	Reference Technology Status	Existing H2A Case (2016)	Current H2A Case (2016)	Future H2A Case (2025)
Parameter 1 (Example: Yield)	4	9	9	9
Parameter 2 (Example: Catalyst Monolith)	4	9	9	9
BoP	**8**	**9**	**9**	**9**

All H2A cases must, by definition, offer a level of performance and integration suitable for full-scale plant production. Thus, all parameters are assumed to be 9 for all H2A cases; fashioning an H2A case based on a lower TRL would not have meaning as an H2A case. Thus, the parameter TRLs are only relevant in terms of serving as a technology reference point as it exists at this moment. As such, the assessment of the system TRL is easily defined after completing a parameter TRL as it is, almost by definition, equal to the lowest of the parameter TRL values.

4 WireTough Wire-Wrapped Pressure Vessel For H_2 Storage

For the successful national deployment of hydrogen-fueled Fuel Cell vehicles, a network of hydrogen refueling stations must be deployed. Bulk hydrogen storage at these stations is required to have an adequate supply for incoming light, medium, and heavy-duty vehicles. For gaseous hydrogen dispensing at 700 bar, station-based high-pressure gaseous storage is typically used for cascade filling of the vehicular tanks.[5] The cascade tanks are typically metal vessels (Type 1) with a rated pressure of ~850 bar (~12,250 psi) to ensure adequate overpressure for a full and rapid fill of the nominally 700 bar vehicle tanks. WireTough Cylinders, LLC developed an alternate design to the conventional Type 1 metal-walled tank that offered the possibility of reduced cascade system cost. Consequently, their pressure vessel design was selected for analysis with the hope of identifying a pathway to lower storage cost, leading to a lower station cost and lower H_2 price.

DFMA® analysis was completed for a model of the wire-wrapped 13,000 psi-rated pressure vessels created by WireTough Cylinders, LLC. The results of the wire-wrapped storage vessel costs may be included in future H2A Forecourt cases as part of the cascade H_2 storage system. The DFMA® process allows for process-based cost estimation of the product, in this case, wire-wrapped steel cylinders. SA modeled WireTough's process at production rates ranging from 240-3,000 pressure vessels per year (40-500 systems/year). Further, SA modeled the costs of the support structure for the H_2 storage tanks that would be used at a Forecourt site.[6]

4.1 WireTough Vessel Costs

Figure 3 shows an image of the WireTough storage vessel.[7] Vessel fabrication was modeled by SA using a DFMA® analysis methodology. Figure 4 shows a 3D model of the WireTough storage vessel. The complete vessel fabrication process is illustrated by the process flow diagram in Figure 5. The wire-wrapping process begins with a steel liner rated for approximately 6,600 pounds per square inch (psi).[8] The liner is carried by a crane to a wire-wrapping station, which combines 24 steel wires into a wire tow band and then wraps the wire tow band around the cylindrical section of the liner. As the wires are wrapped around the liner, epoxy is applied to the wires. As understood, the purpose of the epoxy is to protect the wires from corrosion, provide added strength/rigidity, and prevent wire movement. Finally, the outer layer of wires is taped with non-adhesive dry wall tape and then covered with epoxy. The end domes of the liner are not covered in the wire-wrapping process.

[5] Cascade filling refers to the process in which the H_2 source vessels are used to fill the target vessels. The source vessels (the "cascade tanks" at the station) are composed of multiple banks of vessels at different pressures (typically 3-5 different banks). The target vessels (on the vehicles) are sequentially filled from the source vessel banks in ascending pressure order. In this manner, high pressure is optimally maintained in the source vessels and the system size and cost are minimized.

[6] SA's analysis assumes a peak production rate of 500 Forecourt stations per year, with 6 high-pressure storage tanks per station (each vessel holds approximately 35 kg H_2).

[7] Courtesy of WireTough Cylinders, LLC. https://wiretough.com/wiretough-technology/

[8] For clarity within this report, the solid metal walled pressure vessel is called a liner, while the completed, wire-wrapped product is termed a pressure vessel.

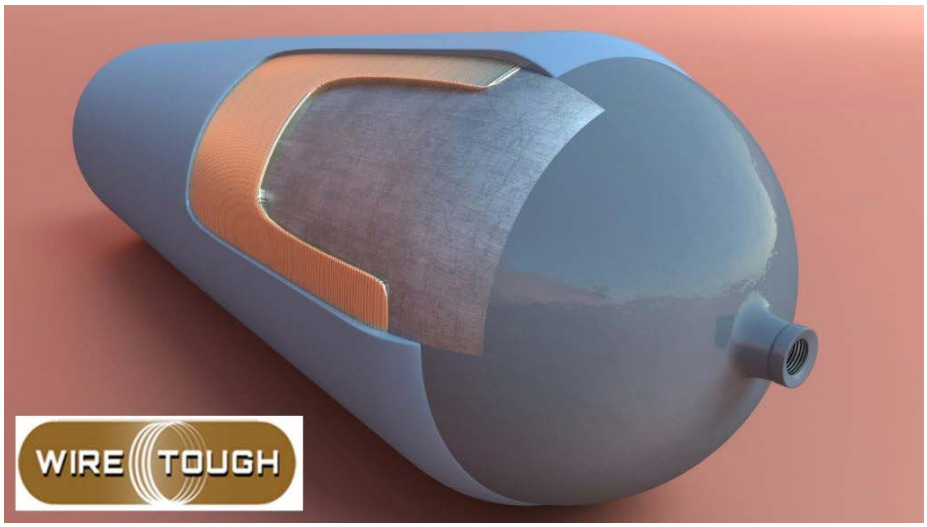


Figure 3 - WireTough storage vessel

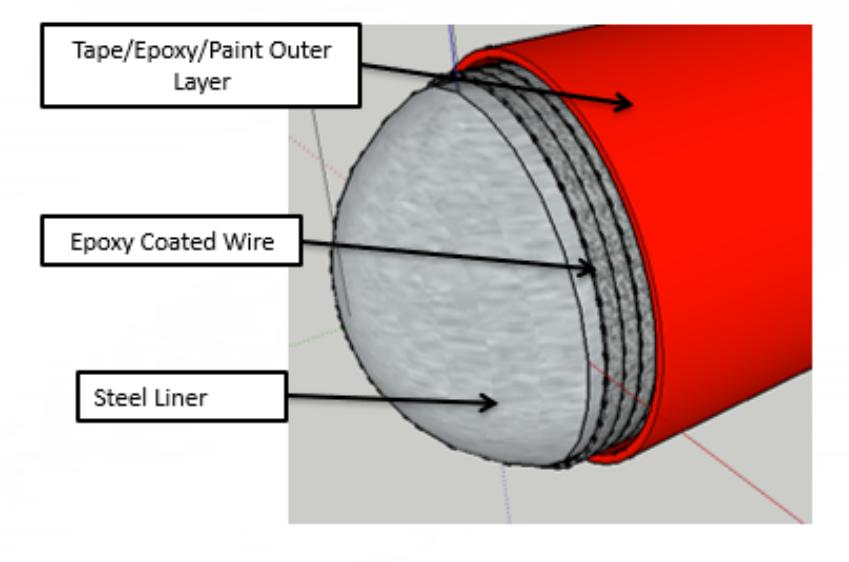


Figure 4 - 3D model of the WireTough storage vessel

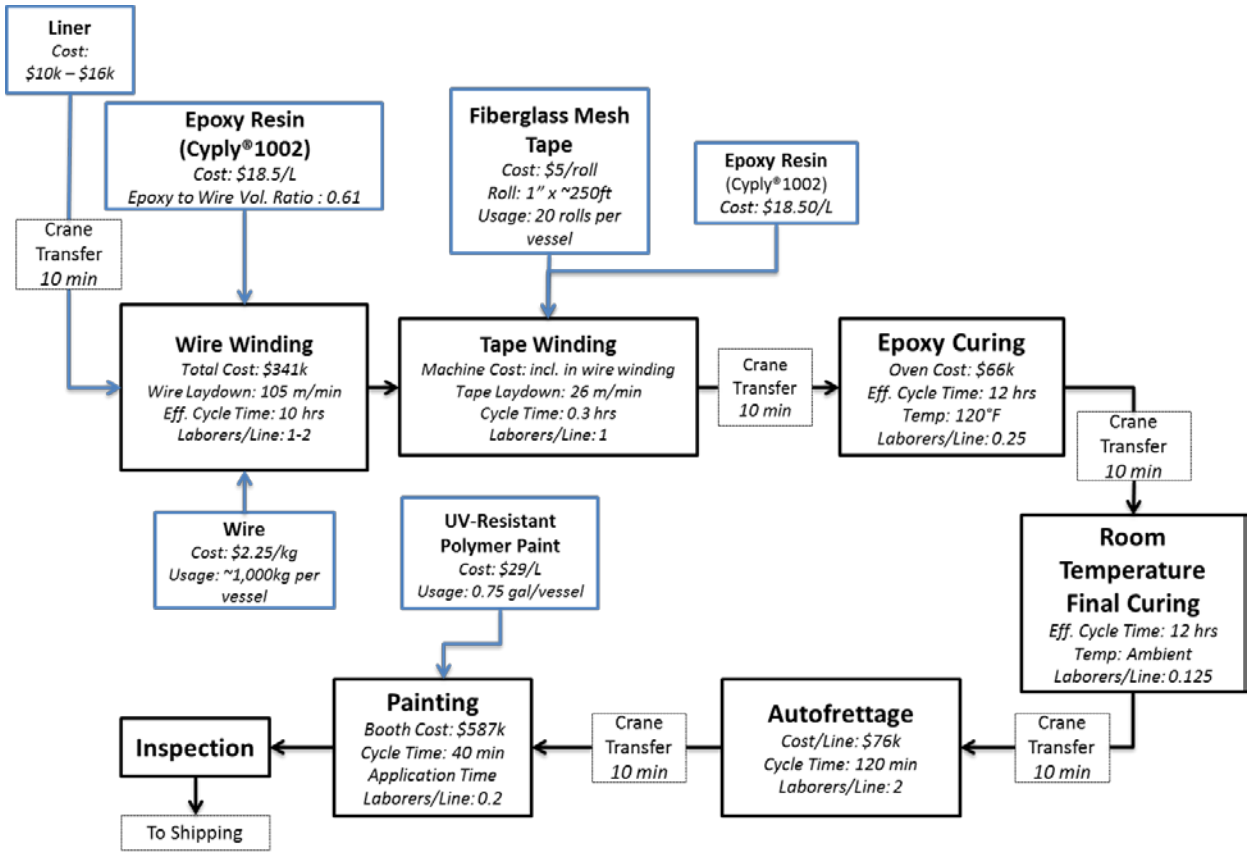


Figure 5 - Process Flow Diagram for the WireTough wire-wrapping process of pressure vessel tanks. The process flow diagram and displayed data apply to production rates of 240-3,000 tanks per year.

After wrapping the liner with wire, the assembly is sent to an oven for partial epoxy curing and is then cured at room temperature to complete the process. The pressure vessel is then put through an autofrettage process. Finally, the pressure vessel is painted with UV resistant paint.

Material costs represent the most significant costs of production, ranging from 73% to 75% of the pressure vessel cost, depending on production rate. The liner alone represents 71% of the total material cost of the vessel and, at $11,000 per unit, is by far the highest single cost element of the system. Liner cost was assessed in two ways: DFMA® cost analysis and via quotation. The results of the DFMA® and the quotations were in general agreement; a quoted price was used within the cost analysis.

A DFMA® analysis of the high-pressure ferritic steel liner was conducted based on processing details for seamless steel liners found during a literature review. The Mannesmann process was determined to be the most suitable method for forming a steel liner of the size required to meet the DOE's storage goals and is suitable for WireTough's wire-wrapping process. A simplified process flow diagram of the Mannesmann process is shown in Figure 6. The process begins with heating an ingot of steel to 1350°C in a rotary hearth furnace. The hot steel ingot is then pierced with a mandrel while the ingot is being rolled. The mandrel hollows the steel and forms a seamless tube while the rollers stretch the steel and help to maintain a cylindrical shape. Following this step, the steel is passed through several other mills and rollers

to elongate and stretch the steel into a seamless pipe of the desired length and diameter. After the pipe is formed, seamless end domes can be created by hot swaging (or a similar process). Once the ends are completed, the liner is ready for use in the wire-wrapping process.

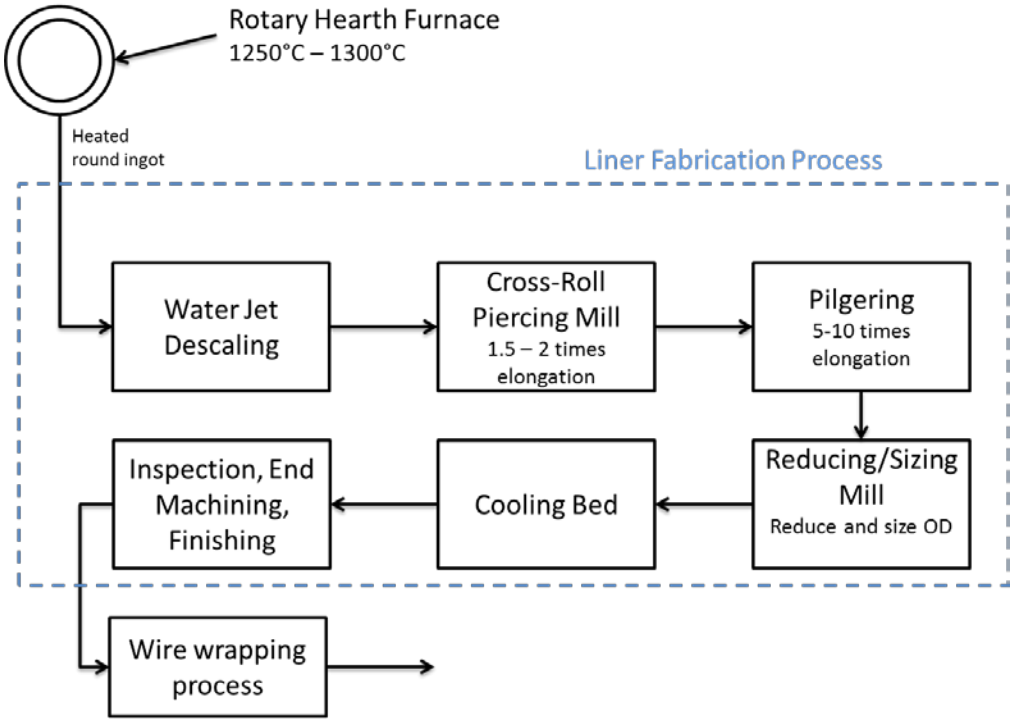


Figure 6 - Simplified Process Flow Diagram of the Mannesmann process, by which seamless steel piping can be created.

The liner analysis is based on rough approximations of equipment capital costs and factory operating values and thus should be taken as a general cost estimate subject to substantial variation. Nonetheless, the analysis provides a valuable understanding of liner cost drivers. The results suggest a liner price between $10,000 and $16,000 per unit. To corroborate these liner price findings, three liner manufacturers were solicited for price quotes. Only one manufacturer provided a quotation, and it was consistent with the DFMA® analysis.

The next largest cost to produce a wire-wrapped pressure vessel is from the wire-winding process. The wire itself makes up approximately 18% of the material cost and the winding process accounts for a significant fraction of the manufacturing costs (~40%). WireTough currently wire-wraps the vessel with a custom winding machine and then applies epoxy to the wire by hand. SA modeled this manual epoxy application process but believes it would be inefficiently slow for higher system production rates. Consequently, capital cost was added to the winding station to model automated epoxy application. Other modifications to WireTough's current low-volume production methods include the addition of a paint booth with automated spray systems for painting the finished pressure vessel (instead of hand painting) and a dedicated, on-site autofrettage station (instead of shipping the vessels to an autofrettage vendor).

The projected price (after markup)[9] of the complete pressure vessel as currently manufactured is approximately $28,266/unit (based on 1 vessel per day production). With process adjustments to account for automation and increased production rates, the cost drops to under $21,000/unit (see Figure 7). The results of a Monte Carlo analysis are presented as error bars in Figure 7. The limited variation in costs at production rates between 240 and 3,000 pressure vessels/year is a result of a constant liner cost being used at each of those production rates. With such a dominant cost being held constant at different production rates, the variation in total cost with varying production rates is minimized. The liner cost shows itself to be the most prominent cost driver in the pressure vessel sensitivity analysis (see Figure 8).

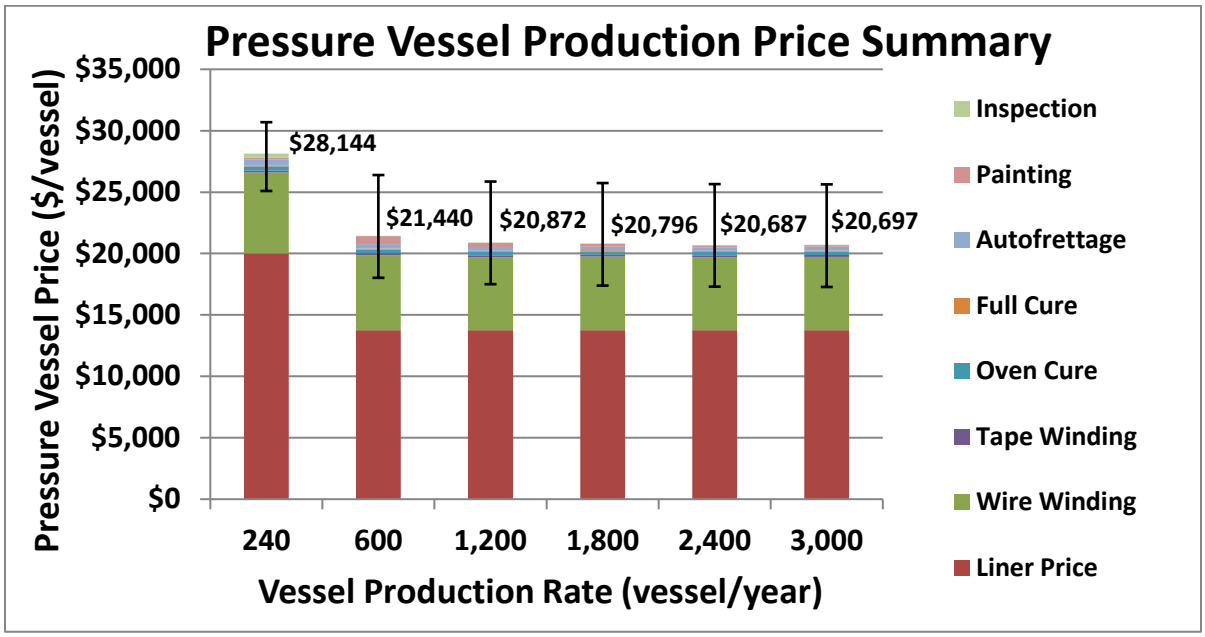


Figure 7 - Pressure Vessel Price. Error bars represent stochastic Monte Carlo analysis with a 90% Confidence Interval. Tank production rates at 240 tanks per year represent a modeling estimate matching the current process at WireTough Cylinders, LLC. All other production rates model systems with previously described system changes for increased production volume.

[9] A markup rate of 25% (at all production rates) was used to translate manufacturing cost into expected sales price (inclusive of company profit, overhead, general and administrative expenses, etc.). This rate is based on information garnered from the annual report of a high-volume pressure vessel manufacturer, Hexagon-Lincoln, and is extrapolated from the company's publicly reported gross margin and cost of goods sold. While markup rates can vary substantially company-to-company, even within an industry, Hexagon-Lincoln is judged to be an industry standard in H_2 and compressed natural gas storage vessels, and thus is thought to be an appropriate markup rate benchmark.

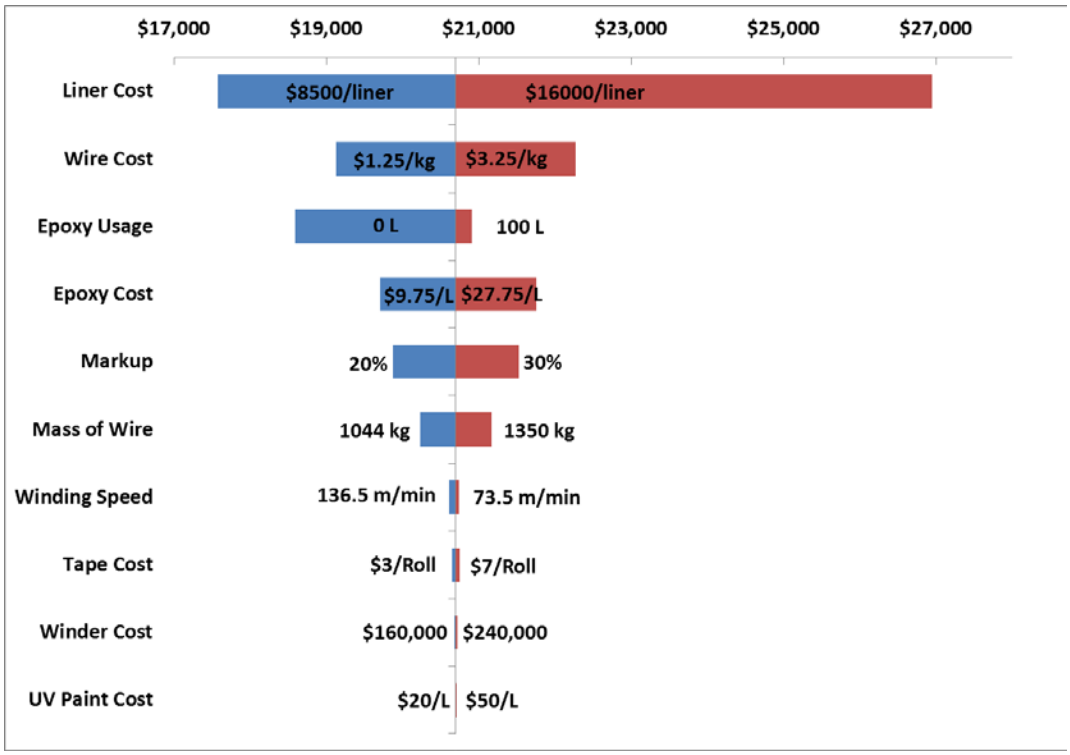


Figure 8 - Sensitivity plot for various parameters affecting the cost of production for wire-wound pressure vessels. The analysis is specifically for production at 3,000 units per year. The estimated cost for a baseline unit is $20,697.

4.2 WireTough System Cost

In conjunction with modeling costs for the pressure vessel, SA evaluated the balance of system (BoS). The storage system is based on the generic requirements of the 1,500 kg H_2/day dispensing station within the H2A model (version 3.101). The storage system is modeled as a cascade high-pressure H_2 storage system, holding six 13,000 psi rated vessels. Each vessel nominally holds approximately 35 kg H_2 for a total system capacity of 210 kg H_2. The vessels are arrayed in 3 banks (high-pressure, medium-pressure, and low-pressure) of 2 vessels each. The system supports six vehicle dispensers that operate independently (i.e., they may refuel six cars simultaneously).

A piping and instrument diagram was created (see Figure 9) to evaluate which supplemental equipment would be needed for the Forecourt cascade storage system. The supplemental equipment was then analyzed by DFMA® techniques and/or price quotation to develop a BoS cost. A mounting frame for the tanks, which consists of a steel beam structure capable of supporting a unit measuring 2 vessels wide by 3 vessels tall, was included in the BoS cost. The mounting frame cost includes materials and assembly, as well as a small allowance for capital costs for purchasing necessary equipment for assembly. A fuel system controller is not included in the cost analysis as it is assumed to be included in the dispenser cost. Finally, an installation cost is applied to the system and consists of cost categories for delivery of the tanks and frames, construction of a concrete pad and retaining wall for the storage assembly, and electrical installation and controls testing. A 15% contingency factor is added to cover non-enumerated costs. A further markup of 25% is added for profit, overhead, and other business-related expenses. The BoS costs for the storage system are detailed in Table 3**.**

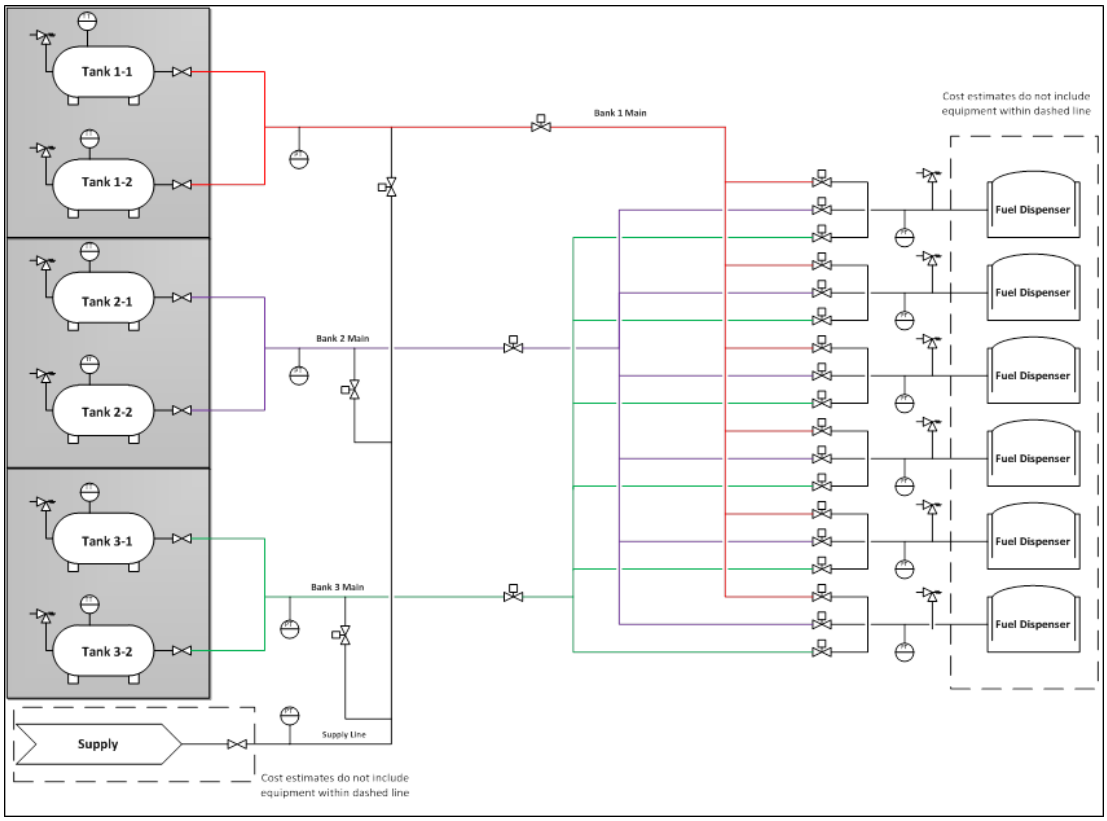
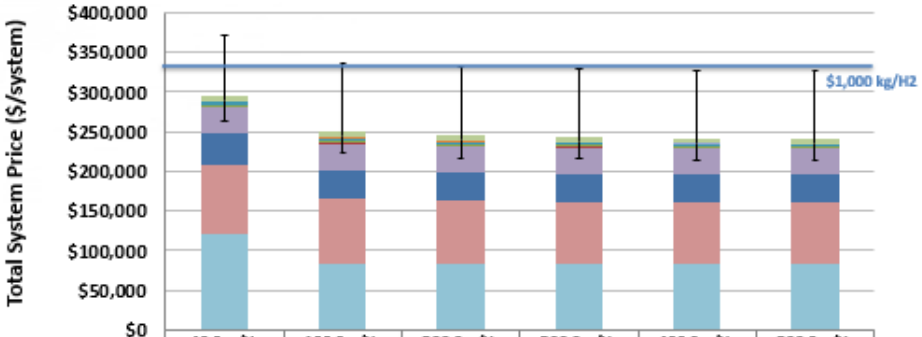


Figure 9 - Piping and Instrument Diagram for the H2A Forecourt system

Table 3 - BoS costs for the storage system

		Annual Manufacturing Rate					
	tanks/system	6	6	6	6	6	6
Annual System Production Rate	system/yr	40	100	200	300	400	500
Annual Tank Production Rate	Tanks/yr	240	600	1,200	1,800	2,400	3,000
BOS							
H_2 Solenoid	**Per System**	**$29,160**	**$29,160**	**$29,160**	**$29,160**	**$29,160**	**$29,160**
Number of $H_{2\ Solenoid}$ Valves	Per System	24	24	24	24	24	24
H_2 Solenoid Valve Cost per unit	Per System	$1,215	$1,215	$1,215	$1,215	$1,215	$1,215
H_2 Solenoid Cost	Per System	$29,160	$29,160	$29,160	$29,160	$29,160	$29,160
Tank Instruments	**Per System**	**$9,095**	**$8,111**	**$7,564**	**$7,304**	**$7,141**	**$7,026**
Manual Tank Valves	Per Tank	$437	$424	$415	$409	$405	$402
High Pressure Temperature Transducer	Per Tank	$478	$327	$245	$207	$184	$168
Pressure Relief Device (PRD)	Per Tank	$601	$601	$601	$601	$601	$601
Temperature Transmitters	Per System	**$3,900**	**$3,900**	**$3,900**	**$3,900**	**$3,900**	**$3,900**
# per System	# per system	**6**	6	6	6	6	6
Total Transmitters	units/year	240	600	1200	1800	2400	3000
Price per Unit	$/unit	**$650**	$650	$650	$650	$650	$650
Other (tubing, mount, etc.)	Per System	**$18,882**	**$16,420**	**$15,069**	**$14,438**	**$14,047**	**$13,794**
Tubing & Fittings	Per System	$7,517	$6,800	$6,393	$6,198	$6,076	$5,989
Pressure Relief Device (PRD)	Per System	$3,606	$3,606	$3,606	$3,606	$3,606	$3,606
High Pressure Transducer	Per System	$4,782	$3,269	$2,452	$2,072	$1,839	$1,676
Manual Defuel Valve	Per System	$982	$751	$624	$567	$532	$528
Mounting Frame	Per System	$1,995	$1,995	$1,995	$1,995	$1,995	$1,995
Contingency (15%)	Per System	**$9,156**	**$8,639**	**$8,354**	**$8,220**	**$8,137**	**$8,082**
Markup		**25%**	25%	25%	25%	25%	25%
BOS Subtotal	$/System	$87,741	$82,787	$80,059	$78,778	$77,982	$77,453
BOS Subtotal	$/Tank	$14,623	$13,798	$13,343	$13,130	$12,997	$12,909
BOS Subtotal	$/kg H_2	$412	$389	$376	$370	$366	$364

Figure 10 shows the storage system price (and price breakdown) at various production rates. At a high production rate of 500 systems per year, the total system price is $235,664 per system. The key cost drivers are the liner price, BoS items, wire winding, and system installation. Figure 11 shows the storage system sensitivity study for production at 3,000 vessels per year. This study used generous percentages for each variable. The liner cost, wire cost, and epoxy usage are the key cost drivers.



	40 Sys/Yr	100 Sys/Yr	200 Sys/Yr	300 Sys/Yr	400 Sys/Yr	500 Sys/Yr
System Assembly	$1,330	$1,019	$915	$881	$915	$895
Inspection	$79	$79	$79	$79	$79	$79
Paint Booth	$825	$3,919	$1,983	$1,338	$1,015	$822
Hydro Test	$3,028	$2,176	$1,892	$1,797	$1,749	$1,835
Final Cure	$496	$496	$496	$496	$496	$496
Oven Cure	$1,983	$1,812	$1,640	$1,583	$1,640	$1,606
Tape Winding	$976	$975	$975	$975	$975	$975
System Installation	$33,128	$33,128	$33,128	$33,128	$33,128	$33,128
Wire Winding	$39,566	$36,689	$35,671	$36,010	$35,671	$35,875
Balance of System (BOS) Items	$87,740.62	$82,787	$80,059	$78,778	$77,982	$77,453
Liner Price	$120,000	$82,500	$82,500	$82,500	$82,500	$82,500

Figure 10 - Storage System Price

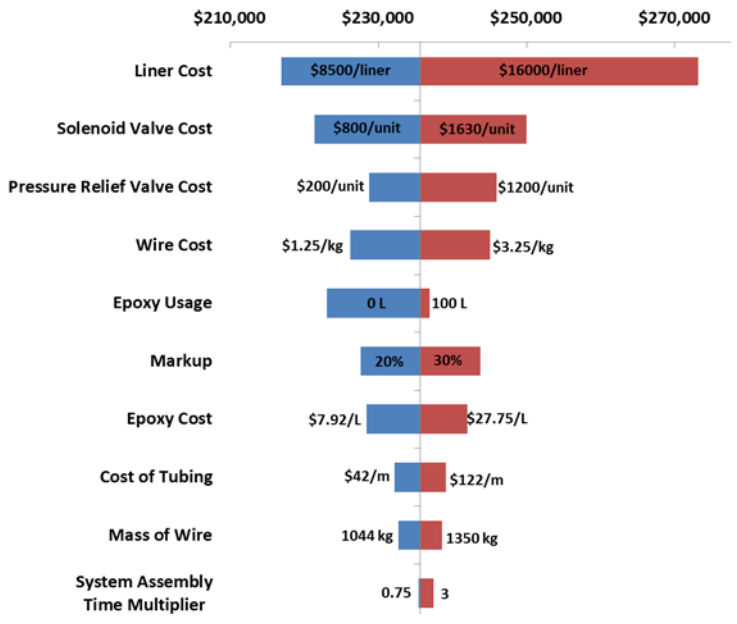


Figure 11 - Sensitivity plot for various parameters affecting the cost of production for the storage system. The analysis is specifically for production at 3,000 vessels per year.

Figure 12 shows a comparison of the cascade storage price and DOE targets. A complete installation of the WireTough system would cost less than the DOE's 2015 target (the DOE's target only includes tank, support frame, painting, cleaning, and testing). The estimated cost for just the WireTough tank falls just below the DOE's 2020 target.

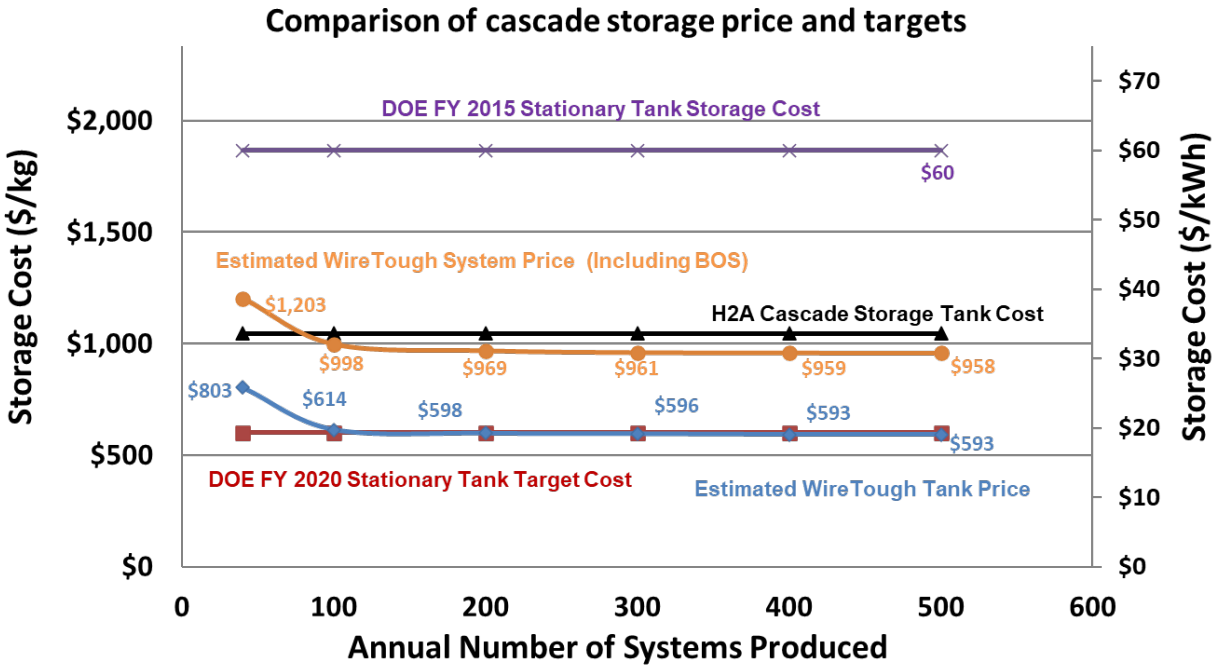


Figure 12 - Comparison of the cascade storage price and DOE targets

5 Proton Exchange Membrane (PEM) Electrolysis

5.1 PEM Electrolysis Overview

Liquid alkaline electrolysis has been the dominant electrolysis technology over the past 100 years but is being challenged by Proton Exchange Membrane (PEM) electrolysis due to its potential for low cost, high current density, and superior intermittent-operation characteristics.[10] PEM electrolysis is characterized by use of solid polymer membrane electrolyte across which hydrogen ions (H^+) are transported as part of the water-splitting process. Figure 13 shows the working principle of a generic PEM electrolyzer as well as the redox reactions.

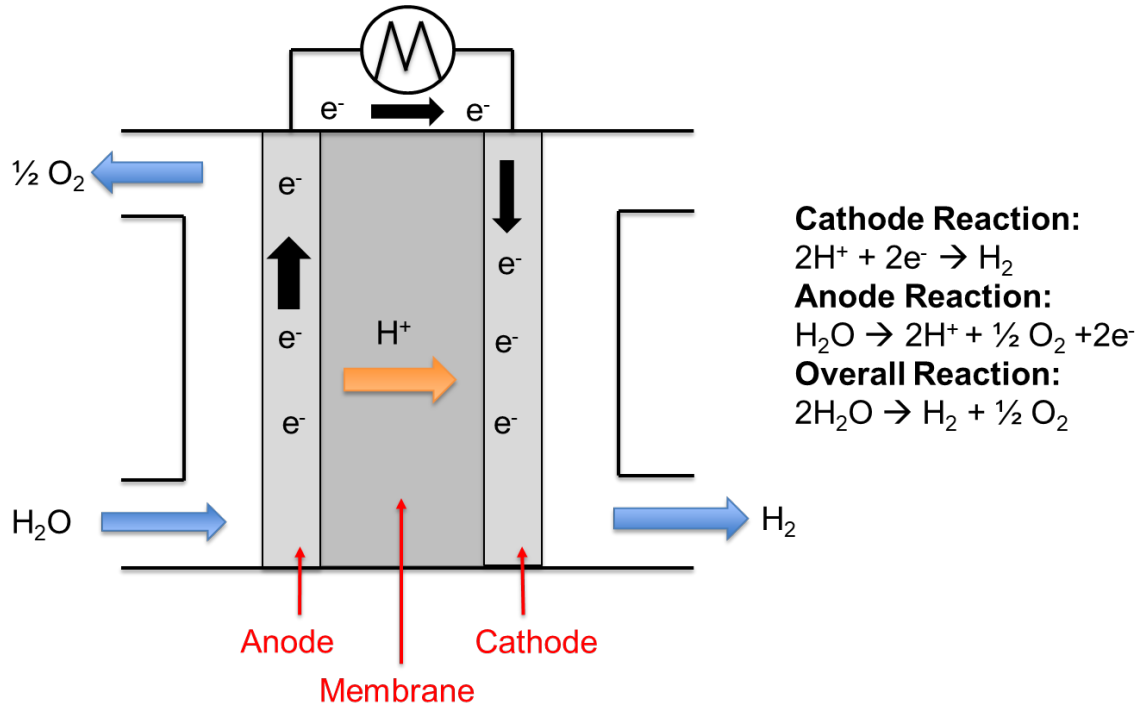


Figure 13 - Working principle of a generic PEM electrolyzer

For this study, we used our previous (2014) PEM electrolysis case study as the basis for our analysis. The PEM electrolyzer system is broken down into 3 major parts: PEM Stack, Mechanical BoP, and Electrical BoP. One module of the mechanical BoP is modeled for Distributed cases. Given the large size of Central plants, it is projected that multiple modules of the mechanical BoP will be required. This will allow for maintenance and partial shutdowns of a module without a complete loss of plant production. Electrical BoP consists primarily of the transformer and rectifier needed to convert the AC input electricity to DC current used by the stacks. A questionnaire was sent to seven PEM electrolyzer companies to solicit information on the current and future status of PEM electrolyzer design and operating parameters. Responses from four companies were collated to identify representative parameter values,

[10] Electrolyzers powered by intermittent electricity supplies such as solar and wind are a vital element in the anticipated renewable hydrogen strategy.

commonalities within the industry, and trends between current and future systems. The process flow diagram for the PEM electrolyzer system is shown in Figure 14.

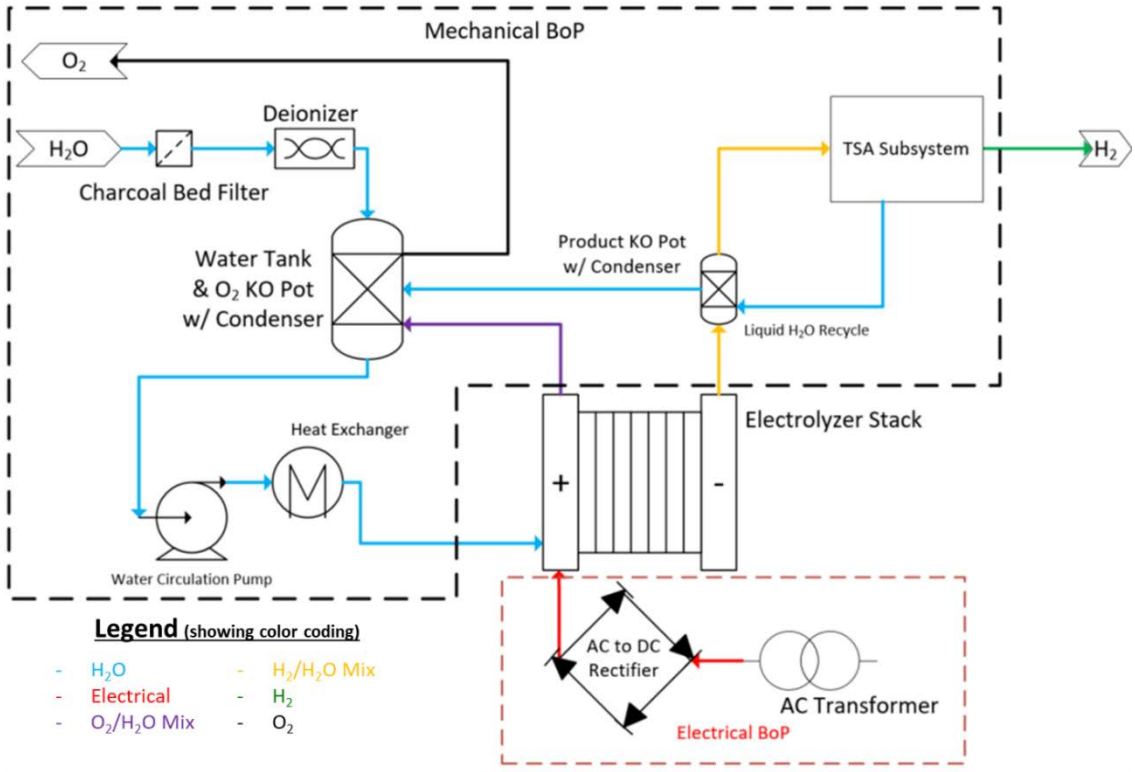


Figure 14 - Process Flow Diagram for the PEM electrolyzer system

Polarization operating points for both Current (2019) and Future (2035) cases were identified from the collected data. A mathematical model described by Hao et al[11] was used to develop polarization curves that pass through the operating points (Equation (**1**)).

$$E(V) = E_o + b * ln\left(\frac{i + i_{loss}}{i_{loss}}\right) + R * i \quad (1)$$

While operating points (specifically, current density and cell voltage) are not specifically required to complete an H2A model, creation of a full polarization curve allows us to conduct an optimization analysis to determine the operating point (impacting both electrolyzer capital cost (size) and efficiency (electricity usage)) leading to the lowest H_2 cost.

Performance degradation rates were incorporated into the cost analysis. Explicit effects from degradation had been omitted from the 2014 H2A PEM electrolysis case study (although degradation was included

[11] Hao, D., Shen, J., Hou, Y., Zhou, Y., Wang, H., "An Improved Fuel Cell Polarization Curve Model based on Review Analysis", International Journal of Chemical Engineering, 2016. https://doi.org/10.1155/2016/4109204

within the SOE study). Degradation rates are asserted on an aggregate basis and were not summed from their mechanistic sources (such as membrane resistance change, loss of catalyst activity, etc.) The analysis assumes constant voltage operation (rather than constant current density operation as was assumed in the previous PEM electrolysis case study). Consequently, all stacks and stack costs are oversized to achieve an average production rate of the target values, 1.5 tons per day (tpd) for Distributed and 50 tpd for Central.

Costs for the various system components were collected in the questionnaire. Questionnaire data were of uneven quantity and sometimes contradictory. Consequently, multiple methods were combined to project internally-consistent component costs for the case studies: questionnaire data, equipment supplier quotes, TEA, DFMA® analysis, and 3rd party equipment cost models (textbook correlations, ASPEN/Hysys® Cost Estimator.

5.2 PEM Electrolysis Stack Cost DFMA® Analysis

In this study, SA was tasked with evaluating the cost to manufacture a PEM electrolysis stack at multiple production rates, system sizes, and stack sizes. Although the PEM electrolysis H2A case has been submitted and finalized, the DOE desired to understand the lowest production rate at which most manufacturing economies of scale were achieved (the knee in the curve). SA utilized existing DFMA® models to develop a PEM electrolyzer stack cost model. Comparisons to NREL's 2019 cost analysis were also conducted to highlight areas of disagreement. The results of this analysis were reviewed by the Hydrogen Production Technical Team (HPTT). Although these results were not used for the published H2A cases, they further support the stack capital cost used within the published H2A cases.

5.2.1 SA PEM Electrolysis Stack Design Basis

SA reviewed current and future manufacturing capacities for PEM electrolyzer companies to establish a baseline stack design. Table 4 compares multiple PEM electrolyzer company system sizes, sales, and operating conditions with SA's selected design for this cost model. In some cases, SA took the midpoint among all the companies. For the manufacturing capacity, SA estimated 2 GW/year as the maximum production volume out to 2030; although there was feedback that SA may want to consider higher production volumes leading to 50 GW/year production. For purposes of this analysis, we are equating manufacturing capacity and actual/projected manufacturing rate.

Table 4 - PEM electrolyzer company information compared to SA's modeled stack design

	Unit	Company 1	Company 2	Company 3	Company 4	Company 5	Company 6	Company 7	SA Model
Manufacturing Capacity (current, estimated)	MW/year		~25		>30	40	>2	60	10
Manufacturing Capacity (projected)	MW/year	500 (2021) >1,000 (2024)	300 (2021) 1,000 (2024)	360 (2024) 1,000 (2030)	500 (2025)				10,000
Max Stack Size	MW	5	2	0.75	2.5	1.25	0.25	1.6	1, 2, 2.5, 5
Largest System	MW	5	24	6	20	13	1	3.2	1, 4, 10, 100
Current Density	A/cm^2	3	3	2		1.8			2
Cell Voltage	V	1.9	2.0	<2.05		1.85 - 1.9			1.9
Operating Pressure	bar	40	20 - 50	35	40	20			35

The cost was evaluated for four different size systems (1, 4, 10, and 100 MW) at six production volumes each (10, 20, 40, 60, 80, 100 systems per year). Stack sizes ranged between 1 and 5 MW depending on the size of the system, although the stack size is easily changed to see how this may affect cost.

The stack was modeled with rectangular cells having active area sizing scaled linearly with the stack power sizing. This scaling came from Giner's publicized stack sizing [12] as seen in Figure 15. The resulting range in active area per cell for 1 MW up to 5 MW stack power is 740 and ~3,000 cm^2 per cell, respectively. Figure 16 shows a cross-sectional view of a representative single electrolysis cell conceptualized by SA. Inspiration for this design came from two main sources: 1) NREL's 2019 report on manufacturing cost of PEM electrolyzers,[13] and 2) a 2019 journal article describing low temperature (LT) electrolyzer designs.[14] Within Figure 16, annotations in blue describe each component material, thickness, and manufacturing process while annotations in red highlight possible changes to the model in the future.

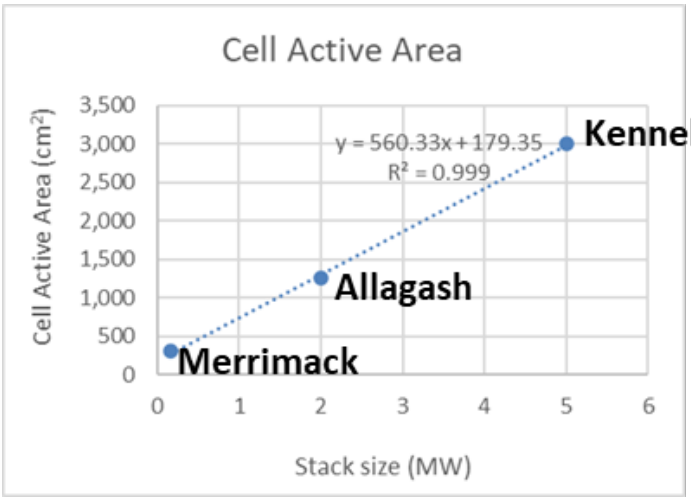


Figure 15 - Electrolyzer cell active area for each of Giner's stacks

[12] https://www.ginerelx.com/electrolyzer-stacks

[13] Mayyas, A., Ruth, M., Pivovar, B, Bender, G., Wipke, K., "Manufacturing Cost Analysis for Proton Exchange Membrane Water Electrolyzers", Technical Report by National Renewable Energy Laboratory, August 2019.

[14] Ayers, K., Danilovic, N., Ouimet, R., Carmo, M., Pivovar, B., Bornstein, M., "Perspectives on Low Temperature Electrolysis and Potential for Renewable Hydrogen and Scale", Annual Review of Chemical and Biomolecular Engineering, 10:219-239, 2019.

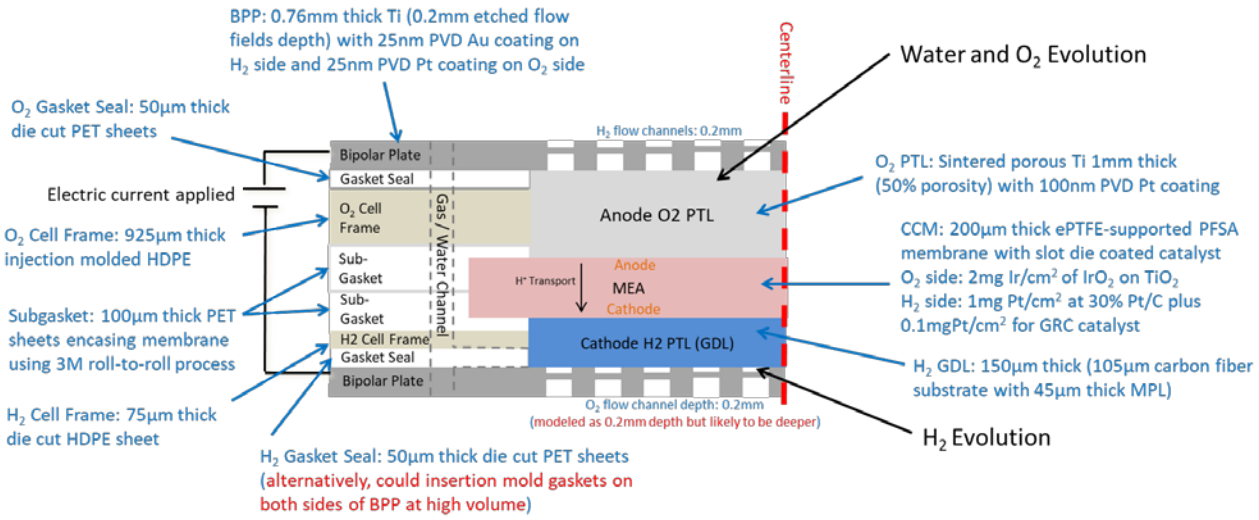
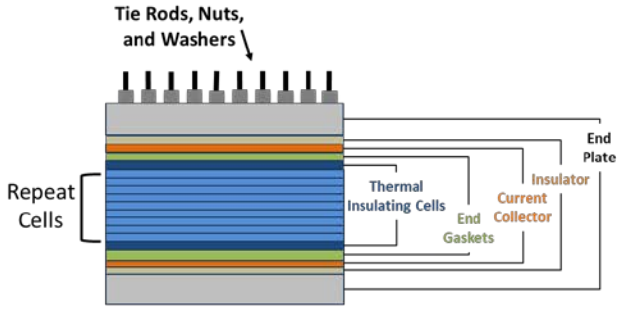


Figure 16 - Cross-sectional view of a single PEM electrolysis cell with descriptions of all modeled components

The stack design described in Figure 17 shows the repeat cells, thermal insulating cells (non-active cells), end gaskets, current collector, electrically insulating plate, and end plates. Not shown in the figure is a cell voltage monitor (CVM) to track possible cell reversals or poor-performing cells.



Dummy Cells: two dummy cells per stack (one at each end) using four 150µm GDLs per dummy cell

End Gaskets: 45µm thick screen-printed polyolefin elastomer

Current Collector: 4mm thick copper stamped sheeting

Insulator: 1mm thick injection molded HDPE

End Plate: ~4cm thick stainless steel machined plates

Cell Voltage Monitor (CVM): Between 2 and 4 cells per sensor

Figure 17 - Full PEM stack design showing additional components as well as repeat active cells

5.2.2 SA PEM Electrolysis Stack Cost

Cost results ($/kW$e_{stack}$) at all production volumes are depicted in Figure 18 for each system size in systems per year. In comparing different system sizes there may look like an extreme difference in cost between a 1 MW system and a 100 MW system when looking at production rate in systems per year. However, this is due to the 1 MW system being produced at fewer MW per year at the same systems per year

production. When comparing cost on a MW per year basis as seen in Figure 19, the different MW-rated size systems tend to fall on a similar curve. The cost estimate for manufacturing the stacks is between about $200/kW and $550/kW.

Based on these estimates, the bulk of the cost reduction due to economies of scale occurs around 100-200 MW/year production. This is highly influenced by the type of manufacturing system chosen for low volume production and whether a 3rd party vendor is used to manufacture a component. Many of the modeled components are assumed to be manufactured by a third-party (job-shopped) and thus a minimum of 30% machine utilization is assumed for these components even though the machine may only use a fraction of a percent for a particular component in low volumes.

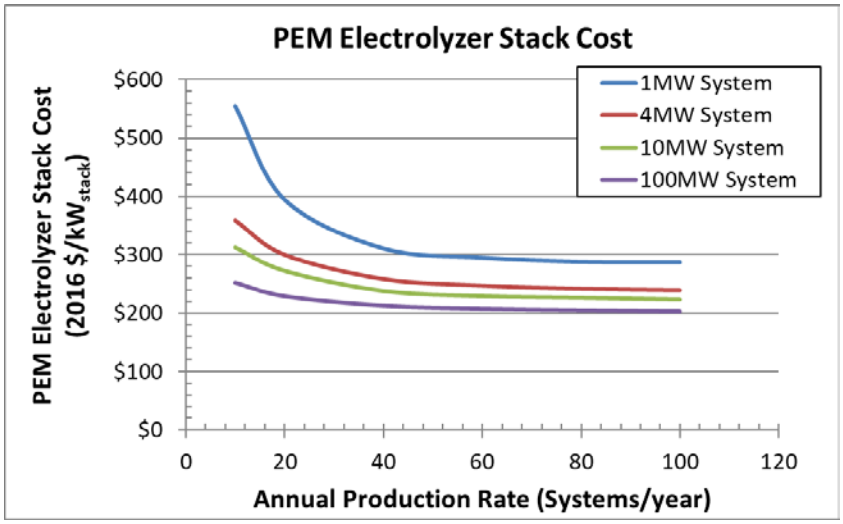


Figure 18 - PEM electrolyzer stack cost over production rate in systems per year for 1, 4, 10, and 100 MW size systems

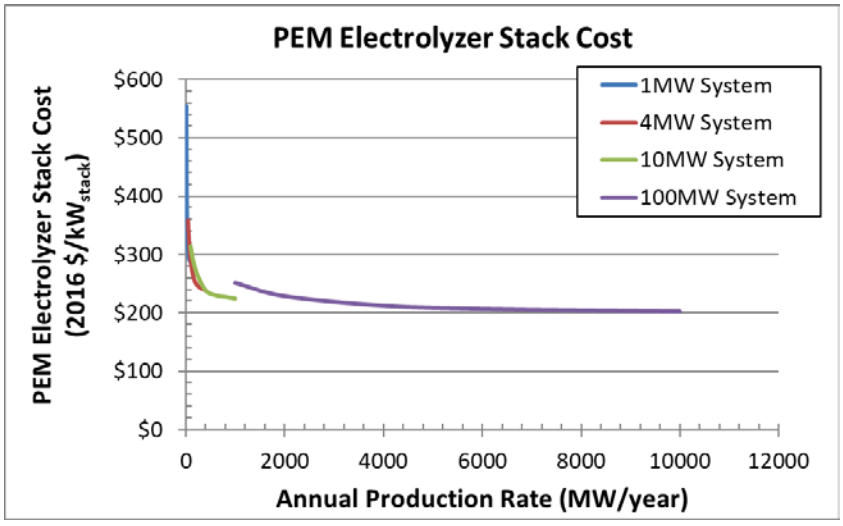


Figure 19 - PEM electrolyzer stack cost over production rate in MW per year for 1, 4, 10, and 100 MW size systems

Figure 20 and Figure 21 show the stack cost breakdown at all production volumes for 1 MW and 100 MW systems, respectively. The stack cost for both system sizes at all volumes is dominated by catalyst cost followed by bipolar plate (BPP), membrane, and porous transport layer (PTL) costs.

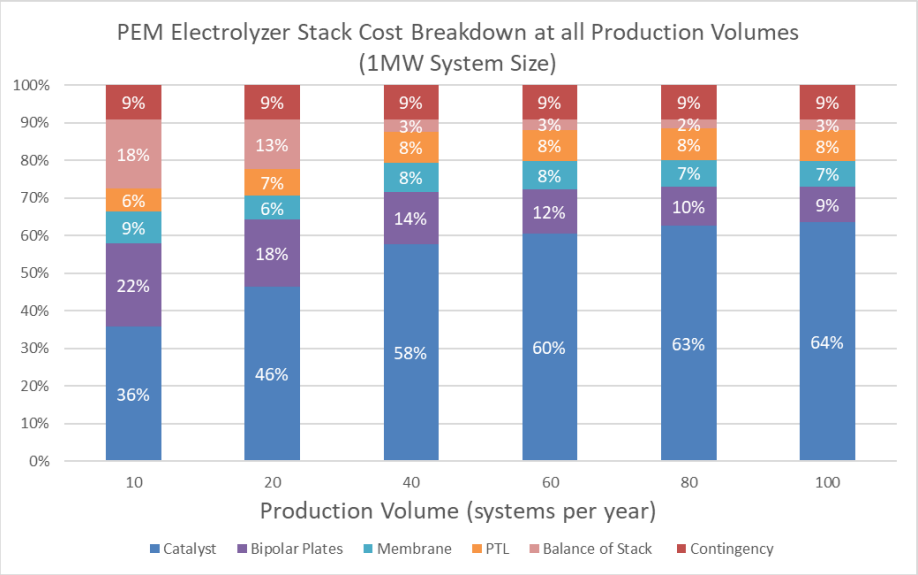


Figure 20 - PEM stack cost breakdown by production volumes for 1 MW system

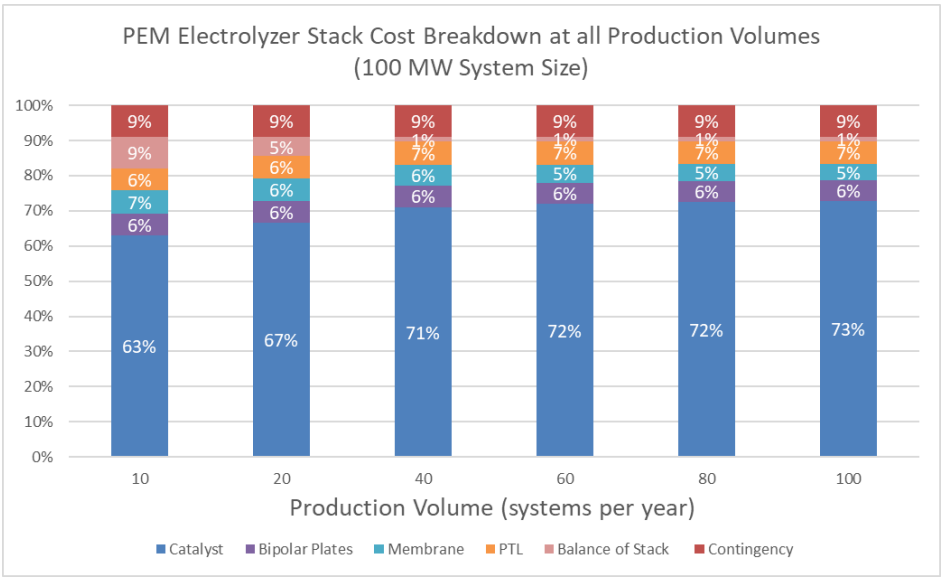


Figure 21 - PEM stack cost breakdown by production volumes for 100 MW system

5.2.3 Comparison of SA and NREL PEM Stack Costs

Given the very similar analysis conducted by NREL in 2019, SA attempted to reconcile cost differences between NREL and SA stack cost estimates for a 1 MW system (the single stack size for which there was

study overlap). SA's cost estimate for the 1 MW stack is roughly 1.5x NREL's cost estimate at a low volume of 10 MW/year (see Figure 22). The majority of this cost difference comes from the BPP and catalyst assumptions. NREL uses a 197-mil thick stamped stainless steel plate for the BPP while SA models a 30-mil thick etched titanium plate. For the catalyst, NREL uses 0.7 mgPt/cm^2 of Pt catalyst on the O_2 electrode (anode) and 0.4 mg/cm^2 of 1:1 Pt-Ir catalyst on the H_2 electrode (cathode) while SA uses 2 mgIr/cm^2 on the anode and 0.24 mgPt/cm^2 of Pt/C on the cathode. Additionally, NREL used an Ir pricing of $700/tr.oz while SA used a price of $5,000/tr.oz. If the amounts of loading for Pt and Ir are aligned and the Ir pricing is aligned to the NREL values, then the SA cost estimate is reduced significantly, and SA projections approach the NREL values as shown in Figure 23. Additional information on the differences between NREL's and SA's designs are listed by component in Table 5.

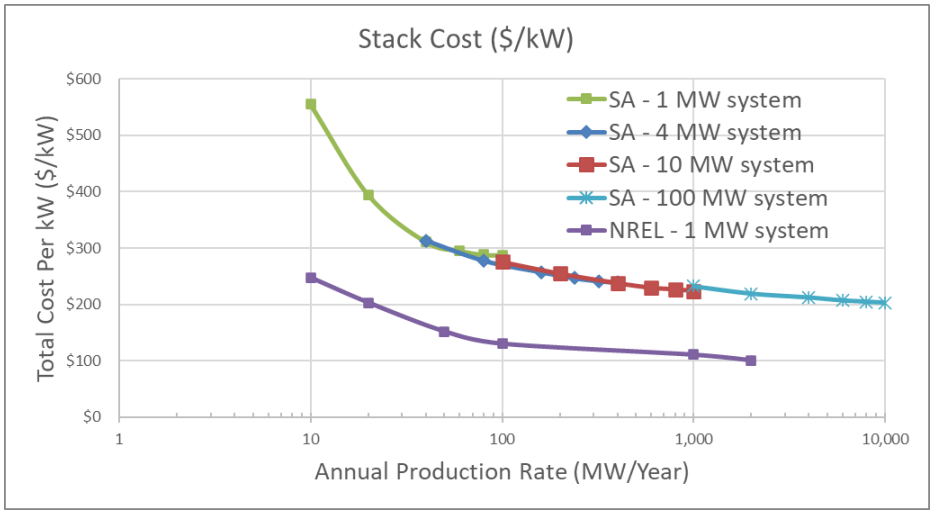


Figure 22 - Comparison of NREL 2019 stack cost estimate (purple) and SA's estimate (green) for a 1 MW system

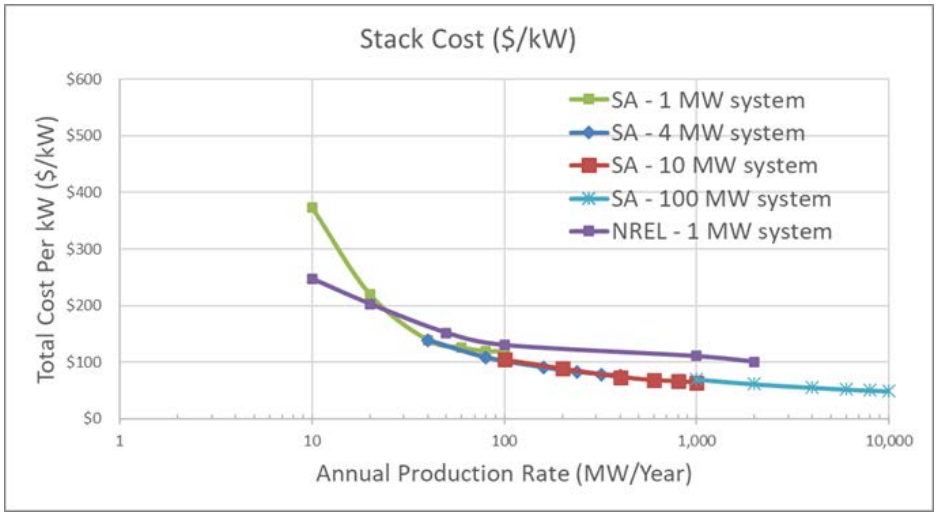


Figure 23 - Comparison of NREL 2019 stack cost estimate (purple) and SA's estimate with adjusted precious metal catalyst loading and material pricing (green) for a 1 MW system

Table 5 - Comparison of design assumptions for NREL's 2019 study, Ayers et al article, and SA's model

	NREL 2019 Study	Ayers/Pivovar Paper	2021 SA Model
Production Vol., Stack Size	Prod. Vol: 2 MW/yr to 50 GW/yr System Sizes: 200 kW, 1 MW/system Stack Sizes: 200 kW and 500 kW/stack		Prod. Vol: 10 to 10,000 MW/yr System Size: 1, 4, 10, 100 MW/system Stack Size: 1, 2, 2.5, 5 MW/stack
Membrane	Nafion 117 (183 µm thick) Assume with $1,500/kg ionomer cost and $500/$m^2$ total membrane lowest estimate	175-250 µm thick PFSA or sulfonated Radel	200 µm thick PFSA + ePTFE support – need to remodel membrane fabrication with Gore coating process
CCM – H2 Electrode (Cathode)	Spray coating of 0.4 mg/cm^2 Pt-Ir ~$700/tr.oz for Ir ~$1500/tr.oz for Pt	Pt/C (Vulcan or Ketjen black) 0.05 mg/cm^2	Slot die coating of 30% Pt/C onto membrane Loading: 0.24 mg Pt/cm^2, 0.8 mg catalyst/cm^2 Using $1500/tr.oz. for Pt
CCM – O2 Electrode (Anode)	Spray coating of 0.7 mg Pt/cm^2 Pt/C Using $1500/tr.oz for Pt	1-3 mg Ir/cm^2 loading for TiO_2-supported IrO_2 catalyst using NSTF or reactive spray deposition tech (RSDT)	Intend to model slot die coating of IrO_2 on TiO_2 onto membrane Loading: 2 mg Ir/cm^2 Currently cost modeled as: cost of d-PtCo, slot-die coated onto membrane, with Ir substitution for Pt (with Ir adjusted loading and price $5,000/tr.oz)
PTL –(H2)	Toray paper 090 – TGP-H-090 (280 µm thick)	Carbon paper (GDL)	Carbon paper (GDL)
PTL –(O2)	Sintered porous Ti ($35/kg Ti price) with 30% porosity, coated with gold (100 nm)	Porous sintered Ti (0.5-2 mm thick with 20-70% porosity) with PGM coating	Sintered porous Ti 1 mm thick with 50% porosity and 100 nm PVD Pt coating
Frame	PPS-40GF or PEEK thermoplastics	Don't mention	Includes cell frame on each side, subgasket, and gasket seals
BPP Base	197 mils (5 mm) SS ($5/piece)	Ti (15-50 mils from Kathy Ayers)	CP2 Ti 30 mils ($50-$85/kg – obtaining high vol. price quote), 48" wide coil cut to ~2 m lengths, etched flow fields, laser cut into individual BPP
BPP Coating	gold (100 nm thick)	Eliminate Pt coating, don't mention other	PVD 25 nm Au on cathode (H_2) and 25 nm Pt on anode (O_2)

5.2.4 SA PEM Stack Price

Assuming a 33% vendor gross margin (50% markup) for a 4 MW stack at 400 MW/year, the price equates to roughly \$1.35/cm^2. This aligns well with the PEM electrolysis stack price used in the published current H2A cases (\$1.30/cm^2 for a 3 MW system at 700 MW/year), graphed in Figure 24 below.

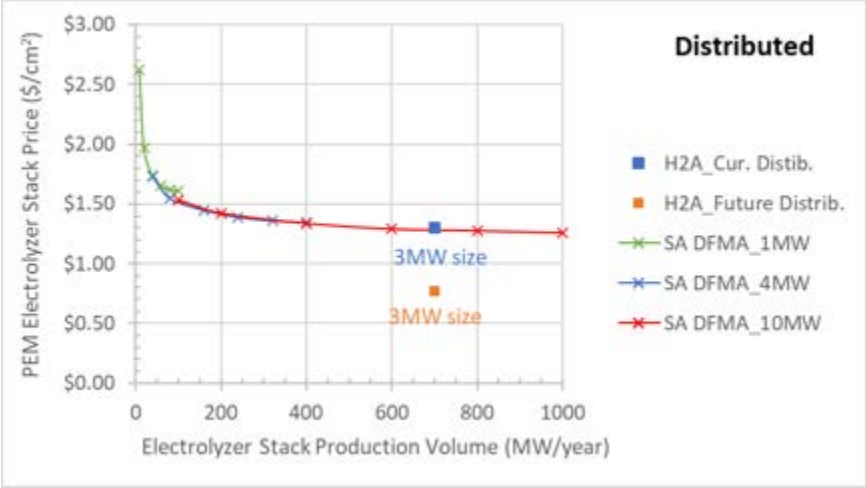


Figure 24 - Comparison of stack cost used in published H2A cases and SA DFMA® cost estimates

Further external review of assumptions and results of the PEM electrolyzer DFMA® model by NREL, Nel, and Giner/Plug Power is planned for the future as well as a sensitivity analysis to show cost reduction pathways based on reviewer comments. SA plans to evaluate the feasibility of a 10 MW stack and the issues that may dictate the extent of stack sizing.

5.3 H2A Case For LT-PEM Electrolysis

An H2A case study for LT-PEM Electrolysis was completed in 2019 and provided an update to the 2014 case study. H2A model v3.2018 was used for the updated analysis. Cost results from the model suggest only small $/kg$H_2$ changes compared to the previous (2014) case study. Costs are approximately $5.00/kg H_2 for Current Cases and $4.50/kg H_2 for Future Cases.

Key system costs and operating parameters for the PEM electrolyzer system are shown in Table 6.

Table 6 - Input parameters for H2A Production cases for PEM electrolysis (costs in 2016$).

Parameter	Current Distributed *1,500 kg/day*	Future Distributed *1,500 kg/day*	Current Central *50,000 kg/day*	Future Central *50,000 kg/day*
Technology Year	2019	2035	2019	2035
Start-up Year	2015	2040	2015	2040
Total Uninstalled Capital (2016$/kW)[15]	$599	$379	$460	$233
Stack Capital Cost (2016$/kW)	$342	$143	$342	$143
BoP CapEx (2016$/kW)	$257	$236	$118	$91
Mechanical BoP Cost (2016$/kW)[16]	$136	$140	$36	$23
Electrical BoP Cost (2016$/kW)	$121	$97	$82	$68
Total Electrical Usage (kWh/kg) [% LHV] (% HHV)	55.8 [59.7%] (70.6%)	51.4 [64.8%] (76.6%)	55.5 [60.1%] (71.0%)	51.3 [65.0%] (76.8%)
Stack Electrical Usage (kWh/kg) [% LHV] (% HHV)	50.4 [66.1%] (78.2%)	47.8 [69.8%] (82.4%)	50.4 [66.1%] (78.2%)	47.8 [69.8%] (82.4%)
BoP Electrical Usage (kWh/kg)	5.4	3.66	5.04	3.54
Stack Current Density (A/cm^2)	2.0	3.0	2.0	3.0
Cell Voltage (V)	1.9	1.8	1.9	1.8
Electrolyzer Power Consumption at Peak Production (MW)	3.56	3.53	119	118
Effective Electricity Price over Life of Plant[17] (2016¢/kWh)	7.27	7.87	7.35	7.91
Outlet Pressure from Electrolyzer (psi)	300	700	300	700
Installation Cost (% of uninstalled capital cost)	12%	10%	12%	10%
Stack Replacement Interval (years)	7	10	7	10
Stack Replacement Cost Percentage (% of installed capital cost)[18]	15%	15%	15%	15%
Plant Life (years)	20	20	40	40
Stack Degradation Rate (mV/khrs)	1.5	1	1.5	1
Cell Active Area (cm^2)	700	700	1,500	1,500
Capacity Factor (%)	97%	97%	97%	97%

[15] All capital costs in this table assume manufacturing at volumes such that economies of scale have been achieved.

[16] Mechanical BoP costs increase slightly between the *Projected Current* and *future* cases due to increased system operating pressure. Costs between the *Distributed* and *Central* cases decrease substantially due an assumption of increased reliability leading to decreased number of Mechanical BoP modules and hence increased unit size which benefits from economies of scale

[17] Effective electricity price over life of plant (20 years for *Distributed* cases and 40 years for *Central* cases)

[18] Stack Replacement Cost Percentage is estimated at 15% of the installed capital cost based on questionnaire responses. This cost is meant to capture the net expense of stack replacement, inclusive of old stack residual value and installation cost.

Table 7 summarizes the projected H_2 cost results for the PEM electrolyzer H2A case study.[19] Cost projections are also made with an electricity price of $0.03/kWh to illustrate the potential for low-cost H_2 if low-cost electricity is available.

Table 7 - PEM electrolyzer H2A case study projected H_2 cost results

Case Study	Low Value ($/kg H_2)	Baseline ($/kg H_2)	High Value ($/kg H_2)	H_2 cost at 3¢/kWh$_{electric}$
***Distributed**: Projected Current Case*[20]	$2.93	**$4.98**	$7.22	**$2.54**
Projected Future Case[21]	$2.16	**$4.48**	$6.07	**$1.92**
***Central:** Projected Current Case*[22]	$2.67	**$4.83**	$6.99	**$2.31**
Projected Future Case[23]	$2.16	**$4.48**	$6.14	**$1.86**

Three sensitivity analyses were conducted:

1) Single Variable Tornado Charts in which one parameter was varied, all others were held fixed at the baseline case values, and the new cost was recorded (Figure 25, showing only Distributed size systems).

2) Two Variable Contour Plots in which electricity cost and stack electrical usage were varied within the bounded ranges and the resulting hydrogen cost plotted in a contour graph (Figure 26, showing only Distributed size systems).

3) Monte Carlo Analysis in which all Table 6 parameters were stochastically and simultaneously varied over their full range to create a probability distribution function of potential hydrogen costs (Table 7).

[19] Peterson, D., Vickers, J., DeSantis, D., "Hydrogen Production Cost From PEM Electrolysis – 2019", DOE Hydrogen and Fuel Cells Program Record # 19009.
[20] For this case, the effective electricity price over the life of the plant is 7.27 ¢/kWh
[21] For this case, the effective electricity price over the life of the plant is 7.87 ¢/kWh
[22] For this case, the effective electricity price over the life of the plant is 7.35 ¢/kWh
[23] For this case, the effective electricity price over the life of the plant is 7.91 ¢/kWh

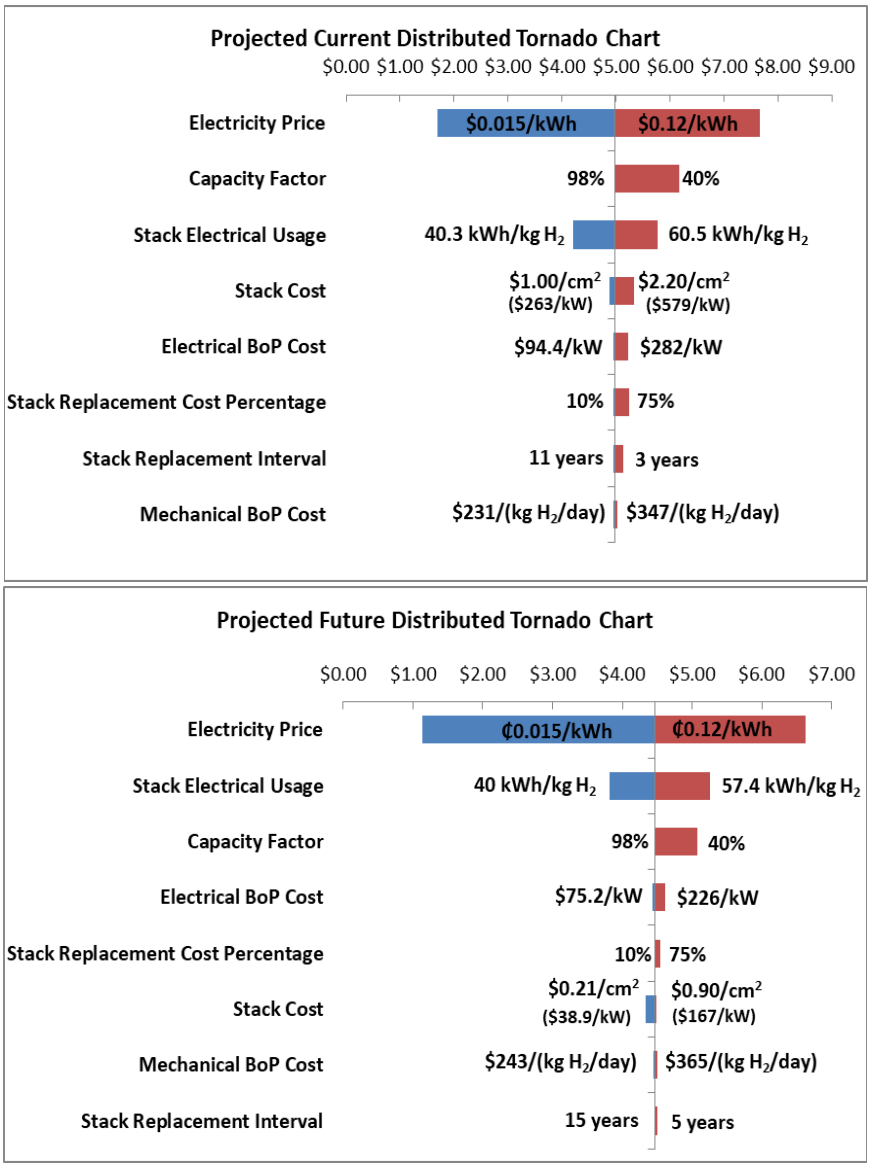


Figure 25 - Tornado chart showing parameter sensitivities for Projected Current and Projected Future Distributed PEM Electrolysis cases.

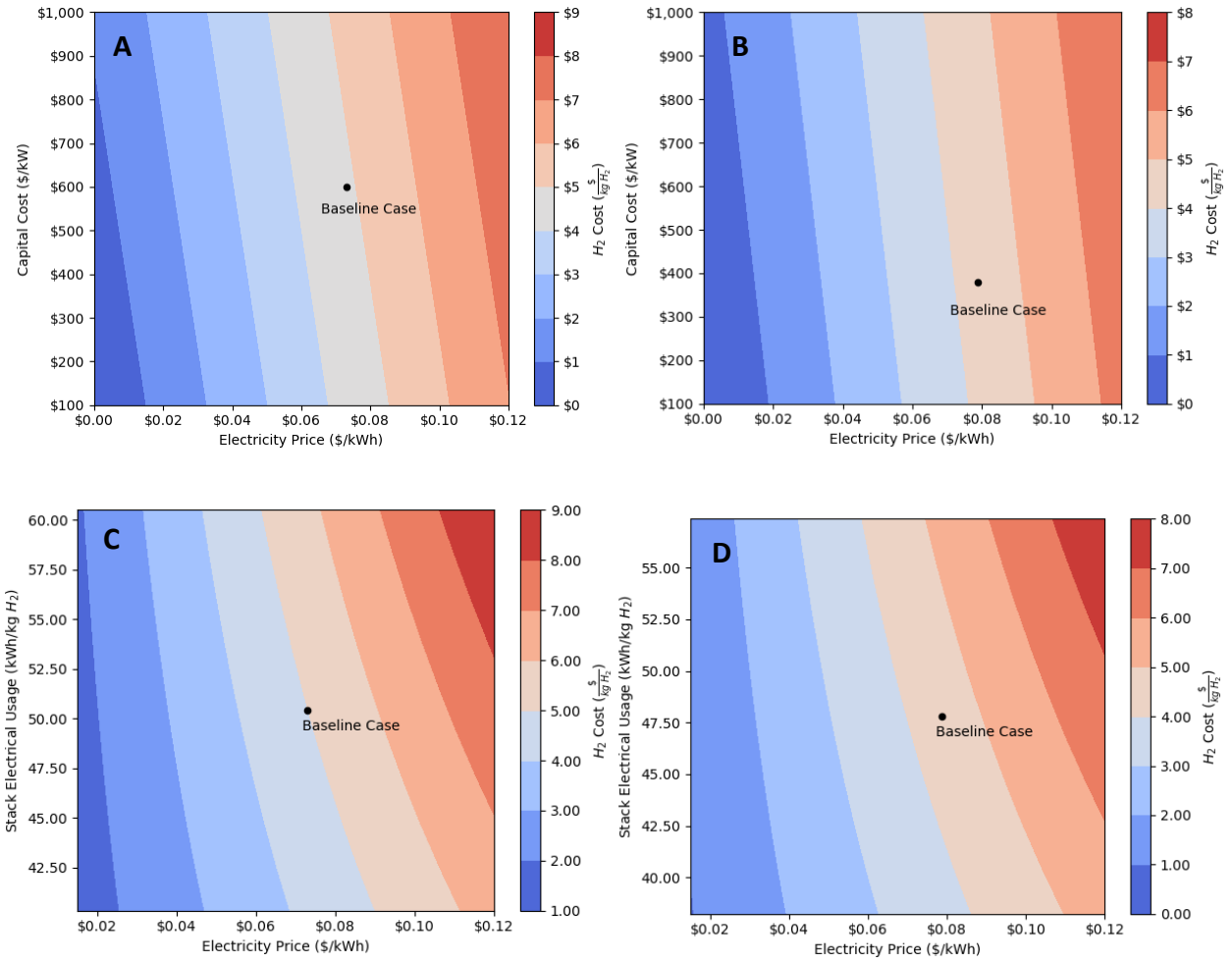


Figure 26 - Contour plots depicting cost variation for H_2 production with changes to electrolyzer system capital cost and electricity price and for: (A) Projected Current and (B) Projected Future Distributed PEM cases. Contour plots depicting cost variation for H_2 production with changes to stack electrical usage and electricity price for: (C) Projected Current and (D) Projected Future Distributed PEM cases.

6 Solid Oxide Electrolysis (SOE)

6.1 SOE Overview

SOE is a promising H_2 production technology due to its very high electrical efficiency (close to 100%) and its potential for low-cost stacks (due to low-cost materials). SOE is characterized by a high temperature solid ceramic electrolyte that conducts oxygen ions (O^{2-}) from the cathode to the anode to split water into oxygen and hydrogen. Figure 27 shows the working principle of a generic SOE as well as the redox reactions.

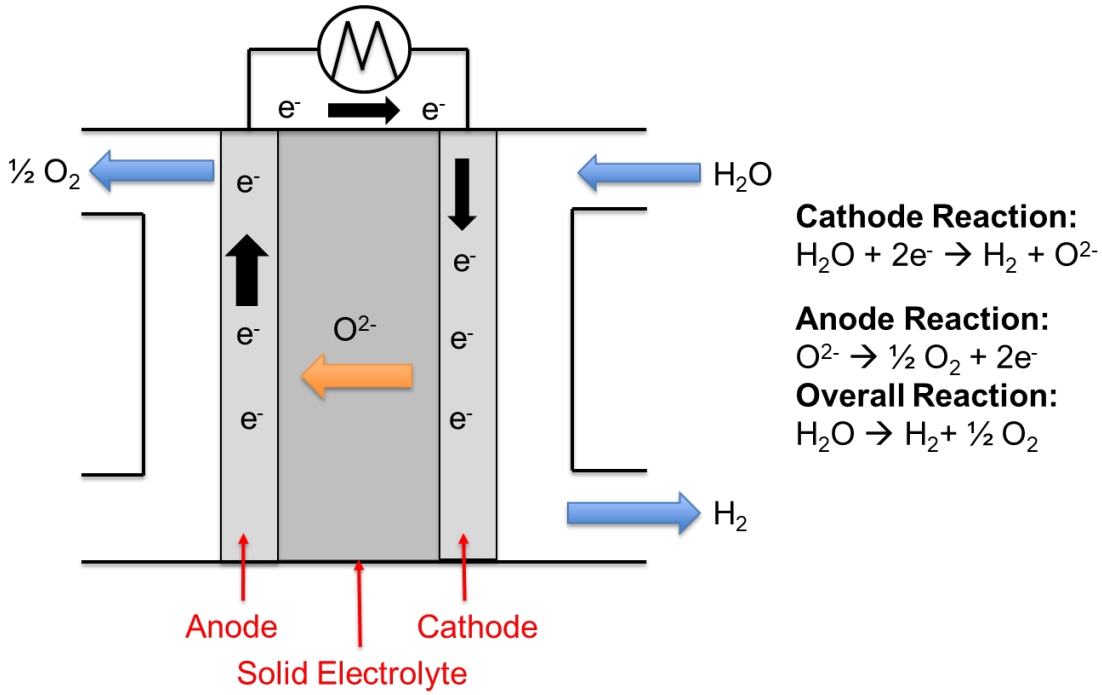


Figure 27 - Working principle of a generic SOE

For this study, we used our previous (2014) SOE case study as the basis for our analysis. The SOE system is broken down into 3 major parts: SOE Stack, Mechanical BoP, and Electrical BoP. A questionnaire was sent to seven SOE companies to solicit information on the current and future status of SOE design and operating parameters. Responses from three companies were collated to identify representative parameter values, commonalities within the industry, and identify trends between current and future systems. Only Central (50 tpd) cases, both Current (2019) and Future (2035), were analyzed. The SOE Current system design is shown in Figure 28. The system design is similar to the 2014 SOE case study and is derived from the Dominion Energy SOE in-depth analysis. Some adjustments to the heat exchanger arrangement and the steam supply have been incorporated into the system design. A complete system mass and energy balance was completed in ASPEN/Hysys®.

To provide a reasonable technological improvement for the Future Case, a proton-conducting SOE system was modeled. The proton-conducting SOE model pushes H^+ ions across the ceramic separator, instead of

O^{2-} ions. The key benefit of the proton-conducting SOE system is that it can operate at temperatures below 650°C, which means that 316 stainless steel could be used for many components that previously required high alloy steels. Additionally, because H^+ ions are transported across the electrolyte, there is less H_2O in the product H_2 steam, allowing the H_2O adsorption subsystem to be eliminated or at least minimized. Cell operating voltage is estimated to be the same 1.285 V (near thermo-neutral) as used in the higher temperature, O^{2-} ion transporting Current Case. The thermo-neutral voltage for a lower temperature system is not estimated to deviate appreciably from 1.285 V. Current density is projected to be 1.2 A/cm^2 (as opposed to 1.0 A/cm^2 for the Current Case) based on general technology improvement, rather than the H^+ ion SOE being intrinsically higher current density than the O^{2-} ion SOE. The SOE Future system design is shown in Figure 29.

A Bill of Materials (BoM) for the BoP was developed. BoM costs were created from price quotes, TEA, and 3rd party equipment cost models. Key system costs and operating parameters for the SOE system are shown in Table 8. SA created representative polarization curves for the SOE case, in a similar fashion as those created for the PEM Electrolysis case. The model assumes that as the area specific resistance increases through degradation, the system temperature can be increased to achieve target H_2 production.

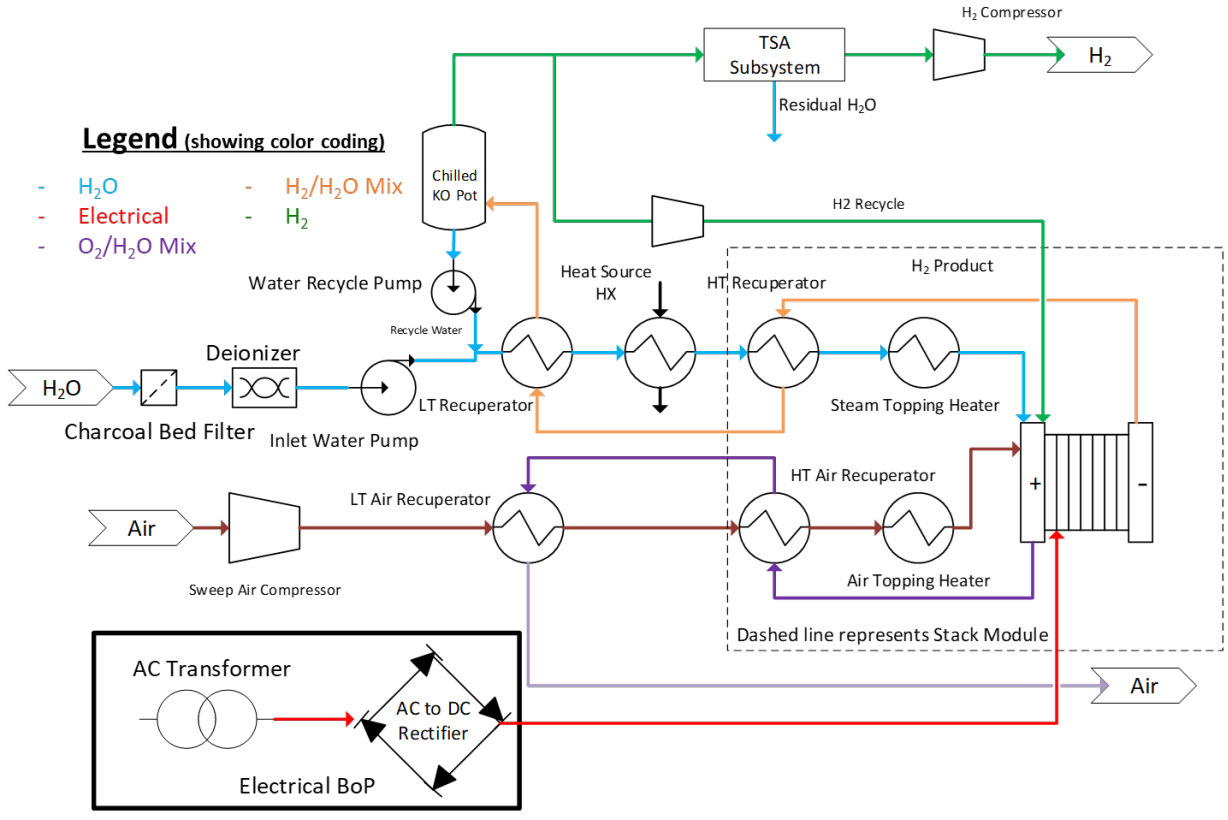


Figure 28 - Process Flow Diagram for the Current Case SOE system

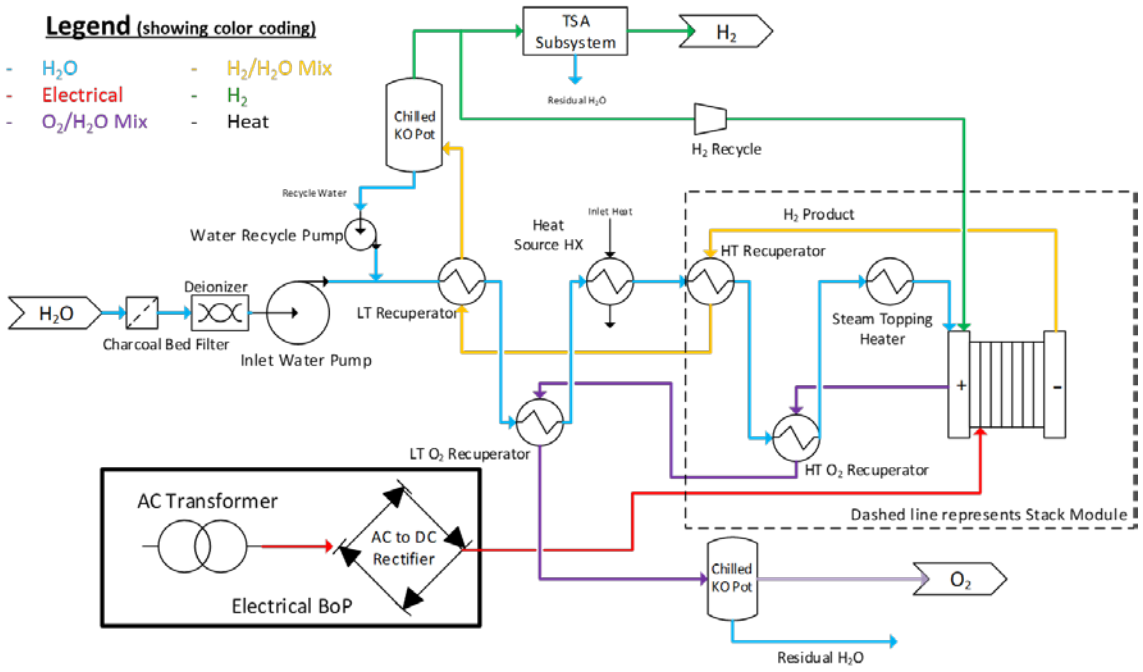


Figure 29 - Process Flow Diagram for the Future Case SOE system

Table 8 - Key parameters for SOE H2A analysis and design

	Units	Current Case	Future Case
Plant Size	kg H_2 day^{-1}	50,000	50,000
Current Density	A cm^{-2}	1.00	1.20
Voltage	V	1.285	1.285
Total Energy Usage	kWh/kg H_2	46.6	44.2
Stack Electrical Usage	kWh/kg H_2	34.0	34.0
Thermal Energy Usage	kWh/kg H_2	6.86	7.10
BoP Electrical Usage	kWh/kg H_2	5.76	3.06
Stack Cost	$ cm^{-2}	$0.20	$0.15
Mechanical BoP Cost	$ $kg^{-1}day^{-1}$	$348	$228
Electrical BoP Cost	$ kW^{-1}	$81	$65
System Cost	$ kW^{-1}	$481	$326
Stack Cost	$ kW^{-1}	$155	$100
Mechanical BoP Cost	$ kW^{-1}	$245	$160
Electrical BoP Cost	$ kW^{-1}	$81	$65

6.2 H2A Case For HT-SOE

An H2A case study for a Current and Future HT-SOE system was completed using our updated SOE results. The H2A model v3.2018 was used for the analysis. The projected price for H_2 produced from SOE is approximately \$4.16/kg H_2 for the Current Case. The projected price for H_2 produced from SOE is approximately \$3.98/kg H_2 for the Future Case. Electricity is the dominant cost contributor in SOE H_2 production. Figure 30 summarizes the projected H_2 cost results for the SOE H2A case study. Cost projections are also made with an electricity price of \$0.03/kWh to illustrate the potential for low-cost H_2 if low-cost electricity is available.

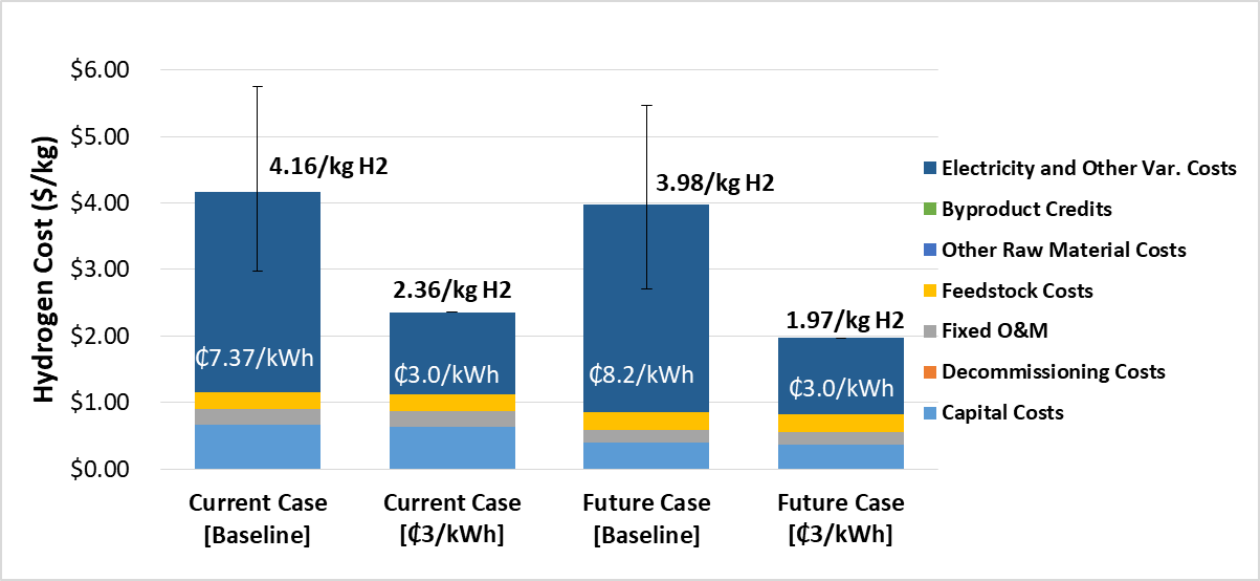


Figure 30 - SOE H2A case study projected H_2 cost results. Electricity is the dominant cost contribution. A case study is shown for both the Current and Future Cases with the electricity price set at \$0.03/kWh for the lifetime of the plant.

7 Anion Exchange Membrane (AEM) Electrolysis

7.1 AEM Electrolysis Overview

AEM electrolysis is another promising H_2 production technology and is characterized by a low-temperature solid polymer membrane electrolyte (conceptually similar to PEM electrolysis) that conducts hydroxide ions (OH^-) from the cathode to the anode to split water. Figure 31 shows the working principle of a generic AEM electrolyzer as well as the redox reactions. AEM systems can be designed in three different ways 1) liquid water is fed only to the anode, allowing a dry cathode thereby reducing the amount of water to remove at the H_2 outlet, 2) liquid water fed to only the cathode (with a dry anode), and 3) liquid water fed to both anode and cathode. Between these three types of designs, flow field designs and BOP components may be different. Water fed to the anode (with a dry cathode) is the method used by Enapter and used in SA's present analysis. Additionally, there is a choice of operating on a pure water feed (using the membrane as the sole electrolyte) or with a salt-water feed (nominally KOH/water) where the salt functions as an auxiliary electrolyte.

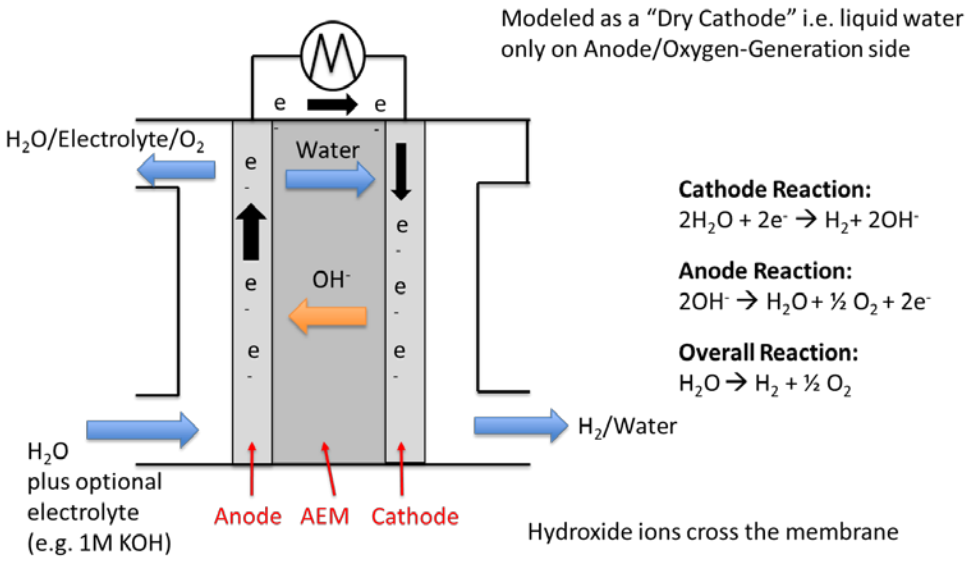


Figure 31 - Working principle of a generic AEM electrolyzer

Within this project, SA initially completed preliminary cost estimates for future and far-future AEM electrolysis systems. However, before finalizing the H2A cases, SA became aware of advancements in the technology when using alkaline electrolyte (1M KOH in water) and publicized larger manufacturing volume capabilities.[24] SA and the DOE decided to evaluate both Near Term and Future cases and re-evaluate the systems for both pure water and 1M KOH electrolytes.

Thus far, Enapter is the only industry manufacturer producing AEM electrolyzers. Since our request for Enapter's participation in this study was unanswered, we collaborated with a research group at the

[24] https://www.enapter.com/press-release/10000-green-hydrogen-generators-per-month

University of Delaware (and Versogen, the spin-off from that university group) who provided periodic technical support and feedback with information regarding AEM electrolysis operation, durability, and cost. Versogen is currently pursuing pure water AEM systems while Enapter is building systems that operate with KOH.

The commercialization of AEM electrolyzers by Enapter provided additional context for SA's understanding of planned future AEM development. Enapter commercialized an AEM electrolyzer system that produces 1 kg H_2/day with approximately 23 cells and 125 cm^2 active area. With a volume flow rate of 500 NL/h and an input power of 2.2 kW, the implied cell voltage is 1.82 V/cell. This AEM electrolyzer operation uses 1 M KOH with a non-platinum group metal (PGM) catalyst. Enapter claims a 30 kh lifetime and a 0.25%/kh degradation rate, which implies 5 mV/kh at constant current. The lifetime and degradation rate are currently dramatically superior to pure-water AEM electrolyzer systems. While the stack size is very small, Enapter has a modular approach and concepts for scale-up to MW system capacities.

A few research areas were further explored to better understand the differences between the Enapter AEM electrolyzer system with alkaline electrolyte and current pure water systems. Three research areas are described below:

1. The Enapter AEM electrolyzer system only feeds liquid H_2O to the O_2-evolution electrode. Consequently, the evolved H_2 is quite dry and requires a lesser level of drying than an H_2-side water-fed AEM or PEM electrolyzer system. The previous drier assumed in the H2A case may be oversized.
2. The Enapter electrolyzer system uses a 1 M KOH supporting electrolyte. However, most AEM research in the past has been focused on pure-water operation. The TEA-related advantages of pure-water operation over KOH have not been fully defined. While KOH addition in general boosts polarization performance, the 1 M KOH is caustic and raises safety and maintenance concerns. Alternate materials for wetted parts may be needed, extra safety components may add cost, and additional scrubber to remove KOH from the H_2 outlet stream will likely be needed.
3. Enapter claims a ~5 mV/kh degradation and ~450 mW/cm^2 power density. Advanced performance, pure-water, durable membrane systems are targeting >1 A/cm^2 current density at 1.8 V/cell and 50 mV/kh degradation. Thus, while pure-water AEM systems offer the potential for similar power density and lifetimes of PEM systems and require non-caustic liquids, AEM pure-water systems currently have inferior performance compared to alkaline electrolyte AEM systems.

7.2 AEM Stack and System Design

Figure 32 shows the AEM electrolyzer cell designs in literature: Figure 32A shows the Park et al cell design,[25] and Figure 32B shows the UD & Enapter cell design. Most cell design schematics follow similar construction utilizing a carbon gas diffusion layer (GDL) for the cathode and a titanium GDL for the anode. Figure 33 shows the modeled cell design for the present AEM electrolyzer study.

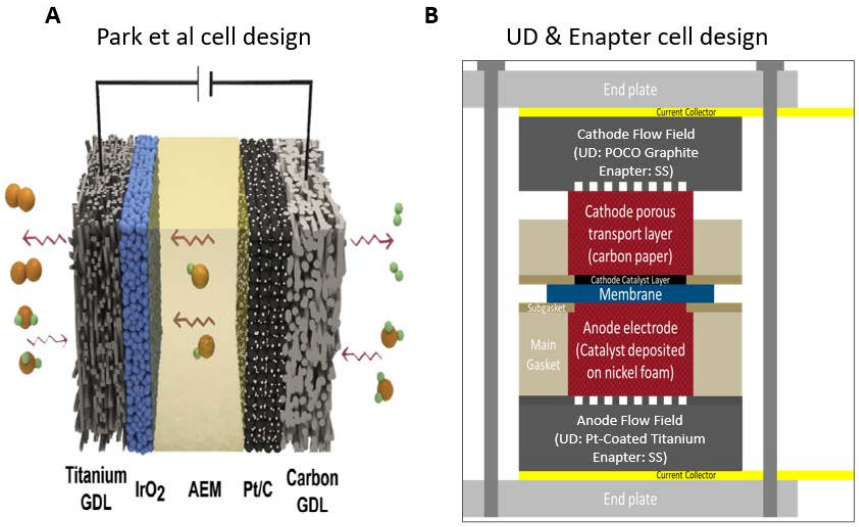


Figure 32 - AEM electrolyzer cell designs. (A) Park et al cell design. (B) UD & Enapter cell design.

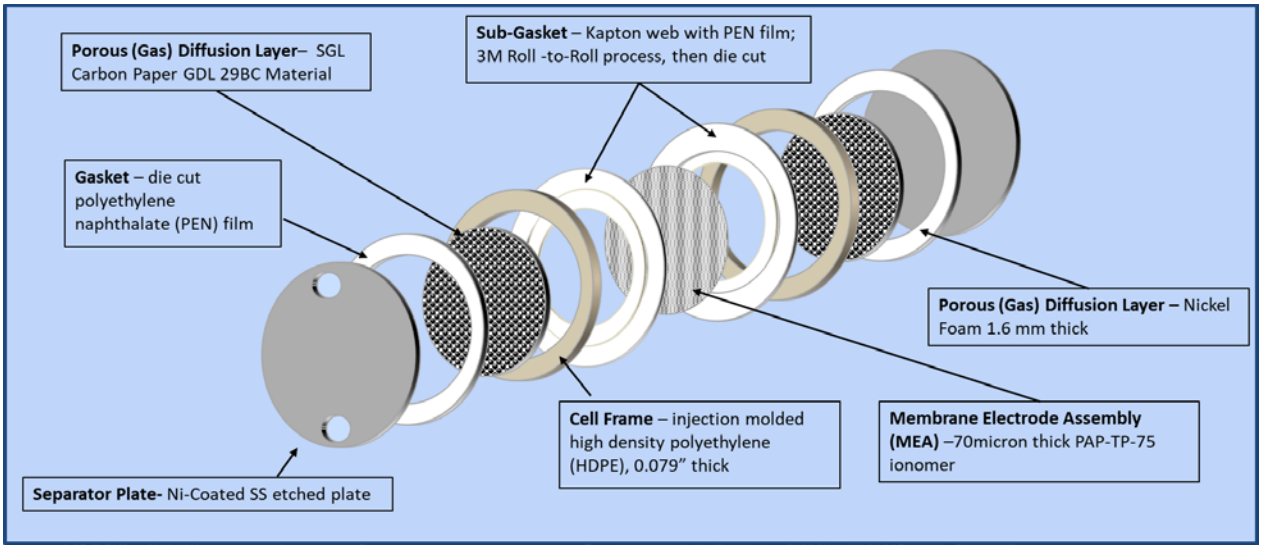


Figure 33 - Modeled cell design for the AEM electrolyzer

[25] Park, J. et al, "High-Performance anion-exchange membrane water electrolysis", Electrochimica Acta, 295 (2019) 99-106.

The AEM electrolyzer system consists of three main parts: AEM Stack, Mechanical BoP, and Electrical BoP. A DFMA® analysis methodology was used to predict the cost of the stack (see Section 7.3). The process flow diagram for the AEM electrolyzer system is shown in Figure 34.

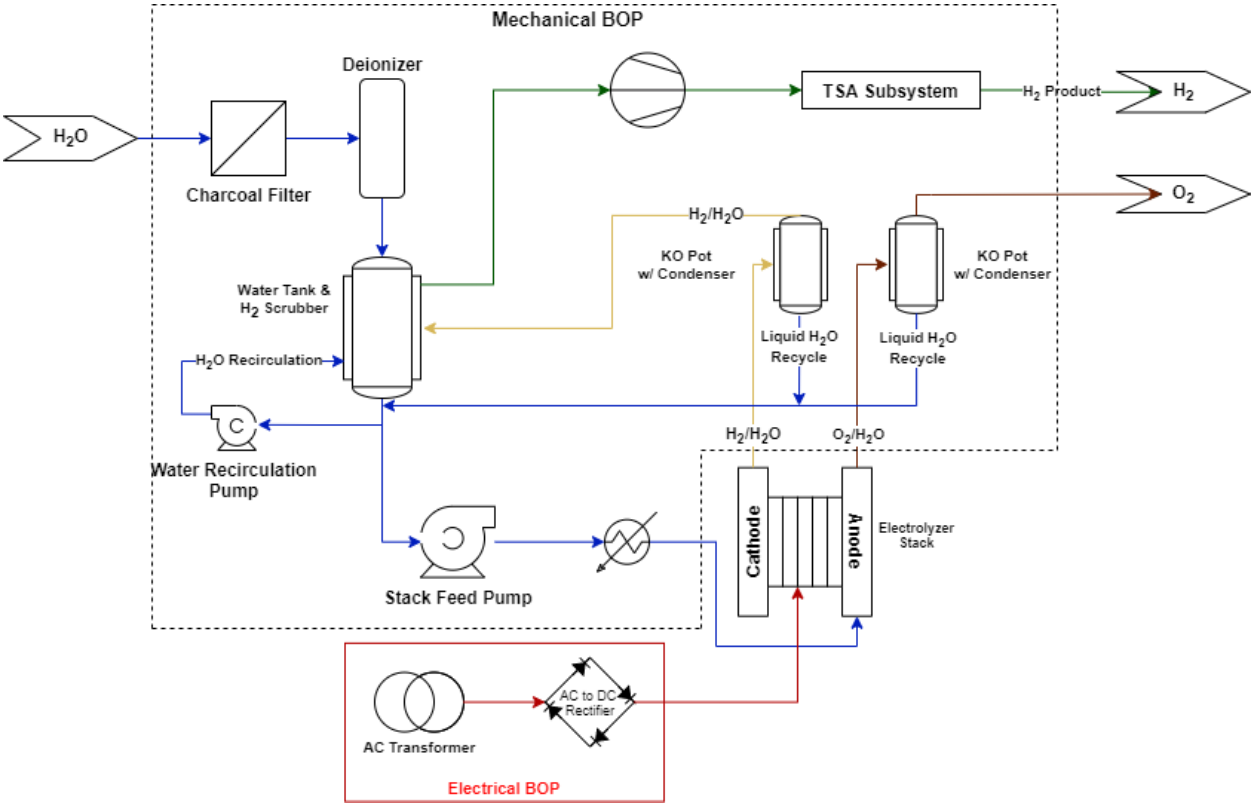


Figure 34 - Process Flow Diagram for the AEM electrolyzer systems

The BoP equipment lists for the AEM electrolyzer systems were developed based on a water-fed cathode design. Separate designs of a pure-water system and alkaline AEM system have not yet been generated. The AEM electrolyzer BoP costs are currently similar to the BoP costs of the PEM electrolyzer, however, there may be lower water feed purification requirements for AEM compared to PEM's strict requirements for deionized water purity and AEM does not have high rejected heat and therefore cooling systems can be lower cost than PEM's.

In 2021, SA's AEM work was presented to the HPTT and they requested that SA evaluate the BOP cost difference between pure-water and KOH AEM systems. Upon initial review, there does not seem to be a significant difference in BoP cost. Differences include a need for an additional KOH scrubber at the H_2 outlet of the stack for the KOH system, alternate materials for wetted parts in sensors, and extra safety components. The KOH scrubber is expected to be of minimal cost (~2% of the BoP cost) while further investigation is needed to quantity the cost impact of the wetted parts and safety components.

7.3 AEM Electrolysis Stack Cost DFMA® Analysis

SA used the PEM electrolysis DFMA stack cost model as a basis to evaluate the cost to manufacture an AEM electrolysis stack at multiple production rates, system sizes, and stack sizes. Various components were changed to align with an AEM system (i.e., non-precious metal catalyst on the anode and coating for the separator plate, SS instead of Ti for the separator plate base material). The cost was evaluated for a Near-Term 3 MW size system at six production volumes (10, 20, 40, 60, 80, 100 systems per year) and a Future 3 MW size system at six production volumes (20, 50, 80, 100, 150, 200 systems per year). Both systems were also evaluated for Pure-Water vs. KOH (assuming only differences in stack performance, all component material and designs remained the same). Stack sizes were 1 MW for Near-Term systems and 1.5 MW for Future systems.

Similar to the PEM electrolyzer stack design, the AEM stack was modeled with rectangular cells having active area sizing scaled linearly with the stack power sizing.[26] The resulting range in active area per cell for 1 MW up to 1.5 MW stack power is 740 and 1,020 cm^2 per cell, respectively. Figure 35 shows a cross-sectional view of a single electrolysis cell. Within Figure 35, annotations in blue describe each component material, thickness, and manufacturing process while annotations in red highlight possible changes to the model in the future.

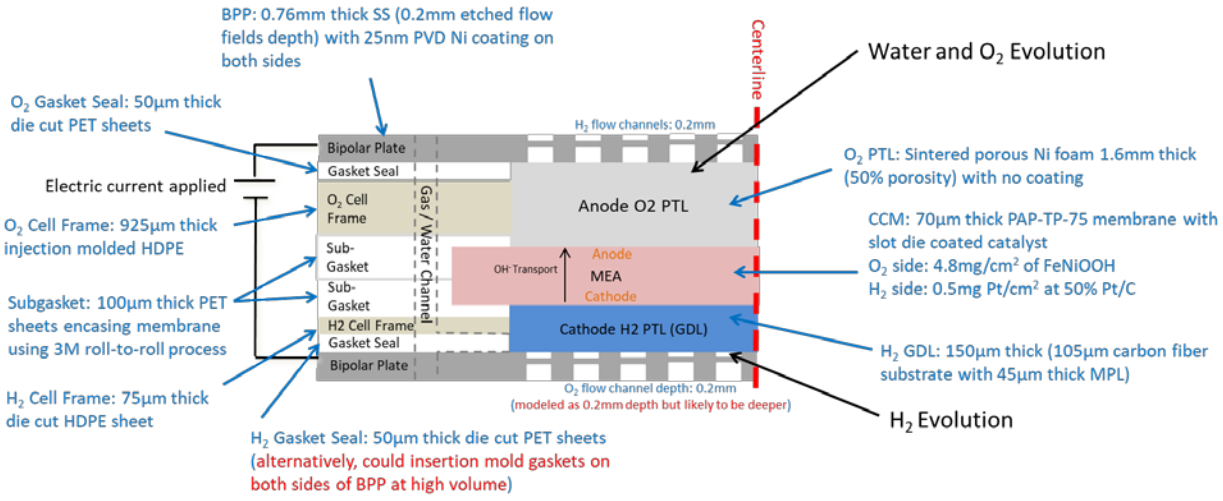
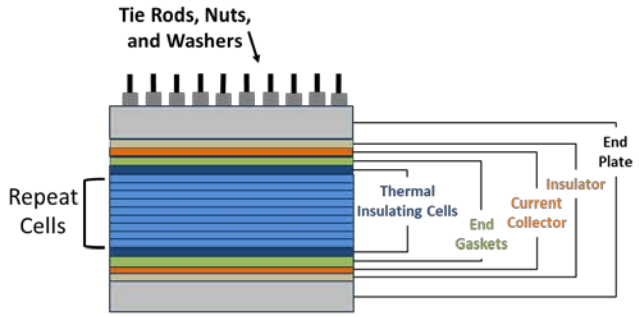


Figure 35 - Cross-sectional view of a single AEM electrolysis cell with descriptions of all modeled components

The stack design described in Figure 36 shows the repeat cells, thermal insulating cells (non-active cells), end gaskets, current collector, electrically insulating plate, and end plates. Not shown in the figure is a CVM to track possible cell reversals or poor-performing cells.

[26] This scaling came from Giner's publicized stack sizing[12] as previously shown in Figure 15.



Dummy Cells: two dummy cells per stack (one at each end) using four 150µm GDLs per dummy cell

End Gaskets: 45µm thick screen-printed polyolefin elastomer

Current Collector: 4mm thick copper stamped sheeting

Insulator: 1mm thick injection molded HDPE

End Plate: ~4cm thick stainless steel machined plates

Cell Voltage Monitor (CVM): Between 2 and 4 cells per sensor

Figure 36 - Full AEM stack design showing additional components as well as repeat active cells

The AEM stack design parameters are listed by component in Table 9. The parameters were determined from various sources (in addition to our own analysis and estimates), including the University of Delaware Questionnaire and the IONOMR white paper.[27]

[27] https://ionomr.com/wp-content/uploads/2021/02/FM-7024-A-Hydrogen-Production-Cost-by-AEM-White-Paper-copy.pdf

Table 9 - AEM stack design parameters

Parameter	Unit	AEM Near Term, Pure Water	AEM Near Term, 1 M KOH	AEM Future, Pure Water	AEM Future, 1 M KOH	AEM Notes
Anode Catalyst						
Catalyst	-	FeNiOOH	FeNiOOH	FeNiOOH	FeNiOOH	From Yushan Yun Questionnaire
Loading	$mg_{catalyst}/cm^2$	4.8	4.8	4.8	4.8	From Yushan Yun Questionnaire
Cost	$/kg	$20-165	$30-239	$20-165	$30-240	Material and Synthesis Cost Est.
Cathode Catalyst						
Catalyst	-	Pt/C	Pt/C	Pt/C	Pt/C	Current: Regularly used
Loading	$mg_{catalyst}/cm^2$	0.94 (0.47 mg Pt/cm^2)	0.94 (0.47 mg Pt/cm^2)	0.2 (0.1 mg Pt/cm^2)	0.2 (0.1 mg Pt/cm^2)	Yushan Yun Questionnaire and similar to loading in other cases
Cost	$/kg	$11k-$13k	$11k-$13k	$11k-$15k	$11k-$15k	Material and Synthesis Cost Est.
Membrane						
Material/Thickness	micron	70 um PAP-TP-75	70 um PAP-TP-75	70 um PAP-TP-75	70 um PAP-TP-75	Wang et al
Cost	$/$m^2$	$88-275	$108-336	$88-276	$108-338	Material, Synthesis, and Casting Cost Est.
Active Area per system	m^2	167	111	83	56	DFMA®
Stack Size	MW	1	1	1.5	1.5	DFMA®
Active area per cell	cm^2	740	740	1020	1020	DFMA®
Bipolar Plate						
BPP Material	-	SS 316	SS 316	SS 316	SS 316	
BPP thickness	cm (mils)	0.0762 (30)	0.0762 (30)	0.0762 (30)	0.0762 (30)	
BPP Cost (incl. coating)	$/plate	$24-64	$25-95	$38-108	$43-156	DFMA® cost calculation
Coating Material	-	Ni	Ni	Ni	Ni	
Coating Thickness	microns	0.025	0.025	0.025	0.025	
Other Stack Components						
Cathode GDL		SGL Carbon GDL 29BC	SGL Carbon GDL 29BC	SGL Carbon GDL 29BC	SGL Carbon GDL 29BC	Same GDL as PEM FC
Anode Electrode		Ni Foam (catalyst coated)	Ni Foam (catalyst coated)	Ni Foam (catalyst coated)	Ni Foam (catalyst coated)	
Current Collector		Stamped Cu Plate	Stamped Cu Plate	Stamped Cu Plate	Stamped Cu Plate	
Endplate		Machined SS	Machined SS	Machined SS	Machined SS	
Compression System		Tie Rods	Tie Rods	Tie Rods	Tie Rods	
Operating Conditions						
Operating Voltage	V	1.84	1.74	1.9	1.8	NT: Based on conv. w/ UD (2021) Future: PEM current and future
Current Density	A/cm^2	1	1.5	2	3	NT: Based on conv. w/ UD (2021) Future: PEM current and future
Stack Pressure	Bar	30	30	30	30	Ionomr white paper
System Performance						
Degradation Rate	mV/khrs	50	13	1.5	1	NT: Based on conv. w/ UD (2021) Future: PEM current and future
Stack Lifetime	years	1.1	3.4	7	10	Based on degradation rates
Stack Cost	$/$cm^2$	$0.2-0.37	$0.21-0.44	$0.18-0.34	$0.34	DFMA® Stack Analysis (no markup)
Stack Cost	$/kW	$110-201	$82-170	$47-90	$32-73	DFMA® Stack Analysis (no markup)

Cost results ($/kW$e_{stack}$) at all production volumes are depicted in Figure 37 for each system size in systems per year and in Figure 38 for each system size on a MW per year basis. The cost estimate for manufacturing the stacks is between about $30/kW and $200/kW. For both the Near-Term and Future systems, KOH reduces the cost across all production rates over Pure-Water. Additionally, Future systems for both KOH and Pure-Water show the potential for a very low-cost option for H_2 production.

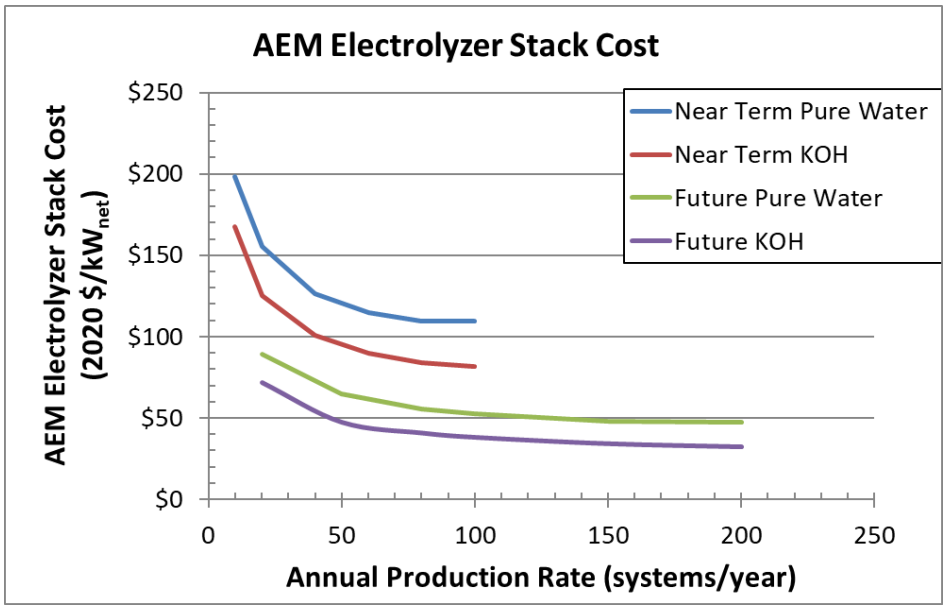


Figure 37 - AEM electrolyzer stack cost over production rate in systems per year for Near-Term and Future systems with Pure-Water vs. KOH

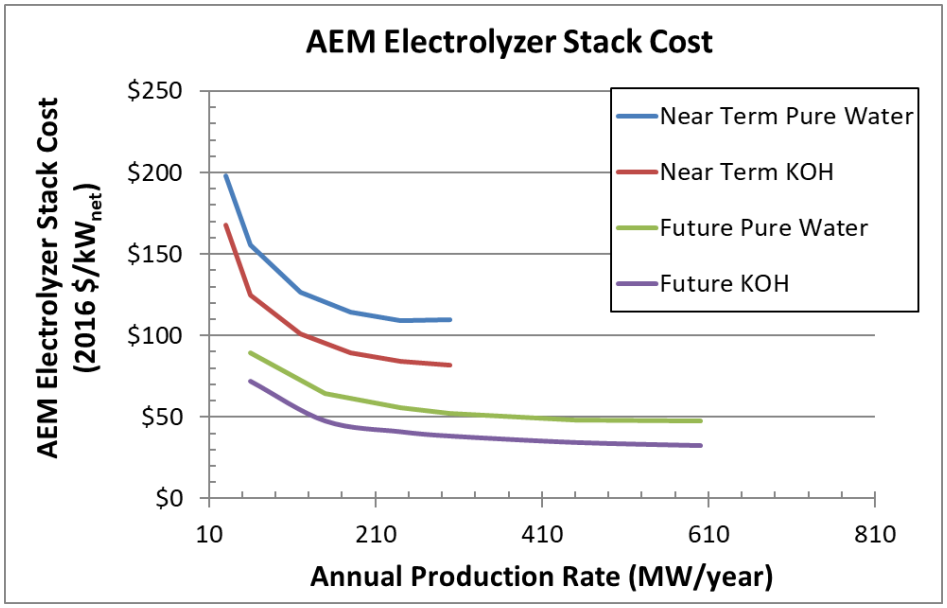


Figure 38 - AEM electrolyzer stack cost over production rate in MW per year for Near-Term and Future systems with Pure-Water vs. KOH

Figure 39 and Figure 40 show the stack cost breakdown at all production volumes for a 3 MW Near-Term system with Pure-Water and a 3 MW Future system with KOH, respectively. The stack cost for both systems at all volumes is dominated by the BPP and PTL costs. In addition, the catalyst cost also dominates for the 3 MW Near-Term system with Pure-Water, but not for the 3 MW Future system with KOH.

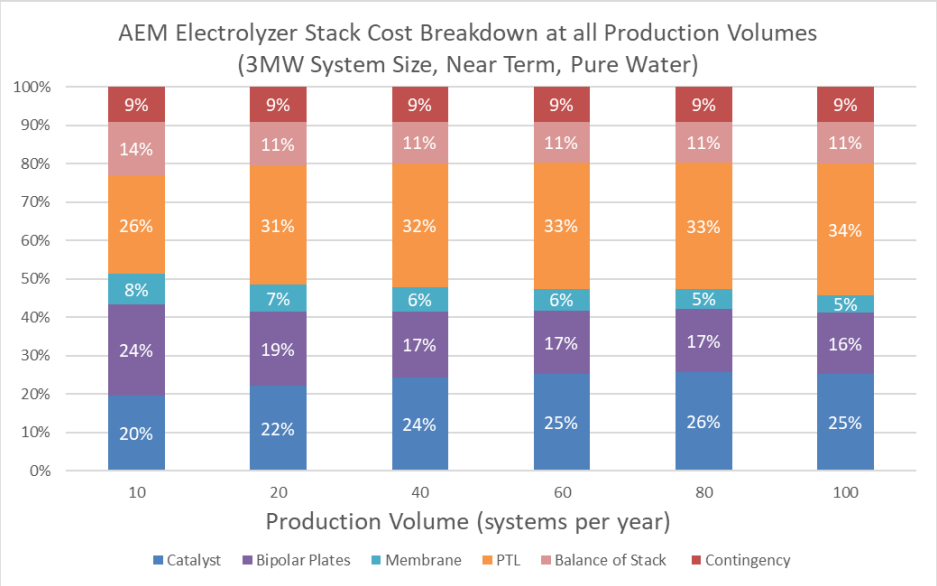


Figure 39 - AEM stack cost breakdown by production volumes for 3 MW Near-Term system with Pure-Water

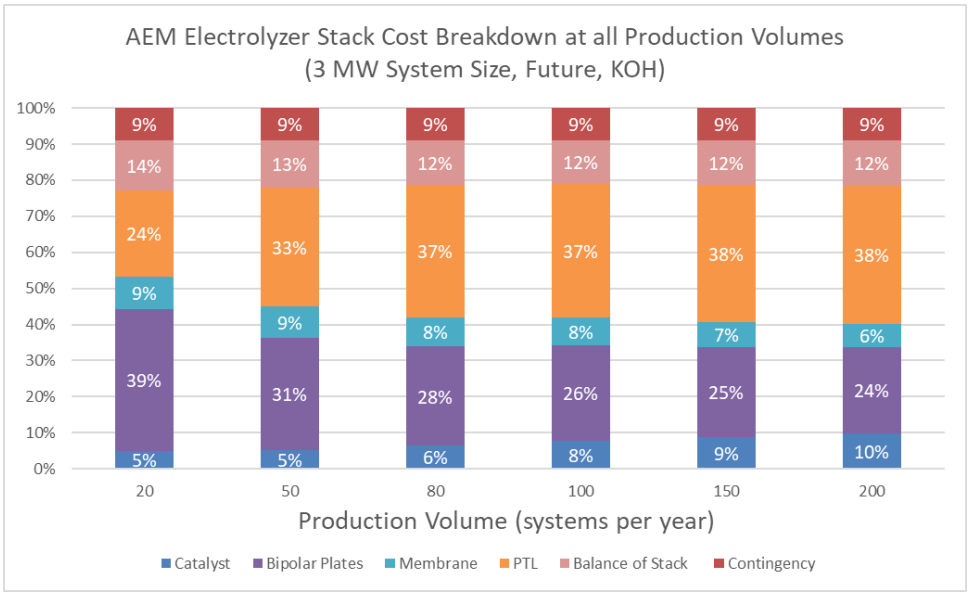


Figure 40 - AEM stack cost breakdown by production volumes for 3 MW Future system with KOH

A comparison of the differences between SA's PEM design, IONOMR white paper, and SA's 2019 and 2021 AEM designs is listed by component in Table 10.

Table 10 - Comparison of design assumptions for SA's 2021 PEM study, IONOMR white paper, and SA's 2019 and 2021 AEM studies

	SA 2021 PEM Study	IONOMR White Paper	SA 2019 AEM Study (F = future, FF = far F)	SA 2021 AEM Study (NT = near term, F = future)
Production Vol., Stack Size	Prod. Vol: 10 - 10,000 MW/yr System Size: 1, 4, 10, 100 MW/system Stack Size: 1, 2, 2.5, 5 MW/stack	Prod. Vol: 100 - 5,000 MW/yr System Size: 1 MW - 5 MW Stack Size: 200 kW - 1 MW	700 MW/yr (F and FF)	Prod. Vol: 300 MW/yr (NT), 600 MW/yr (F) System Size: 3 MW (NT and F) Stack Size: 1 MW (NT), 1.5 MW (F)
Electrolyte	Water	1 M KOH	Water	1 M KOH electrolyte (NT/F) Pure water electrolyte (NT/F)
Membrane	200 µm thick PFSA + ePTFE support 0.1 mg Pt/cm^2 GRC catalyst	Aemion+^{TM} \$220/m^2 (with bounds between \$175/m^2 and 250/m^2)	70 microns PAP-TP-75, \$864/kg @ 700 MW/yr – F and FF	70 microns PAP-TP-75 (NT and F)
CCM – H2 Electrode (Cathode)	Slot die coating of 30% Pt/C onto membrane. Loading: 1 mg Pt/cm^2, 3.33 mg Catalyst/cm^2 Using \$1500/tr.oz. for Pt	2 mg/cm^2 NiCrMo @\$100/m2 with 1 mg/cm2 ionomer loading	0.5 mg/cm^2 Pt/C @ \$32,000/kg (F) 0.94 mg/cm^2 $CuCoO_x$ @\$300/kg @700mW/yr (FF)	0.47 mg Pt/cm^2 Pt/C (NT) (0.94 mg catalyst/cm2) 0.1 mg Pt/cm^2 Pt/C (F) (0.2 mg catalyst/cm2)
CCM – O2 Electrode (Anode)	Slot die coating of IrO_2 on TiO_2 onto membrane. Loading: 2 mg Ir/cm^2 Currently cost modeled as: cost of d-PtCo, slot die coated onto membrane, with Ir substitution for Pt (with Ir adjusted loading and price \$5,000/tr.oz)	2 mg/cm^2 NiMo @ \$100/m2 with 0.273 mg/cm2 ionomer loading	4.8 mg/cm^2 FeNiOOH @ \$27.60kg @ 700MW/yr – same for F and FF	4.8 mg/cm^2 FeNiOOH – same for NT and F
PTL –(H2)	Carbon paper (GDL)	Carbon paper (GDL)		Carbon paper (GDL)
PTL –(O2)	Sintered porous Ti 1 mm thick with 50% porosity and 100 nm PVD Pt coating	Ni Foam		Sintered porous Ni Foam 1.6 mm thick with 50% porosity and no coating
Frame	Includes cell frame on each side, subgasket, and gasket seals	PPS-40GF		Includes cell frame on each side, subgasket, and gasket seals
BPP Base	CP2 Ti 30 mils (\$50-\$85/kg – obtaining high vol. price quote), 48" wide coil cut to ~2 m lengths, etched flow fields, laser cut into individual BPP	SS		30 mils SS 316, using \$13.60/kg placeholder (quote for thinner material), 48" wide coil cut to ~2 m lengths, etched flow fields, laser cut into individual BPP
BPP Coating	PVD 25 nm Au on cathode (H_2) and 25 nm Pt on anode (O_2)	Ni		25 nm Ni on cathode and anode size of BPP

There was also a comparison to IONOMR's cost study outlined in a white paper.[28] IONOMR assumed an anode and catalyst cost of \$100/m^2 for NiMo or NiCrMo while SA anode catalyst cost is between \$10 and \$260/m^2 for FeNiOOH and cathode catalyst cost between \$95 and \$400/m^2 for Pt/C. Figure 41 shows SA estimates to be lower than IONOMR estimates at both low (100 MW/yr) and high volume (5 GW/yr). It is difficult to discern the source of the difference as IONOMR cost estimates are not broken down by component and material sets for components other than catalyst are similar.

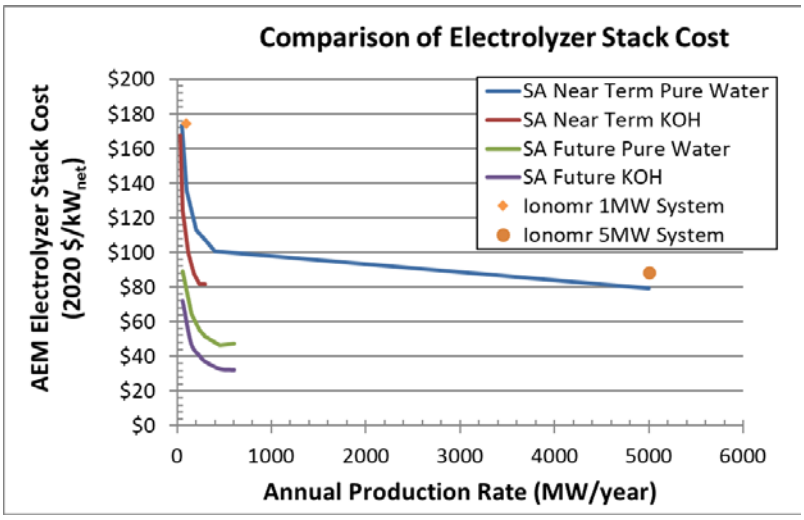


Figure 41 - Comparison of SA estimate (3 MW systems) and IONOMR estimate for AEM electrolyzer stack cost

[28] https://ionomr.com/wp-content/uploads/2021/02/FM-7024-A-Hydrogen-Production-Cost-by-AEM-White-Paper-copy.pdf

7.4 Future AEM Studies

Several studies are planned for the future including complete H2A cases. Before finalizing the cases, however, SA plans to have assumptions for the new cases vetted by external viewers. Additionally, SA plans to compare the production cost of H_2 when the stack cost and operating power density are changed. This will show whether an inexpensive stack operating at low power density may be more beneficial to cost compared to an expensive stack operating at high power density. Based on previous studies, inexpensive stacks may offer a reduced H_2 production cost, not primarily by reducing stack cost, but rather by allowing a reduced operating voltage for a physically larger stack thus reducing electricity consumption.

AEM electrolyzer polarization curves will be postulated for all cases based on a simple first-principles electrolysis model previously developed by SA. Those polarization curves will then combine with the AEM electrolyzer stack cost model and an AEM electrolyzer H2A model to estimate the resulting cost of H_2 at various current densities.

Finally, a detailed list of BOP components for both pure-water and alkaline systems will be generated to highlight cost differences and if one technology may have superior cost benefits than the other in the future.

8 Photoelectrochemical (PEC) Electrolysis

8.1 PEC Electrolysis Overview

Photoelectrochemical (PEC) production of hydrogen is a promising method for generating hydrogen using 100% renewable energy at a low cost. PEC devices use photons to split water thereby combining the functionality of photovoltaics and electrolyzers. This holds the potential to increase efficiency and decrease cost although PEC is at a substantially reduced maturity level compared to traditional PV or electrolysis systems.

In 2009, SA conducted a techno-economic evaluation of conceptual PEC hydrogen production systems.[29][30] This was followed up with a journal article documenting those cost assumptions coupled with a summary of the demonstrated photoelectrochemical capabilities as of 2013.[31] There have been several follow-up studies that compared PEC technologies to alternative hydrogen production pathways. Shaner et al (2016) suggested that PEC Type 3 and 4 had a lower levelized cost of hydrogen (LCH) than a base case photovoltaic-electrolyzer (PV-E) system but could not outcompete a grid supplemented photovoltaic electrolysis (GSPV-E) or solely grid-supplied electricity from fossil fuels coupled with electrolysis.[32] Ardo et al (2018) postulated that a PEC system would need to compete against an equivalent PV-electrolyzer system on both cost and efficiency, which may be difficult to achieve.[33]

After examining four different potential PEC technologies and holding discussions with the DOE, Type II and Type IV technologies were selected for further analysis (see Figure 42 and Figure 43). SA examined and refreshed the previously-developed cost models associated with PEC Type II and PEC Type IV. PEC Type II is a dual water-bed colloidal suspension of PV nanoparticles, with one bed carrying out an oxygen evolution reaction and the other bed carrying out a hydrogen evolution reaction. Hydrogen ions travel between beds via an ion bridge and complete the net reaction. PEC Type IV systems use reflectors to concentrate solar flux at greater than 10:1 intensity ratio onto multi-junction PEC element receivers immersed in a water reservoir and pressurized to approximately 300 psi.

[29] In 2011, Strategic Analysis Inc. acquired Directed Technologies Inc. (DTI). Thus, the 2009 PEC analysis was conducted by DTI.

[30] James, Brian D., George N. Baum, Julie Perez, and Kevin N. Baum. "Technoeconomic Analysis of Photoelectrochemical (PEC) Hydrogen Production," December 1, 2009. https://doi.org/10.2172/1218403.

[31] Pinaud, Blaise A., Jesse D. Benck, Linsey C. Seitz, Arnold J. Forman, Zhebo Chen, Todd G. Deutsch, Brian D. James, et al. "Technical and Economic Feasibility of Centralized Facilities for Solar Hydrogen Production via Photocatalysis and Photoelectrochemistry." Energy & Environmental Science 6, no. 7 (2013): 1983. https://doi.org/10.1039/c3ee40831k.

[32] Shaner, Matthew R., Harry A. Atwater, Nathan S. Lewis, and Eric W. McFarland. "A Comparative Technoeconomic Analysis of Renewable Hydrogen Production Using Solar Energy." Energy & Environmental Science 9, no. 7 (2016): 2354–71. https://doi.org/10.1039/C5EE02573G.

[33] Ardo, Shane, David Fernandez Rivas, Miguel A. Modestino, Verena Schulze Greiving, Fatwa F. Abdi, Esther Alarcon Llado, Vincent Artero, et al. "Pathways to Electrochemical Solar-Hydrogen Technologies." Energy & Environmental Science 11, no. 10 (October 10, 2018): 2768–83. https://doi.org/10.1039/C7EE03639F.

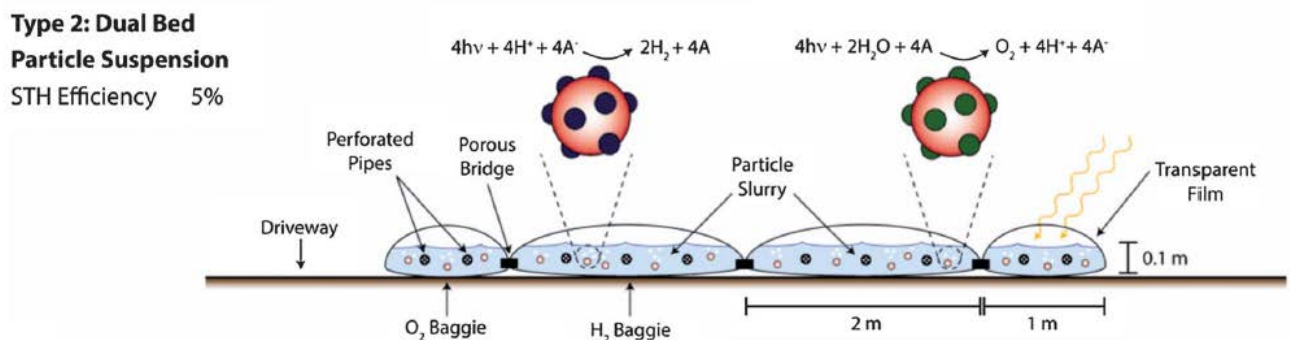


Figure 42 - Type II PEC parallel baggie graphical representation. Listed efficiency from past analysis may not be the value used in present H2A analysis.

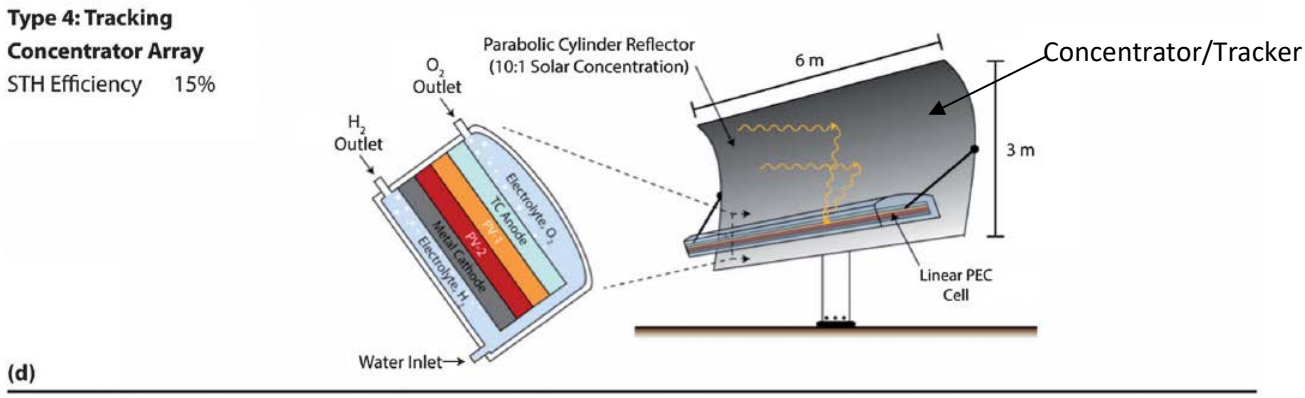


Figure 43 - Type IV PEC graphical representation. Listed efficiency from past analysis may not be the value used in present H2A analysis. The dimensions for the concentrator/tracker correspond to the photon capture area before it is curved for installation.

To update the technical and cost assumptions for PEC Type II and IV, SA pursued the following activities:

- Literature review of PEC designs and operating parameters
- Teleconference communication with NREL PEC experts
- Teleconference communication with JCP to review system designs
- Technical reviews with academic research experts

Assumptions and results for the updated PEC Type II and IV are described below.

8.2 General Solar to Hydrogen Assumptions

Type II PEC systems separate the H_2 evolution reaction (HER) and the O_2 evolution reaction (OER) into separate bags by locating only an HER or OER catalyst in each bag (Figure 42). PEC Type IV systems evolve H_2 and O_2 on opposite faces of the electrode panel and thus naturally separate H_2 and O_2 by the structure of the PEC cells and casement (Figure 43). The O_2 can be vented to the atmosphere as waste or to storage (for pressurization and resale). Regardless of the PEC type used, each PEC type will have a similar top-level system process design appropriate for a large Central plant 10-100 tonnes of H_2 per day composed of independent production modules operating in parallel. Each module is designed for a production rate of 1,000 kg H_2/day. Namely, the module will be a solar collection field with the PEC technology of choice. Pumps will feed an H_2O/electrolyte solution to the field. The O_2 can be vented to the atmosphere as waste or to storage (for pressurization and resale). The hydrogen is pressured to 20 bar (300 psi) consistent with standard H2A assumption and a typical pipeline pressure. The product hydrogen is saturated with water and purified to industrial grade quality (two 9's purity) using one condenser and two intercoolers. Further drying is not considered within this study but could be achieved using a refrigeration cycle or a Temperature Swing Adsorption (TSA) system. The standard H2A purity specification is five 9's purity.

The method for calculating the solar insolation rate and subsequent hydrogen production is common to all PEC methods. For this study, systems were designed for solar conditions in Dagget, CA. Type II systems have a lower solar insolation rate (horizontal flat plate, direct and diffuse light) than the modeled Type IV systems (1-axis solar tracking, concentrated, direct light only). Daily insolation data for PEC Type II are shown in Figure 44 and Figure 45.

The hydrogen production rate per photon capture area can be calculated using the average annual insolation rate and the solar to hydrogen efficiency. Using Equation (**2**), the mass flow rate of H_2 generated can be found on a per area basis.

$$\frac{\dot{m}}{A}\left(\frac{g\ H_2}{sec \cdot m^2}\right) = \frac{\eta_{STH} \cdot MW_{H_2} \cdot I_r}{n_{H_2} \cdot F \cdot E_{Eq}} \tag{2}$$

E_{Eq} = 1.229 V, represents the Equilibrium potential at standard conditions.
F = 96,485 s A $mol_{electron}^{-1}$, Faraday's Constant.
η_{STH} = Solar To Hydrogen (STH) efficiency (values listed in tables below).
n_{H2} = 2, electrons per mol H_2 atom.
I_r = Irradiance (W m^{-2}).
MW_{H2} = 2.016 g/mol, Molecular Weight of H_2.
$\dot{m}$ = mass flow rate (kg/day).
I_s represents the insolation across a given photon capture area (HER bed size or concentrator platform area).

The PEC II and PEC IV average annual irradiance rates (averaged over 24h/day, 365 days/yr), calculated from the average insolation rates shown in Figure 44 and Figure 45, are 240 W m^{-2} and 311 W m^{-2}, respectively.[34]

With the area specific mass flow rate and a module mass flow rate (average of 1,000 kg H_2/day)[35] the total Photon Capture Area (PCA) can be calculated (Equation (**3**)).

$$PCA(m^2) = \left(\frac{\dot{m}}{A}\right)^{-1} \frac{1{,}000\ kg\ H_2}{day} \frac{1\ day}{24\ hrs} \frac{1\ hr}{3{,}600\ secs} \tag{3}$$

With the total PCA determined and the module mass flow rate, other key parameters for the module can be identified.

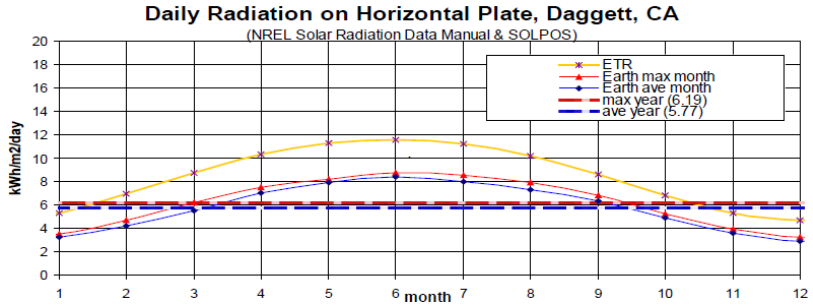


Figure 44 - Insolation data taken for Daggett California with a horizontal bed such as at Type II PEC. The annual average is 5.77 kWh day^{-1} m^{-2} for this location.

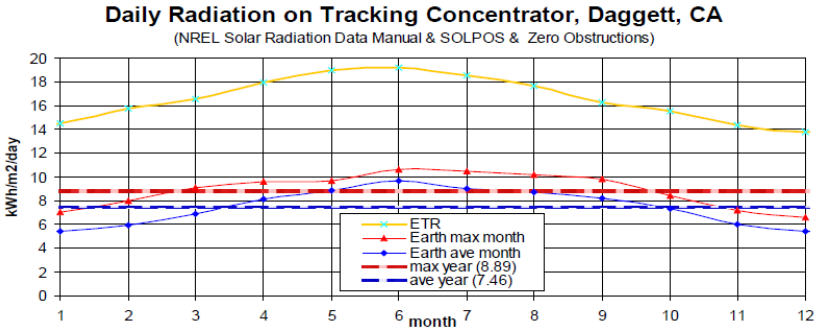


Figure 45 - Insolation data for tracking concentrator panels used in a type IV PEC system. The location is taken as Daggett, California. The average annual irradiance is 7.46 kWh day^{-1} m^{-2}.

[34] To calculate the insolation rate from irradiance, divide the irradiance by 24 hours/day and multiply the result by 1000 W/kW.

[35] This value will be reduced below the average for days with limited sunlight or increased above the average for days with extended periods of sunlight.

8.3 Design Considerations for PEC Type II

PEC Type II was originally conceptualized as multiple parallel baggies connected by ion bridges (see Figure 42). However, this design required protons generated via photocatalytic OER to travel a long distance to reach the ion bridge. This reduced the reaction speed and efficiency and therefore required measures to increase fluid circulation (which are costly due to the large areas of the beds). In addition, there were concerns that the parallel baggie concept would be challenging to deploy since the materials used for the baggie may not be sufficiently tear-resistant if placed on cleared (but otherwise bare) ground. To mitigate these constraints, PEC Type II design has undergone multiple revisions in an attempt to improve the performance, manufacturability, and economics of the concept.

Discussions with Shane Ardo (UC Irvine) and Rohini Bala Chandran (University of Michigan) suggested two major open questions for PEC system design: 1) Is passive mixing sufficient or will active mixing be required?, and 2) How can we ensure that O_2 and H_2 will evolve and stay separated?

To address the first point, SA considered light extinction as a function of bed depth. Ardo estimated that a bed depth of 1-10 cm is sufficient for 40 nm particles at a concentration of 200 nm equivalent thickness (depth of particle layer if settled at bottom). In addition, a pool bed depth of 1 cm may be possible from a photon-capture perspective but is thought to be impractical and further work is necessary to confirm feasibility. In addition, Bala Chandran's group performed calculations that suggested that temperature-induced buoyancy in shallow beds of 1-5 cm may provide adequate mixing via natural convection. Based on these discussions, SA selected a nominal bed depth of 10 cm and assumed that natural convective mixing was sufficient with no additional mixing required. This assumption must be validated experimentally during benchtop and pilot-scale testing.

For the second point, SA considered variations of the baggie-on-baggie concept originally conceptualized by Ardo.[36] A focus was placed on the physical arrangement of the baggies and ion bridge materials that would provide sufficient separation between the evolved O_2 and H_2. The effectiveness of the ion barrier must be validated experimentally during benchtop and pilot-scale testing.

After consideration the above points, the first PEC Type II revision is the baggie-within-baggie concept shown in Figure 46A. This variant hosts the HER process within long cylinders encapsulated by a larger OER process bed. The increased number of HER cylinders reduces the distance protons must travel and therefore improves hydrogen evolution efficiency and production rate. The baggie-within-baggie concept was further iterated as shown in Figure 46B, with the leading concept selected in collaboration with Ardo's research group being the Flexible V-Tube Design. This design further reduced the average distance required to cross the ion bridge, further increasing reaction efficiency. In addition, the concept would not be overly burdensome to manufacture compared to the alternative baggie-within-baggie concepts.

[36] The Ardo concept has several innovative aspects. The primary innovation is a nanoparticle system with optimized absorption wavelengths, that when stacked, grant a higher system efficiency. That aspect is not currently captured in the SA analysis. Instead, the current analysis focused on the physical layout and design of the baggie system.

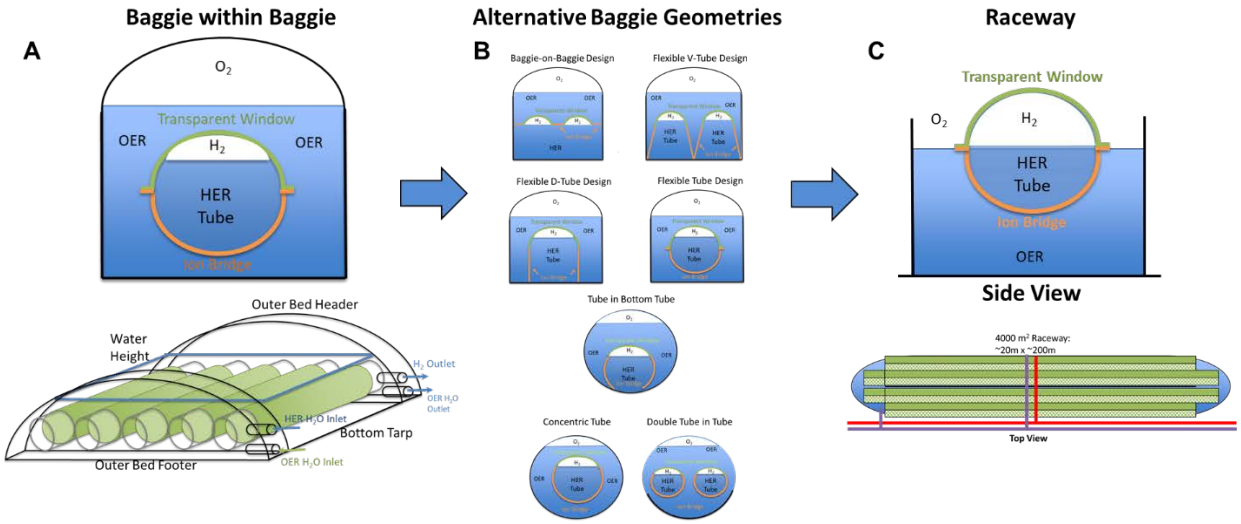


Figure 46 - A) Baggie within baggie concept explored at the beginning of the PEC project period. B) Variants on baggie within baggie concept to improve system performance and cost. C) Raceway with floating baggie concept that increases manufacturability while also minimizing costs. Raceway concept selected for final PEC Type II cost estimate.

The final variant explored was a raceway concept inspired by algae production raceways (Figure 46C). This baggie-on-raceway concept has the OER process occurring in a large, shallow pool (pool depth of 10 cm) with the HER process occurring within floating cylinders (cylinder diameter of 10 cm). This system exposes the pool to the air, thereby venting the oxygen product and reducing the risk of oxygen and hydrogen mixing. As an added benefit, exposure to the air may allow rainwater to collect within the pool and reduce the necessary water feed. (Conversely, it may also lead to higher evaporation.) The baggie-on-raceway concept was selected for the current cost estimate due to implied manufacturability from analogous commercially available algae raceways and enhanced reaction efficiency consistent with a shallow bed system.

8.4 Raceway Design for PEC Type II

The PEC Type II raceway design is inspired by algae production raceways, which have the benefit of already being commercially available and seem to be a capital cost efficient design relative to earlier baggie designs. In particular, MicroBio Engineering, Inc. (Founded 2005) offer algae raceway products up to 1 acre in size. For this study, we also assumed PEC raceways of a 1 acre size.

The bulk raceway pool hosts the OER process with the HER process occurring within long cylinders floating on the surface. A schematic of the proposed design is shown in Figure 47. The top transparent window is made from High Density Polyethylene (HDPE) film while the bottom ion bridge is made from polypropylene filter membrane. It is hypothesized that the cylinders can be manufactured from a large sheet of HDPE and a large sheet of polypropylene filter membrane that are heated in a pattern so as to fuse the two materials together into long cylinders. The buoyancy of the evolved hydrogen gas and the weight of the polypropylene ion bridge are expected to keep a generally circular cross-sectional shape to the HER baggie.[37] The spacing between cylinders is perforated to allow oxygen (evolved from raceway particles) to vent to the atmosphere, in addition to allowing rainwater to pass into the raceway. In this design, we assume that the perforated region is 25% of the total surface area (a 3.3 cm span between cylinders). The specific dimensions of the pool height, the HER cylinders, and the tubing spacing are subject to revision as a function of STH efficiency, diffusion distance of ions, and light collection effectiveness. The ability of the system to promote passive mixing of suspended particles is particularly important to maximize light collection and improve ion diffusion efficiency. Further experimental work will be required to validate these initial design choices.

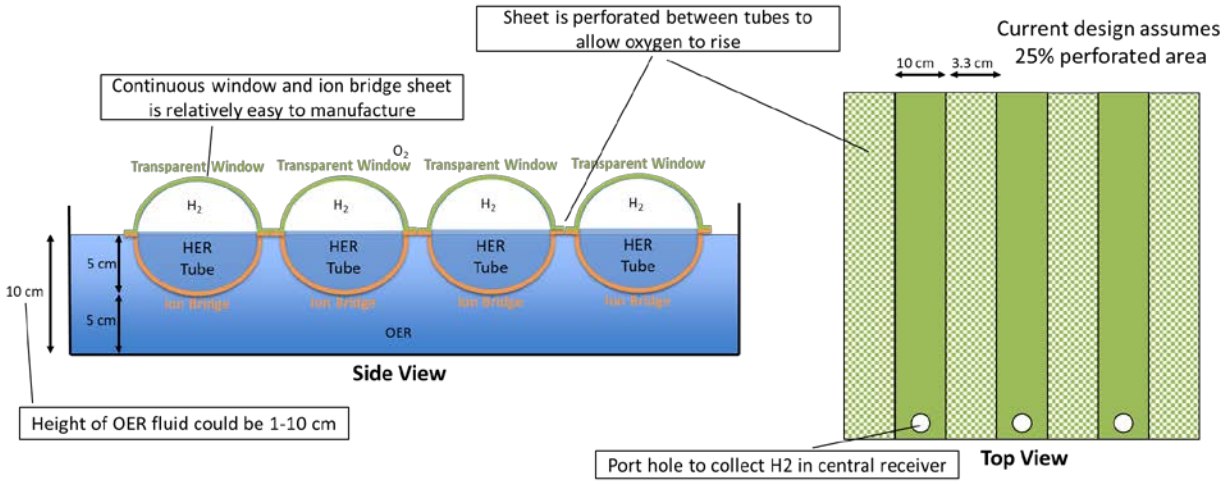


Figure 47 - Side view and top view of floating cylinder system.

[37] Additional weight at the bottom of the baggie may be needed to maintain baggie shape and proper water content.

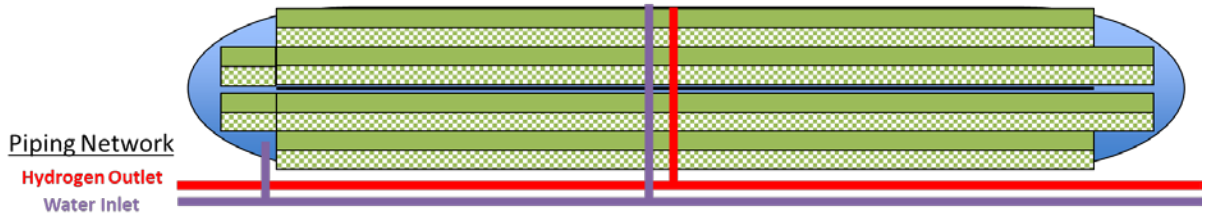


Figure 48 - Top view of floating raceway with proposed piping network.

A top-down view of the whole PEC raceway is shown in Figure 48. Since the liquid level is only 10 cm, the current PEC raceway does not assume a paddlewheel is necessary. If mixing of particles and ions is necessary, an additional system to induce mixing may be required.

Two piping networks are used to distribute water/slurry and to collect the hydrogen product. Each piping network consists of a main branch that connects all the raceways together and a branch pipe that is directly connected to each cylinder via a port valve. The water network has an additional branch that is connected to the pool. The water network provides clean water to both the pool and the cylinders, whereas the hydrogen network collects hydrogen product from the cylinders. During installation and maintenance of the raceways, it is imagined that an isolation valve will close off the main water pipe with the branching water pipes and the catalyst slurry can be distributed into the branches. A water purge can then be run into the pipe branches to clear them of slurry prior to normal operation.

8.5 PEC Type II Catalyst Fabrication

SA conceptually defined a manufacturing process flow for PEC Type II Material Fabrication and used that to develop a range of material costs. Nano-particle materials synthesis and production is an active area of investigation, so a bottom-up cost approach was chosen for flexibility and to allow specification of separate cost components. Solvothermal and hydrothermal are scalable synthesis pathways for bulk production of doped metal oxides and are used as the basis for this study. Other pathways (e.g., vapor deposition, pyrolysis, etc.) and coating options may be addressed as needed.

The general process flow diagram can be seen in Figure 49 below. The final active material cost range is estimated to be $105/kg - $1,200/kg, with assumptions shown in Table 11. With a 50% manufacturing markup, the estimated price range is $158/kg - $1,800/kg. While this cost range is quite wide, the bulk of the cost variation arises from the platinum dopant: with a range of 0.2-2wt% Pt, the resulting cost range is $100/kg - $1,000/kg at a platinum cost of $50k/kg Pt (~$1,500/troy ounce). Synthesis cost excluding Pt cost is $5-$200/kg, which is a useful cost result when postulating non-Pt nanoparticles. This overall nanoparticle active material price range (rounded to $150-$1,800/kg) was used as the bounds of the Type II parametric study described below.

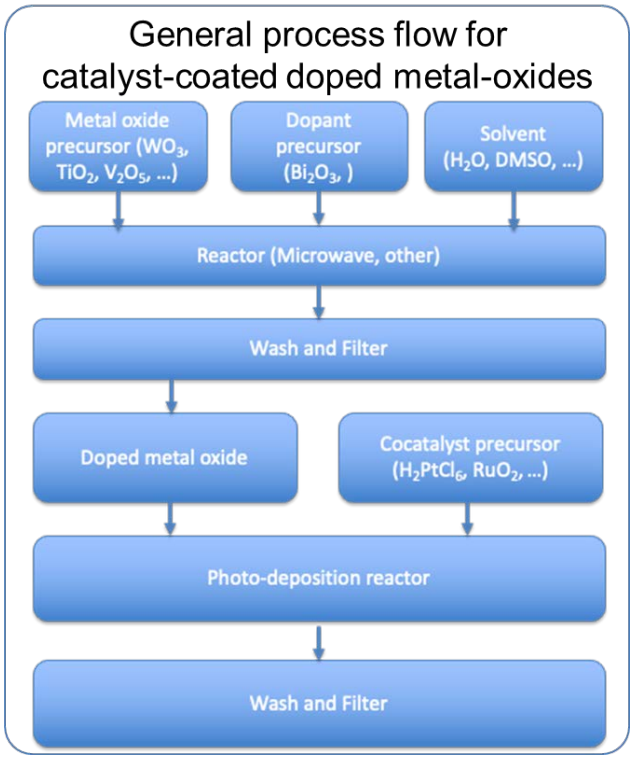


Figure 49 - General process flow for catalyst-coated doped metal-oxides for use as the active material in PEC Type II

Table 11 - SA assumptions for the fabrication of PEC Type II active materials

Item	Expected Range of Values	Rationale
Annual production	60-3,000 tonnes/year	Assumed 100 kg per 1 TPD module, 50 TPD systems, 60-300 systems installed, Particle lifetime: 0.5-5 years
Metal oxide salts	$1-$100/kg	Range of quotes for bulk (10-1,000 kg) orders of metal oxides
Plant capital cost	$2M-$10M	Estimate based on analogous metal organic framework (MOF) analysis scaled for annual material production
Unrecovered solvent costs	$0-$25/kg PEC	Based on analogous MOF work. Range depends on yield, solvent choice, and recovery
Co-catalyst cost	$100-$1,000/kg PEC	Modeled as 0.2-2 wt% Pt:metal oxide, $50k/kgPt
Range of nanoparticle costs	$105/kg PEC - $1,200/kg PEC	Materials + Synthesis (approx. first-pass range of particle price)
Range of nanoparticle prices	$150/kg PEC - $1,800/kg PEC	50% Manufacturer markup assumed

8.6 PEC Type II Case Study

8.6.1 General PEC Type II Assumptions

A modular PEC design is envisioned in this analysis where each module has a capacity of 1,000 kg H_2/day. Multiple modules are strung together to reach the desired H_2 production. Key parameters are listed in Table 12.

Table 12 - PEC Type II Design Specifications

System Parameters	Units	Value
PEC Type	-	Type II
Average Insolation	kWh m^{-2} day^{-1}	5.77
STH Efficiency	%	8%
Average H_2 Mass Flow	kg day^{-1}	1,000
Area Specific Mass Flow	kg H_2 hr^{-1} m^{-2}	5.89E-04
Total Solar Area Required	m^2	70,790
Raceway Length	m	200
Raceway Width	m	20
Raceway Height	m	0.01
Raceway Area	m^2	3800
Floating Cylinder Width	m	0.01
Number of Floating Cylinders per Raceway	#	1425
Number of Raceways	#	19
Assumed Particle Density	kg m^{-2}	0.00105
Particle Mass	kg	74

8.6.2 Labor Assumptions

In light of the new raceway concept, we reviewed the labor costs associated with the PEC II concept. Borrowing from recent NREL reports,[38] the labor count was extrapolated from a bottom-up labor estimate of a 10 acre algae pond. The labor assumptions used for PEC Type II are shown in Table 13.

[38] NREL, 2016. Process Design and Economics for the Production of Algal Biomass: Algal Biomass Production in Open Pond Systems and Processing Through Dewatering for Downstream Conversion.

Table 13 - Labor Assumptions for 50 TPD PEC II Plant

Role	FTE	Rationale (compared to 10 acre NREL analysis)
Plant Manager	1	Constant
Plant Engineer (Civil)	2	Constant
Plant Engineer (Env.)	2	Constant
Maintenance Supervisor	1	Constant
Maintenance Tech	5	Extrapolated from # of raceway ponds
Lab Manager	1	Constant
Lab Tech	3	Extrapolated from # of raceway ponds
Shift Supervisor	2	Extrapolated from # of direct reports
Module Operator – Production	25	Extrapolated from # of raceway ponds
Clerks & Secretaries	1	Extrapolated from labor total
Labor Total	**43**	**50 TPD Plant**
Labor Total	**0.86 FTE/TPD**	**FTE per H_2 tonne per day**

Using the previous labor assumptions, Equation (**4**) can be used to estimate the labor needed for a module within a 50 TPD raceway-based facility:

$$Module\ FTE = \left(\frac{0.0042\ FTE}{Particle\ lifetime\ (yrs)} + 0.018\ FTE\right)\frac{(Solar\ Land\ Required\ in\ m2)}{4000m^2} + \frac{13\ FTE}{50\ TPD}(Module\ H2\ Production\ in\ TPD) \quad (4)$$

The first term describes the maintenance tech labor required for replacement of photoactive particles at the intervals required and scales with particle lifetime. The second term refers to module operator labor used to run the raceways. The solar land required refers to the total area covered by photoactive particles and scales with the number of ponds required to meet the H_2 production target. The third term represents personnel necessary for managing the entire 50 TPD facility including the plant manager, plant engineers, supervisors, lab managers, and office staff. For a 1 TPD H_2 production module, this yields an estimate of 0.86 FTE of annual labor compared to the previous estimate of 3.06 FTE derived based on solar panel installations. We believe the new estimate better represents the PEC raceway concept in scale and will be a better estimate for H2A cases going forward.

8.6.3 Particle Density

Light extinction analysis conducted by Eric McFarland at UCSB indicates that a bed depth of 10 cm for a nanoparticle size of 40 nm and a particle concentration of 200 nm equivalent thickness of 40 nm particles is sufficient to capture all light entering the bed. Equivalent thickness is defined as the depth of the particle layer if all particles settled to the bottom. For purposes of cost estimation, the PEC nanoparticles are modeled as 40 nm particles of iron oxide (Fe_2O_3) upon which 5 nm layers of TiO_2 have been deposited. Assuming even distribution, the required particle areal density is 0.00105 kg/m^2 for complete light extinction.

8.6.4 Product Specifications

Similar to the 2009 PEC report,[39] PEC Type II is assumed to use 1 condenser and 2 intercoolers coupled with a hydrogen compressor to achieve product gas output purity of 99.6% H_2 and 0.4% water vapor. The hydrogen is cooled to 40℃ using cooling water to achieve this product purity. This is consistent with the hydrogen purity required for industrial applications but may be insufficient for transportation applications (which typically specify 99.999% H_2 purity). Further cost analysis for hydrogen compression and dehydration methods should be explored to understand how the levelized cost of hydrogen changes.

8.6.5 H2A Assumptions

The standard H2A workbook has continued to evolve since the 2009 PEC report.[36] Where appropriate, the changes implemented in this cost study are shown in Table 14, Table 15, and Table 16.

[39] James, Brian D., George N. Baum, Julie Perez, and Kevin N. Baum. "Technoeconomic Analysis of Photoelectrochemical (PEC) Hydrogen Production," December 1, 2009. https://doi.org/10.2172/1218403.

Table 14 - PEC Type II H2A Financial Parameters

Financial Parameters	2009 Analysis Value	2021 Analysis Value
Operating Period	20 years	
Facility Life	20 years	
Construction Period and Cash Flow	1 year	20% Year 1, 80% Year 2
Installation Cost Factor	1.3	1.4
Land Cost	$500/acre	$5000/acre
Property Taxes and Business Insurance	2%/year of the total initial capital cost	
IRR	10% after tax	8% after tax
Working Capital Rate	15% of the annual change in total operating costs	
Income Taxes	35% Federal, 6% State, 38.9% effective	21% Federal, 6% State, 25.74% Effective
Sales Tax	Not included - facilities and related purchases are wholesale and through a general contractor entity	
Decommissioning	10% of initial capital	
Salvage Value	10% of initial capital	

Table 15 - PEC Type II H2A Operating Parameters

Operating Parameters	2009 and 2021 Analysis Value
Hydrogen Pressure at Central Gate	300 psig
Hydrogen Purity	98% minimum; CO < 10 ppm, sulfur < 10 ppm [99.6% H_2, Industrial purity assumed]
Burdened Labor Rate for Staff	$50/hour
G&A Rate	20% of the staff labor costs

Table 16 - PEC Type II H2A System Parameters

System Parameters	2009 Analysis Value	2021 Analysis Value
Operating Capacity Factor	90%	
Site Preparation	1% of direct costs minus unique excavation costs	2% (No unique excavation)
Engineering & Design	7% of direct costs (compared to default of 13% due to modularity)	7% of direct costs
Process Contingency	20% of direct costs (compared to default of 15% due to greater uncertainties in system configuration)	20% of direct costs
Project Contingency	$0	$0
Up-Front Permitting Costs	0.5% of direct capital costs (compared to default of 9% due to modularity of design and environmental benefits)	5% of direct capital costs
Annual Maintenance & Repairs	0.5% of direct capital costs	0.5% of direct capital costs

8.6.6 PEC Type II Case Study Results

At the time the PEC study was performed, the target levelized cost for green hydrogen was \$2/kg H_2 by 2030. A set of possible conditions were selected as a case study for what future PEC Type II facility would cost: 8% STH, 1 year catalyst particle lifetime, and ~\$150/kg particle price. The floating cylinder materials were assumed to be replaced every 5 years and a 50% manufacturer markup was included in the price of the combined floating cylinder materials. A bill of materials for this 1 TPD module is shown in Table 17.

Table 17 - PEC Type II Case Study Bill of Materials (1 TPD H_2 Modules)

Component	Units	Cost Unit	Cost per Unit	Total Cost
PEC Particle	74	$/kg	$150	$11,150
Cylinder Material Top	103109	$/m^2	$0.51	$52,585
Cylinder Material Bottom	85059	$/m^2	$1.74	$148,002
Geomembrane	72200	$/m^2	$4.00	$288,800
Circulation Pump	19	$/pump	$500	$9,500
Port Hardware	54150	$/port	$4.50	$243,675
Port Installation	54150	$/port	$4.50	$243,675
Hydrogen PVC Piping	4180	$/m	$0.82	$3,428
Water Piping	4218	$/m	$1.71	$7,213
Raceway Wiring Panel	19	$/raceway	$117	$2,223
Water Level Controllers	19	$/raceway	$40	$760
Pressure Sensors	19	$/raceway	$278	$5,282
Hydrogen Area Sensors	19	$/raceway	$299	$5,681
Instrument Wiring	19	$/raceway	$77	$1,463
Power Wiring	19	$/raceway	$39	$741
Conduit	19	$/raceway	$103	$1,957
H2 Compressor	1	$/module	$872,400	$872,400
HX - Condenser	1	$/module	$10,626	$10,626
HX - Intercooler 1	1	$/module	$11,464	$11,464
HX - Intercooler 2	1	$/module	$11,870	$11,870
Makeup Water Pump	1	$/module	$213	$213
Forklift	1	$/module	$950	$950
Bag Unroller	1	$/module	$37,000	$37,000
Computer & Monitor	1	$/module	$1,500.00	$1,500
Labview	1	$/module	$4,300.00	$4,300
PLC	1	$/module	$2,000.00	$2,000
Control Room Building	1	$/module	$8,000.00	$8,000
Control Room Wiring Panel	1	$/module	$3,000.00	$3,000
Hydrogen Flow Meter	1	$/module	$5,500.00	$5,500

The uninstalled capital cost for the 1 TPD module is $1.99M and the resulting levelized cost of hydrogen is $1.94/kg H_2.

8.7 PEC Type II Sensitivity Study

SA performed a sensitivity study on several input parameters for the PEC Type II system. This includes active material cost, active material lifetime, and STH efficiency. Results for PEC Type II are shown in Figure 51 and Figure 51.

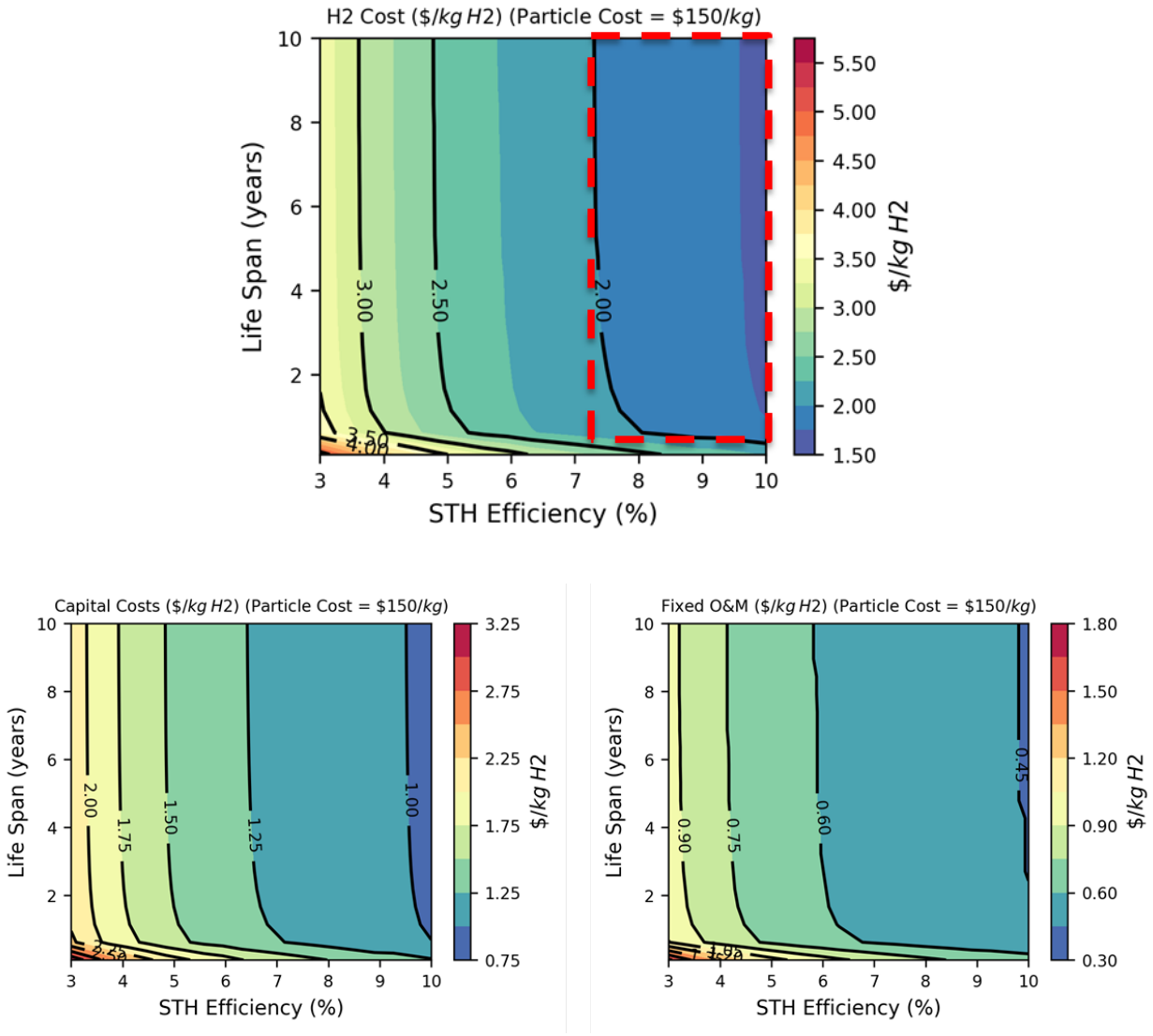


Figure 50 - PEC Type II results for an active material cost of $105/kg. Optimal range for achieving an H_2 cost of $2/kg H_2 is highlighted by a dashed red box.

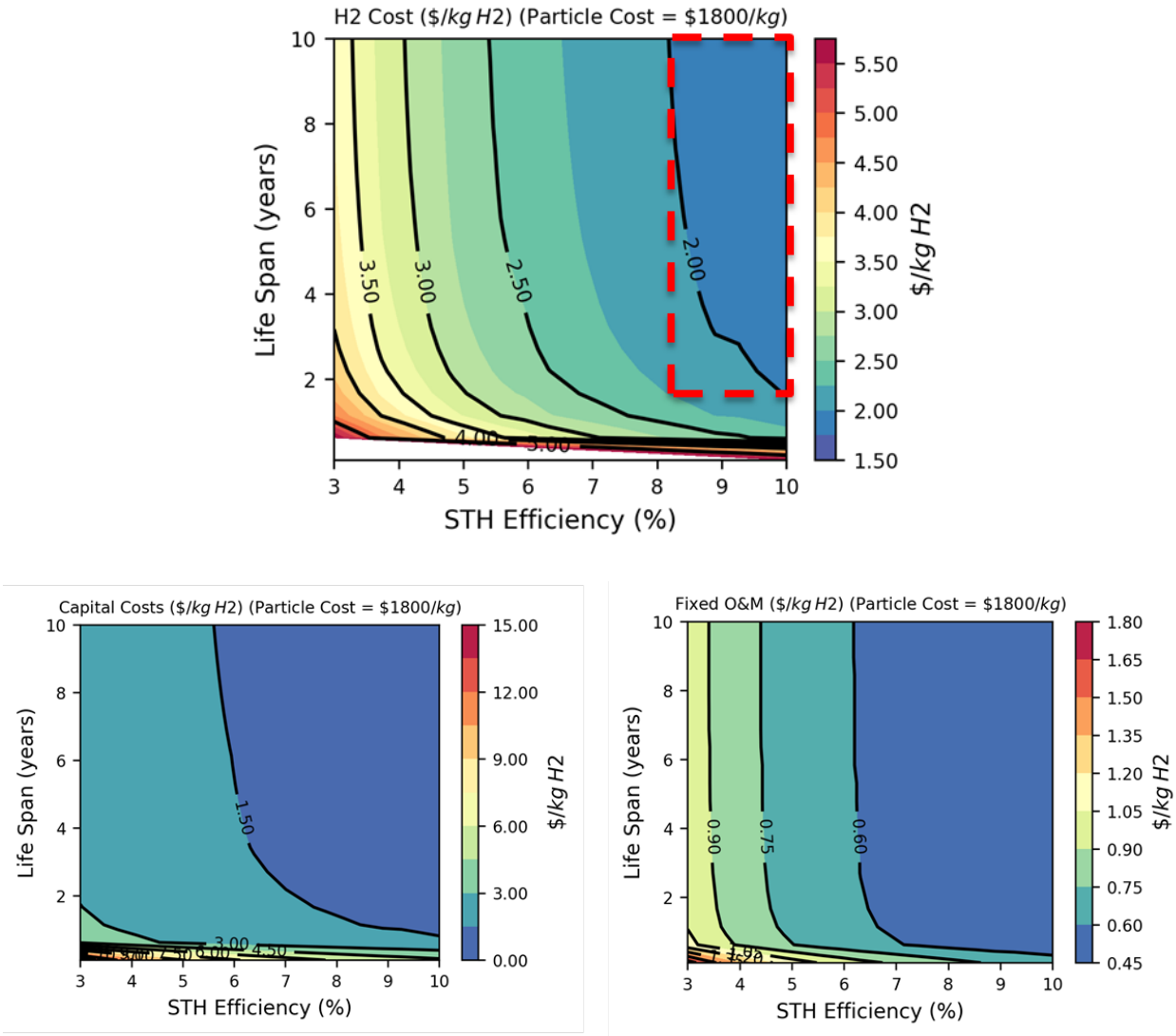


Figure 51 - PEC Type II results for an active material cost of $1,200/kg. Optimal range for achieving a H_2 cost of $2/kg H_2 is highlighted by a dashed red box.

As seen in Figure 51, the STH efficiency must >7% with a particle lifetime of >1 year to achieve a target H_2 cost of $2/kg. The acceptable STH efficiency and particle lifetime become more stringent as the particle cost increases, with the approximate minimum requirements moving to an STH Efficiency of 8% and a particle lifetime of >3 years.

The cost of H_2 for the SA was broken out into its constituent parts including capital cost and labor, which allowed SA to observe that the cost of labor and maintenance can account for 30%-40% of H_2 with capital cost taking up the majority of the balance.

8.8 Design Considerations for PEC Type IV

PEC Type IV conceptualizes a photoelectrochemical electrode surrounded by water to convert concentrated solar energy into hydrogen and oxygen. The PEC receiver panel and water are pressurized to 300 psi to eliminate the need for a downstream H_2 compressor and to prevent vaporization of the water at higher temperatures due to thermal energy accumulated via concentrated solar flux. As the solar concentration ratio (CR) increases, the levelized cost of hydrogen from a PEC Type IV system is expected to decrease since the PEC electrode area will be lower and thus require less catalyst and capital investment. However, a higher concentration ratio leads to a higher water temperature, which will eventually cause water vaporization and prevent the exchange of protons between the OER and HER sides of the PEC electrode. At 300 psi, the boiling temperature of water is ~214 ℃, which limits the highest allowable water temperature under maximal solar load for PEC Type IV. A thermal calculation was performed as a function of concentration ratio and solar to hydrogen conversion efficiency to determine the maximum allowable concentration ratio (for a pressure of 300 psi).

A rigorous thermal balance was calculated for the electrode material, the water above and below the electrode, and the plexiglass on the top and bottom of the receiver panel. A schematic is shown in Figure 52. The calculations accounted for the radiative heat transfer and convective heat transfer between regions of interest. The radiative heat transfer included transfer between the electrode and the covers, and between the covers and the atmosphere. The convective heat transfer included natural convection of the water and convection induced by wind on the surface of the covers.

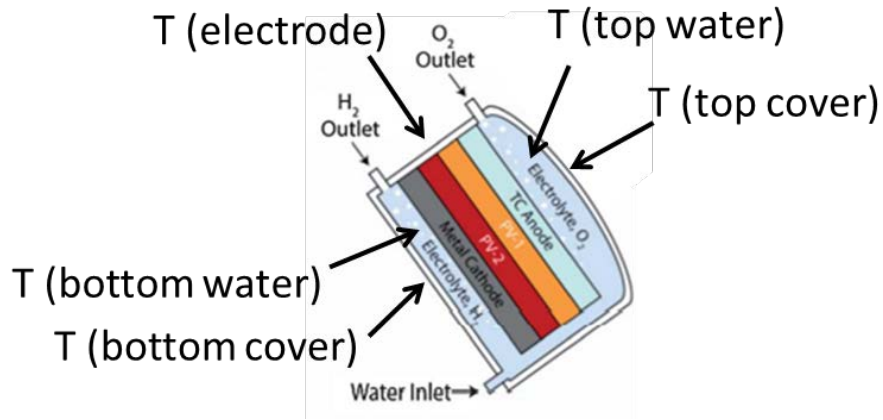


Figure 52 - A steady-state thermal calculation was performed for the five distinct regions shown in the figure: electrode, bottom water region, top water region, bottom panel cover, and top panel cover. (Image from Pinaud et al.)

Since the ion exchange is necessary for completing the HER and OER half-reactions, it was assumed that at steady-state, no water flowed out of the system in liquid form. The panel was assumed to lie at 45°, which helps promote natural convection, in addition to being required for solar collection from the concentrating panels. Ambient temperature was assumed to be 25 ℃ and atmospheric temperature (for heat rejection to the sky) was assumed to be 10 ℃. Wind speed was set at 5 m/s, to match the global average wind speed. The receiver panel is assumed to be a series of parallel plates, with the electrode

thickness being 7 cm, the distance between the electrode and cover being 1.5 cm, and the overall panel having a thickness of 10 cm.

A case study was performed for an STH of 25% and a concentration ratio of 20 with results shown in Table 18. The results suggest that natural convection of water combined with radiative heat transfer ensures that the largest temperature differential is limited to ~10℃ (between the electrode and the top cover).

Table 18 - Steady-State Temperatures for PEC Receiver Panel with STH of 25% and CR of 20

Component	Temperature
Top Cover	175 ℃
Top Water	180 ℃
Electrode	185 ℃
Bottom Water	184 ℃
Bottom Cover	183 ℃

A sensitivity study on the electrode temperature was performed as a function of the solar-to-hydrogen efficiency and the concentration. A temperature map was produced in Figure 53, with curves marked for atmospheric boiling of water (100 ℃) and 300 psi boiling of water (214.08 ℃). At 300 psi, the concentration ratio can increase up to 30 before the boiling temperature constraint is met. If an unpressurized system is required, then the concentration ratio will be limited to ~10.

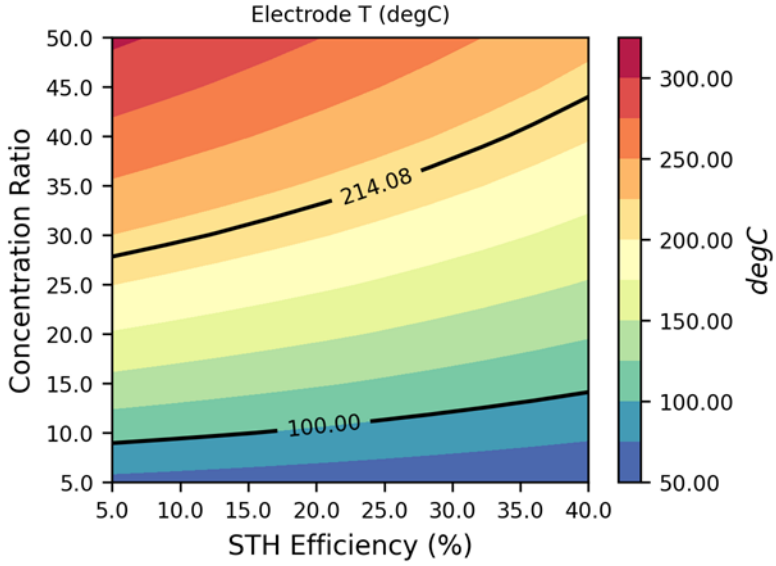


Figure 53 - Electrode temperature as a function of STH Efficiency and Concentration Ratio. Curves for atmospheric boiling of water (100 ℃) and 300 psi boiling of water (214.08 ℃) are marked on the figure.

Consistent with Figure 53, the upper and lower bound for concentration ratio were set to 30 and 10, respectively. Future thermal studies can improve the model by updating the wind heat transfer coefficient as a function of cover temperature and rigorously calculating thermal losses incurred through the edges and supports.

8.9 PEC Type IV Electrode Fabrication

The biggest capital cost driver for PEC Type IV is the concentrator panel, followed by the PEC electrode. Depending on the cost of the PEC electrode, PEC Type IV is predicted to have dramatically different levelized costs of hydrogen. Studies by NREL (Todd Deutsch) have used a dramatically higher electrode cost than that used by other authors such as SA. Table 19 shows sample parameter values used in alternative comparison studies performed by NREL (Todd Deutsch) vs SA for a Type IV PEC panel. In order to understand what PEC panel costs are possible now and in the future, a DFMA® cost analysis was performed. The cost analysis is based on a nominal system with a 15% STH efficiency and 10:1 solar concentration ratio.

Table 19 - Parametric comparison study between NREL and SA on a Type IV PEC panel

Parameter	NREL (Todd Deutsch)	SA
PEC Absorber Cost ($/m² USD 2016)	2,300 - 34,700	100 - 300
PEC Absorber Lifetime (years)	0.001 - 0.5 (Current Status)	1 - 10

Figure 54 shows a generic Type IV PEC panel with a tracking concentrator system. [40] The parabolic cylinder tracking concentrator system focuses sunlight on a linear PEC panel, which consists of perovskite/perovskite tandem cells that generate the electricity to split the water within the plexiglass encasement. The DFMA® model for the Type IV PEC panel was based on the works of Li et al[41] and Song et al[42] from the University of Toledo on cost analysis of a perovskite/perovskite tandem photoelectrode PV panel, with appropriate adjustments made for our Type IV PEC panel such as adding PEC catalyst layers.

[40] Pinaud, Blaise A., et al. "Technical and economic feasibility of centralized facilities for solar hydrogen production via photocatalysis and photoelectrochemistry." *Energy & Environmental Science* 6.7 (2013): 1983-2002.
[41] Li, Zongqi, et al. "Cost analysis of perovskite tandem photovoltaics." *Joule* 2.8 (2018): 1559-1572.
[42] Song, Zhaoning, et al. "A technoeconomic analysis of perovskite solar module manufacturing with low-cost materials and techniques." *Energy & Environmental Science* 10.6 (2017): 1297-1305.

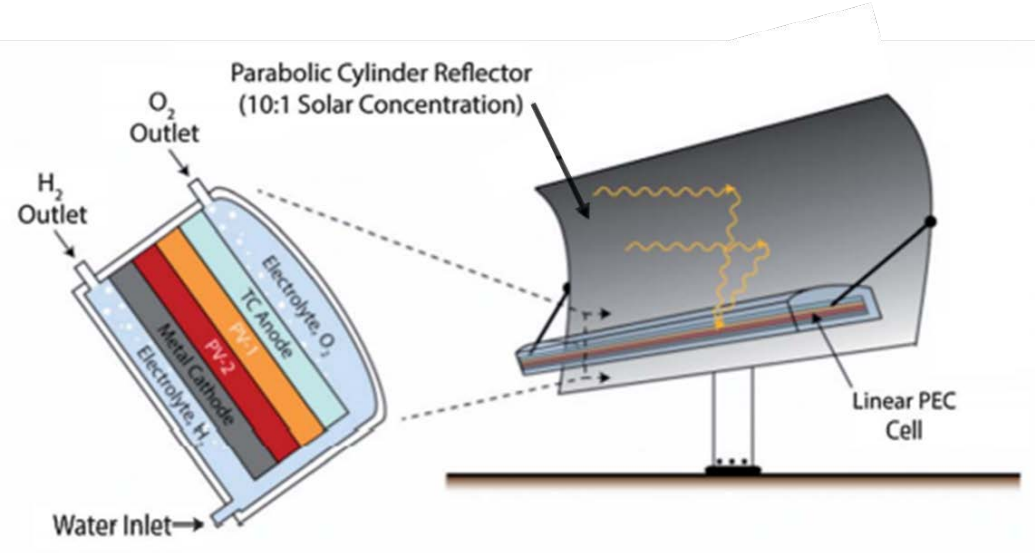


Figure 54 - Generic Type IV PEC panel with a tracking concentrator system (Image from Pinaud et al.)

The dimensions for our Type IV PEC panel are 1 m x 2 m x 0.70182 mm. Figure 55 shows the Type IV PEC panel configuration along with the thickness for each layer. Our Type IV PEC panel was modeled based on the panel configurations from Li et al[41] and Yan et al from the University of Toledo in their 2021 AMR presentation.[43] The stainless steel substrate is the cathode, and the ITO layer is the anode. NiO is the hole transport layer, SnO_2 is the electron transport layer, and C_{60} is the tunnel junction layer. PEDOT:PSS is the interconnecting layer to integrate the thick layer of $FAPbI_3$ (low band gap of 1.45 eV) and the thin layer of $CsPbIBr_2$ (high band gap of 1.9 eV). An ultrathin TiN layer (5 nm) is deposited on each electrode to protect it from photocorrosion and moisture ingress. An Ir catalyst is deposited on the O_2 reaction side, and a Pt catalyst is deposited on the H_2 reaction side, to facilitate the respective gas evolution half-reactions.

[43] https://www.hydrogen.energy.gov/pdfs/review21/p191_yan_2021_o.pdf

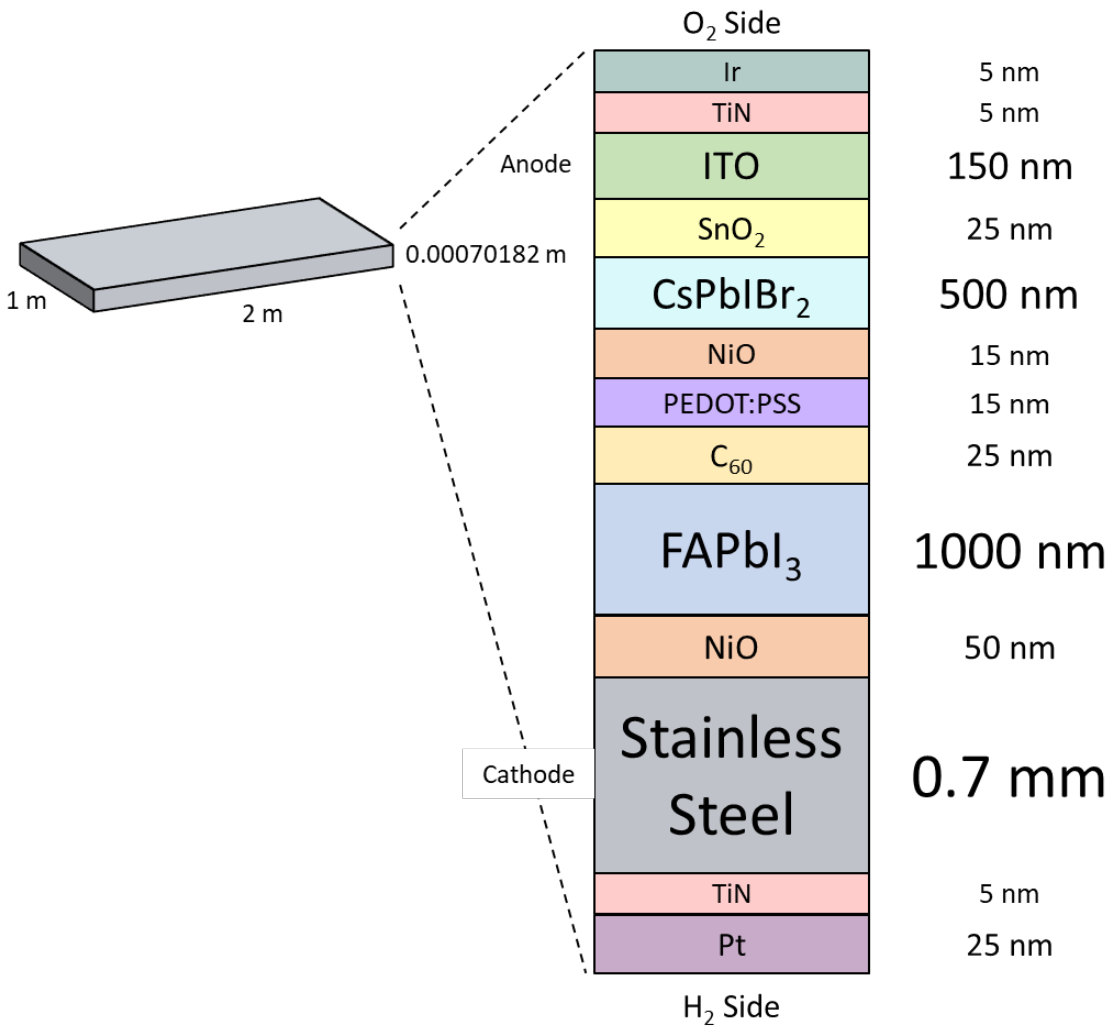


Figure 55 - Type IV PEC panel configuration

Table 20 shows the process flow for the Type IV PEC panel. The entire manufacturing process consists of 13 stages that were modeled separately. The 13 stages include:

- 5 sputtering stages,
- 1 evaporation stage,
- 6 screen printing stages, and
- 1 ultrasonic bath cleaning stage.

The data for the input parameters of the DFMA® model was obtained from the works of Li et al[41] and Song et al[42] from the University of Toledo. Table 21 shows the Type IV PEC panel material input parameters (price, density, thickness).

Table **22** shows the equipment input parameters (active area, capital cost, laborers, power consumption, operating time, material utilization). For the equipment input parameters listed as "per line" (capital cost, laborers, power consumption), the "per line" refers only to the listed equipment's section of the entire process line, which includes the 13 stages. Table 23 shows the process input parameters.

Table 20 - Process flow for the Type IV PEC panel

PEC Panel Process Flow
Stage 1 - Stainless Steel Cleaning
Stage 2 - NiO Screen Printing 1
Stage 3 - $FAPbI_3$ Screen Printing
Stage 4 - C_{60} Evaporation
Stage 5 - PEDOT:PSS Screen Printing
Stage 6 - NiO Screen Printing 2
Stage 7 - $CsPbIBr_2$ Screen Printing
Stage 8 - SnO_2 Screen Printing
Stage 9 - ITO Sputtering
Stage 10 - TiN Sputtering (O_2 Side)
Stage 11 - Ir Sputtering (O_2 Side)
Stage 12 - TiN Sputtering (H_2 Side)
Stage 13 - Pt Sputtering (H_2 Side)

Table 21 - Material input parameters for the Type IV PEC panel

Material	Material Price ($/kg)	Material Density (g/cc)	Thickness
Stainless Steel Sheeting	$6.94	7.93	0.7 mm
NiO Layer 1	$200	6.67	50 nm
$FAPbI_3$ Layer	$700	6.16	1000 nm
C_{60} Layer	$160,000	1.65	25 nm
PEDOT:PSS Layer	$2,700	1.45	15 nm
NiO Layer 2	$200	6.67	15 nm
$CsPbIBr_2$ Layer	$700	6.16	500 nm
SnO_2 Layer	$100	6.95	25 nm
ITO Layer	$3,330	7.14	150 nm
TiN Layer (O_2 Side)	$6,684	5.4	5 nm
Ir Layer	$160,754 ($5,000/Tr.Oz)	22.56 (0.01128 mg/cm^2)	5 nm
TiN Layer (H_2 Side)	$6,684	5.4	5 nm
Pt Layer	$48,226 ($1,500/Tr.Oz)	21.45 (0.05363 mg/cm^2)	25 nm

Table 22 - Equipment input parameters for the Type IV PEC panel

Equipment	Active Area (m x m)	Capital Cost Per Line ($/line)	Laborers Per Line (laborers/line)	Power Consumption Per Line (kW/line)	Operating Time: Per Cycle (min/cycle) Per Panel (min/panel)	Material Utilization (%)
Ultrasonic Bath	2 x 3	$7,500	0.5	28	(76 panels/cycle) 6 0.079	n/a
Screen Printer	2 x 2.5	$200,000	0.15	10	(1 panel/cycle) 0.25 0.25	80%
Evaporator	5 x 12	$500,000	0.25	840	(20 panels/cycle) 60 3	60%
Sputterer	2.5 x 10	$2,000,000	0.25	500	(9 panels/cycle) 27 (5 nm), 28 (25 nm), 30 (150 nm) 3 (5 nm), 3.11 (25 nm), 3.33 (150 nm)	80%

Table 23 - Process input parameters for the Type IV PEC panel

PEC Panel Process Parameters	
Possible Runtime Per Year Per Line (hr/year/line)	3,360
Interest Rate (%)	10%
Corporate Income Tax Rate (%)	40%
Default Machine Lifetime (years)	15
Average Equipment Installation Factor	1.4
Average Maintenance/Spare Parts (% of CC) Per Year (%/year)	10%
Average Miscellaneous Expenses (% of CC) Per Year (%/year)	7%
Average Labor Rate ($/hr/laborer)	$48
Electricity Utility Cost ($/kWh)	$0.0787

8.10 PEC Type IV Electrode Fabrication Cost Results

Table 24 shows the process results per line for the Type IV PEC panel. Here, the "per line" refers to the entire 13-stage process line. The total capital cost per line is $11,707,500 with a maximum possible production capacity of ~120,000 m^2/year (60,000 panels/year) based on an effective panel production rate of 3.33 min/panel/line for a 2 m^2 panel.

Table 24 - Process results per line for the Type IV PEC panel

PEC Panel Process Results		
Annual Production Capacity (for a 2 m^2 panel)	**Area Per Year (m^2/year)**	120,000
	Panels Per Year (panels/year)	60,000
Total Capital Cost ($/line)		$11,707,500
Total Laborers (laborers/line)		2.9
Total Power Consumption (kW/line)		3,428
Effective Panel Production Rate (min/panel/line)*		3.33

* Effective rate panels are produced per line based on simultaneous processing of each of the 13 manufacturing stages

Table 25 shows the tabular cost results per annual production rate for the Type IV PEC panel from our DFMA® model. Six different annual production rates were analyzed (from 1,000 m^2/year to 10,000,000 m^2/year). The corresponding panels per year (panels/year), power input to electrolyzer per year (MW-in/year), and mass of H_2 produced per year (kg H_2/year) are also detailed in Table 25. The results of our model are shown in various cost forms, including $/$m^2$ (total, material, manufacturing), $/kW-in, $/kg H_2, $/panel, and $/year. The results for the highest production volume (10,000,000 m^2/year) show a corresponding cost of $160.31/$m^2$, with a material cost of $117.71/$m^2$ (73% of the total cost) and a manufacturing cost of $42.60/$m^2$ (27% of the total cost); and in the other cost forms, the corresponding costs are $234.30/kW-in, $1.28/kg H_2, $321/panel, and $1,603,099,357/year. Figure 56 graphically shows the cost results ($/$m^2$) per annual production rate (m^2/year) for the Type IV PEC panel.

Table 25 - Tabular cost results per annual production rate for the Type IV PEC panel

Annual Production Rate						
Area Per Year (m^2/year)	1,000	10,000	50,000	100,000	1,000,000	10,000,000
Power Input To Electrolyzer Per Year (MW-in/year)	1	7	34	68	684	6,842
Mass Of H_2 Produced Per Year (kg H_2/year)	125,049	1,250,489	6,252,443	12,504,887	125,048,866	1,250,488,657
Panels Per Year (panels/year)	500	5,000	25,000	50,000	500,000	5,000,000

PEC Panel Results						
PEC Panel Cost Per Area ($/m²)	$4,979.89	$611.75	$223.47	$174.94	$162.56	$160.31
PEC Panel Material Cost Per Area ($/m²)	$117.71	$117.71	$117.71	$117.71	$117.71	$117.71
PEC Panel Manufacturing Cost Per Area ($/m²)	$4,862.18	$494.03	$105.76	$57.22	$44.84	$42.60
PEC Panel Cost Per Power Input To Electrolyzer ($/kW-in)	$7,278.42	$894.11	$326.62	$255.68	$237.58	$234.30
PEC Panel Cost Per Mass Of H_2 Produced ($/kg H_2)	$39.82	$4.89	$1.79	$1.40	$1.30	$1.28
PEC Panel Cost ($/panel)	$9,960	$1,223	$447	$350	$325	$321
PEC Panel Annual Cost ($/year)	$4,979,893	$6,117,491	$11,173,578	$17,493,775	$162,555,242	$1,603,099,357

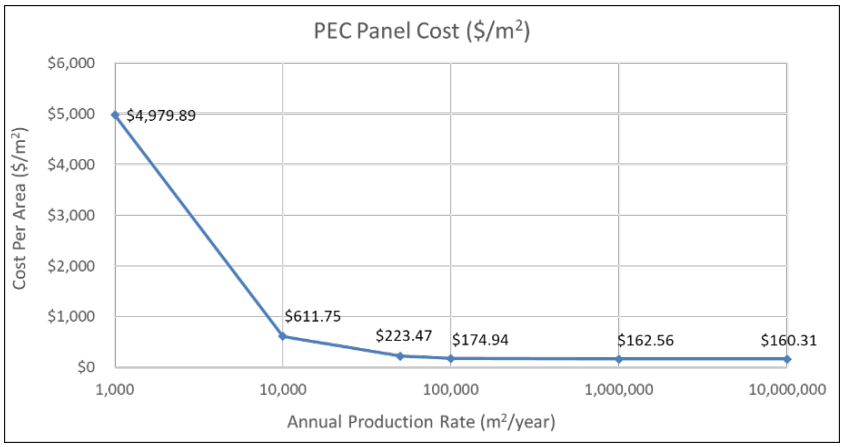


Figure 56 - Graphical cost results ($/m²) per annual production rate (m²/year) for the Type IV PEC panel

Table 26 shows the tabular cost per stage breakdown ($/m^2) per annual production rate for the Type IV PEC panel. Figure 57 graphically shows the cost per stage breakdown/comparison ($/m^2) between material and manufacturing costs at a 10,000,000 m^2/year production rate. At the highest production volume of 10,000,000 m^2/year, the major cost drivers are the material costs for the stainless steel substrate, Pt catalyst, Ir catalyst, and C_{60} layer; and the moderate cost drivers are the material costs for the perovskite layers and ITO layer and the manufacturing costs for sputtering and evaporation. The manufacturing costs for screen printing have a minimal cost contribution. Alternative manufacturing processes, materials, and thicknesses are being explored to further reduce the cost of the Type IV PEC panel.

Table 26 - Tabular cost per stage breakdown ($/m²) per annual production rate for the Type IV PEC panel

Annual Production Rate						
Area Per Year (m²/year)	1,000	10,000	50,000	100,000	1,000,000	10,000,000
Power Input To Electrolyzer Per Year (MW-in/year)	1	7	34	68	684	6,842
Mass Of H_2 Produced Per Year (kg H_2/year)	125,049	1,250,489	6,252,443	12,504,887	125,048,866	1,250,488,657
Panels Per Year (panels/year)	500	5,000	25,000	50,000	500,000	5,000,000
Stage 1 - Stainless Steel Cleaning						
Total Cost Per Area ($/m²)	$41.66	$38.85	$38.60	$38.57	$38.54	$38.54
Stage 2 - NiO Screen Printing 1						
Total Cost Per Area ($/m²)	$83.01	$8.39	$1.76	$0.93	$0.18	$0.16
Stage 3 - $FAPbI_3$ Screen Printing						
Total Cost Per Area ($/m²)	$88.32	$13.70	$7.07	$6.24	$5.49	$5.47
Stage 4 - C_{60} Evaporation						
Total Cost Per Area ($/m²)	$220.25	$33.68	$17.10	$15.03	$14.61	$14.51
Stage 5 - PEDOT:PSS Screen Printing						
Total Cost Per Area ($/m²)	$83.00	$8.38	$1.75	$0.92	$0.17	$0.15
Stage 6 - NiO Screen Printing 2						
Total Cost Per Area ($/m²)	$82.95	$8.33	$1.70	$0.87	$0.13	$0.10
Stage 7 - $CsPbIBr_2$ Screen Printing						
Total Cost Per Area ($/m²)	$85.62	$11.00	$4.37	$3.54	$2.80	$2.77
Stage 8 - SnO_2 Screen Printing						
Total Cost Per Area ($/m²)	$82.95	$8.33	$1.70	$0.87	$0.12	$0.10
Stage 9 - ITO Sputtering						
Total Cost Per Area ($/m²)	$835.01	$88.80	$22.47	$14.18	$13.35	$12.77
Stage 10 - TiN Sputtering (O_2 Side)						
Total Cost Per Area ($/m²)	$830.63	$84.42	$18.09	$9.80	$8.14	$7.73
Stage 11 - Ir Sputtering (O_2 Side)						
Total Cost Per Area ($/m²)	$853.07	$106.86	$40.53	$32.24	$30.58	$30.17
Stage 12 - TiN Sputtering (H_2 Side)						
Total Cost Per Area ($/m²)	$830.63	$84.42	$18.09	$9.80	$8.14	$7.73
Stage 13 - Pt Sputtering (H_2 Side)						
Total Cost Per Area ($/m²)	$862.78	$116.57	$50.24	$41.95	$40.29	$40.13

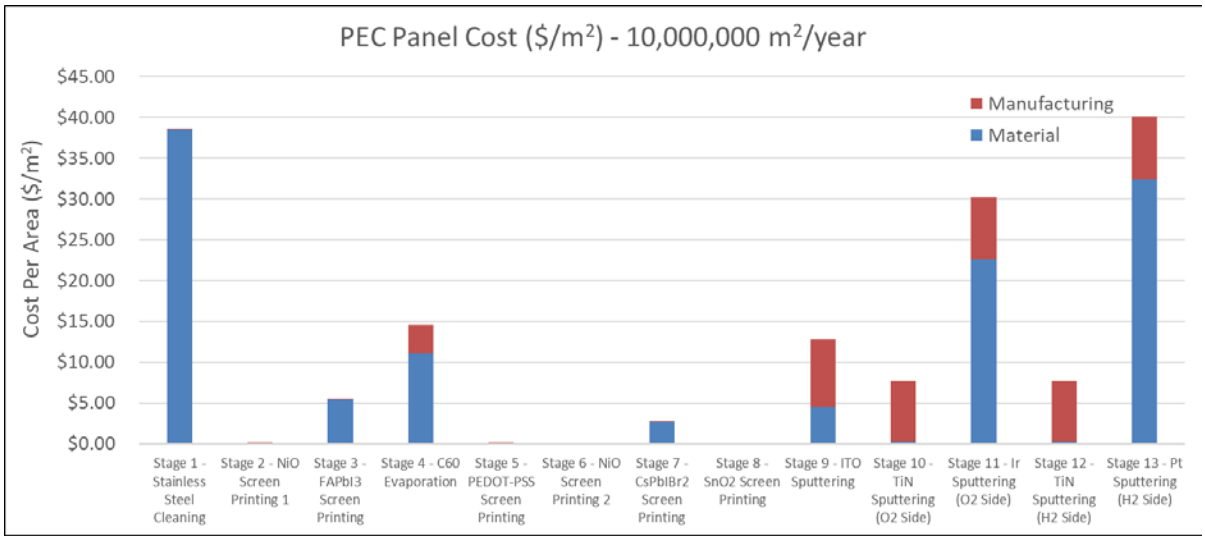


Figure 57 - Graphical cost per stage breakdown/comparison ($/m²) between material and manufacturing costs at a 10,000,000 m²/year production rate for the Type IV PEC panel

8.11 Facility Design for PEC Type IV

The 2009 PEC report imagined long rows of concentrating panels alternating with large empty stretches of land between rows to prevent solar shadowing.[44] This current study made only minor modifications to reflect current sizing assumptions. A schematic of the general layout is shown in Figure 58.

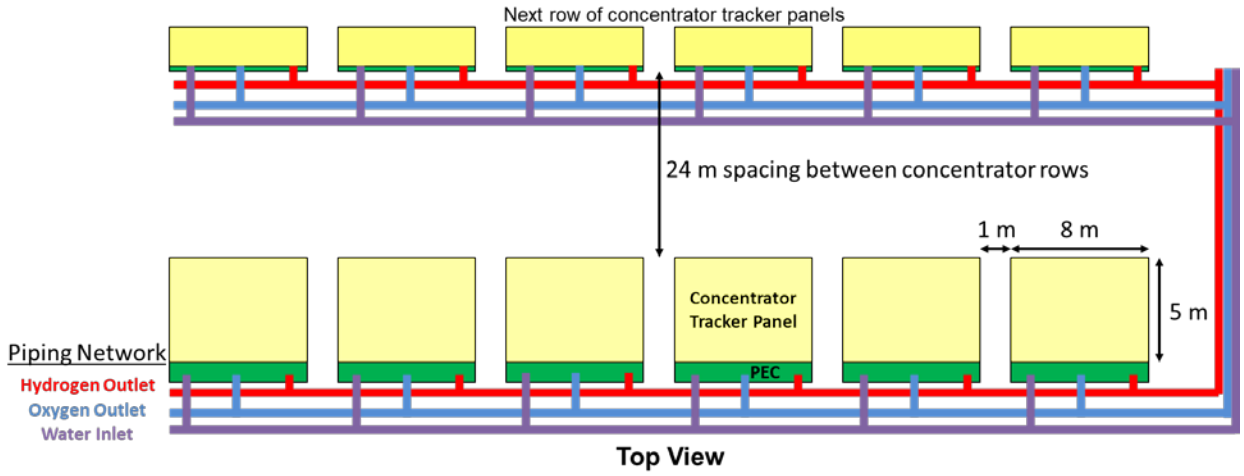


Figure 58 - General layout specifications for PEC Type IV

The concentrator tracker panel size was set to 8 m by 5 m, which is consistent with the size currently achievable by parabolic troughs. The distance between rows was linearly extrapolated from the results calculated in the 2009 PEC report. A 1 m distance was enforced between concentrator tracker panels to allow for 2-axis tracking and maintenance support.

Three piping networks are deemed sufficient to service the PEC Type IV receiver panels. PVC piping is assumed to be acceptable for this application, with one piping network providing clean water to the receiver panels and two piping networks to return hydrogen and oxygen products.

8.12 PEC Type IV Case Study

8.12.1 General PEC Type IV Assumptions

A modular PEC design is envisioned for which each module has a capacity of 1,000 kg H_2/day and multiple modules are strung together to reach the desired H_2 production rate. Key parameters are listed in Table 27.

Table 27 - PEC Type IV Design Specifications

System Parameters	Units	Value
PEC Type	-	Type IV

[44] James, Brian D., George N. Baum, Julie Perez, and Kevin N. Baum. "Technoeconomic Analysis of Photoelectrochemical (PEC) Hydrogen Production," December 1, 2009. https://doi.org/10.2172/1218403.

Average Insolation	kWh m^{-2} day^{-1}	7.46
STH Efficiency	%	35%
Overall Solar Efficiency	%	33.3%
Collector Efficiency	%	95%
Average H_2 Mass Flow	kg day^{-1}	1,000
Area Specific Mass Flow	kg H_2 hr^{-1} m^{-2}	3.33E-03
Total Area Collector Required	m^2	13,200
Collector Length	m	8
Collector Width	m	5
Collector Area	m^2	40
Number of Collectors	#	330
PV Area Required	m^2	440

8.12.2 Labor Assumptions

In addition to the capital cost, we reviewed the labor costs associated with the PEC IV concept. The labor rate was updated to scale similarly to the maintenance of a solar panel facility. As shown below (Equation (**5**)), this equates to 1 field worker tending to the maintenance of 62.5 acres of PEC field (solar capture area) plus the workers in the control room. While not verified, we judge that PEC fields, which include H_2O/H_2/electrical lines, will have higher maintenance cost than PV fields, which include only electrical lines. PEC maintenance labor estimates require further study and NREL has been asked to explore whether a more accurate labor rate could be developed from recent solar installation estimates or from analogous bioreactor installations.

$$Module\ FTE\ = \frac{0.016\ FTE}{1000\ m^2}(Total\ Land\ Required) + \frac{3\ FTE}{50\ TPD}(Total\ Facility\ H2\ Production) \quad (5)$$

8.12.3 Solar Concentrator Cost

Solar concentrators often represent about a quarter of an entire concentrating solar-thermal power (CSP) system cost and reducing the capital investment for these systems is of high importance. We assume in this study that advancements in solar concentrators for CSP electricity generation systems will directly improve CSP PEC H_2 generation applications as well. The 2015 NREL report on parabolic trough solar concentrators includes estimates between \$170 and \$180/m^2 for two different designs (from companies

SkyFuel and FLABEG).[45] In 2018, SkyFuel claimed achievement of \$100/m^2 using a parabolic trough CSP,[46] close to reaching the DOE's SunShot target for CSP (\$75/m^2).[47] For the purposes of forecasting future possible levelized costs of hydrogen, the DOE SunShot target cost was selected (\$75/m^2). A future study can explore a sensitivity to this cost and whether a higher-cost parabolic trough solar concentrator would still be feasible.

8.12.4 Product Specifications

Similar to the 2009 PEC report,[48] PEC Type IV is assumed to use 1 condenser and 2 intercoolers to achieve product gas output of 99.6% H_2 and 0.4% water vapor. The hydrogen is cooled to 40°C using cooling water to achieve this product purity. Since the water is already pressurized, no hydrogen compressor is assumed to be necessary. This is consistent with the hydrogen purity required for industrial applications but may be insufficient for transportation applications. Further cost analysis for hydrogen compression and dehydration methods should be explored to understand how the levelized cost of hydrogen changes.

8.12.5 H2A Assumptions

The standard H2A workbook has continued to evolve since the 2009 PEC report.[39] Where appropriate, the changes implemented in this cost study are shown in Table 28, Table 29, and Table 30.

Table 28 - PEC Type IV H2A Financial Parameters

Financial Parameters	2009 Analysis Value	2021 Analysis Value
Operating Period	20 years	
Facility Life	20 years	
Construction Period and Cash Flow	1 year	20% Year 1, 80% Year 2
Installation Cost Factor	1.3	1.4
Land Cost	\$500/acre	\$5,000/acre
Property Taxes and Business Insurance	2%/year of the total initial capital cost	
IRR	10% after tax	8% after tax

[45] Kurup, P., Turchi, C., "Parabolic Trough Collector Cost Update for the System Advisor Model (SAM)", NREL report, November 2015. https://www.nrel.gov/docs/fy16osti/65228.pdf

[46] Schuknecht, N., McDaniel, J., Filas, H., "Achievement of the \$100/m2 Parabolic Trough", SkyFuel publication in the AIP 2018 Conference Proceedings, November 2018.

[47] https://www.nrel.gov/docs/fy12osti/55451.pdf

[48] James, Brian D., George N. Baum, Julie Perez, and Kevin N. Baum. "Technoeconomic Analysis of Photoelectrochemical (PEC) Hydrogen Production," December 1, 2009. https://doi.org/10.2172/1218403.

Working Capital Rate	15% of the annual change in total operating costs	
Income Taxes	35% Federal, 6% State, 38.9% effective	21% Federal, 6% State, 25.74% Effective
Sales Tax	Not included - facilities and related purchases are wholesale and through a general contractor entity	
Decommissioning	10% of initial capital	
Salvage Value	10% of initial capital	

Table 29 - PEC Type IV H2A Operating Parameters

Operating Parameters	2009 and 2021 Analysis Value
Hydrogen Pressure at Central Gate	300 psig
Hydrogen Purity	98% minimum; CO < 10 ppm, sulfur < 10 ppm [99.6% H_2, Industrial purity assumed]
Burdened Labor Rate for Staff	$50/hour
G&A Rate	20% of the staff labor costs

Table 30 - PEC Type IV H2A System Parameters

System Parameters	2009 Analysis Value	2021 Analysis Value
Operating Capacity Factor	90%	
Site Preparation	1% of direct costs minus unique excavation costs	2% (No unique excavation)
Engineering & Design	7% of direct costs (compared to default of 13% due to modularity)	7% of direct costs
Process Contingency	20% of direct costs (compared to default of 15% due to greater uncertainties in system configuration)	20% of direct costs

Project Contingency	$0	$0
Up-Front Permitting Costs	0.5% of direct capital costs (compared to default of 9% due to modularity of design and environmental benefits)	5% of direct capital costs
Annual Maintenance & Repairs	0.5% of direct capital costs	0.5% of direct capital costs

8.12.6 PEC Type IV Case Study Results

At the time the PEC study was performed, the target levelized cost for green hydrogen was \$2/kg H_2 by 2030. A set of possible conditions were selected as a case study for what future PEC Type IV facility would cost: 35% STH, 1 year catalyst lifetime, \$240/m^2 electrode cost, and 30:1 concentration ratio. While the theoretical maximum multi-layer STH conversion efficiency for photoelectrolysis exceeds 40%, PEC systems to date have only demonstrated 8%-12.4% efficiencies and future efficiencies are projected to reach above 30%. Therefore, this case study is highly aspirational in concept.

The plexiglass and casing materials were assumed to be replaced every 10 years and a 50% manufacturer markup was included in the price of the electrode, plexiglass, and casing. A bill of materials for this 1 TPD module is shown in Table 17.

Table 31 - PEC Type IV Case Study Bill of Materials (1 TPD H_2 Modules)

Component	Units	Cost Unit	Cost per Unit	Total Cost
Concentrator Tracker	13200	\$/m^2	\$75.00	\$990,000
PEC Cells	440	\$/m^2	\$240.00	\$105,600
PEC Panels	440	\$/m^2	\$175.91	\$77,399
Port Hardware	990	\$/port	\$4.50	\$4,455
Port Installation	990	\$/port	\$4.50	\$4,455
Piping	9628	\$/m^2	\$1.71	\$16,463
PEC Wiring Panel	330	\$/panel	\$117.00	\$38,610
Water Level Controllers	330	\$/panel	\$40.00	\$13,200
Circulation Pump	33	\$/pump	\$500.00	\$16,500
Pressure Sensors	33	\$/block	\$345.00	\$11,385
Hydrogen Area Sensors	33	\$/block	\$299.00	\$9,867

Instrument Wiring	33	$/block	$77.00	$2,541
Power Wiring	33	$/block	$39.00	$1,287
Conduit	33	$/block	$103.00	$3,399
H2 Compressor	1	$/module	$0.00	$0
HX - Condenser	1	$/module	$7,098.00	$7,098
Makeup Water Pump	1	$/module	$213.00	$213
Computer & Monitor	1	$/module	$1,500.00	$1,500
Control Software	1	$/module	$4,300.00	$4,300
PLC	1	$/module	$2,000.00	$2,000
Control Room Building	1	$/module	$8,000.00	$8,000
Control Room Wiring Panel	1	$/module	$3,000.00	$3,000
Hydrogen Flow Meter	1	$/module	$5,500.00	$5,500

The uninstalled capital cost is $1.33M and the levelized cost of hydrogen is $1.84/kg H_2.

8.13 PEC Type IV Sensitivity Study

SA performed a sensitivity study on several PEC Type IV input parameters. These include electrode cost, active material lifetime, STH efficiency, and concentration ratio. Two scenarios were tested, corresponding to a worse case and best case scenario for PEC Type IV.

- Scenario 1 uses a low concentration ratio of 10, which is compatible with an unpressurized PEC receiver, and a high electrode cost of $915/$m^2$, which is likely achievable with today's manufacturing capabilities.
- Scenario 2 uses a maximum concentration ratio of 30, which is compatible with a system pressurized to 300 psi, and a low electrode cost of $240/$m^2$, which is consistent with a future high manufacturing rate of 100,000 m^2/year.

Results for PEC Type IV Scenarios 1 and 2 are shown in Figure 59 and Figure 60.

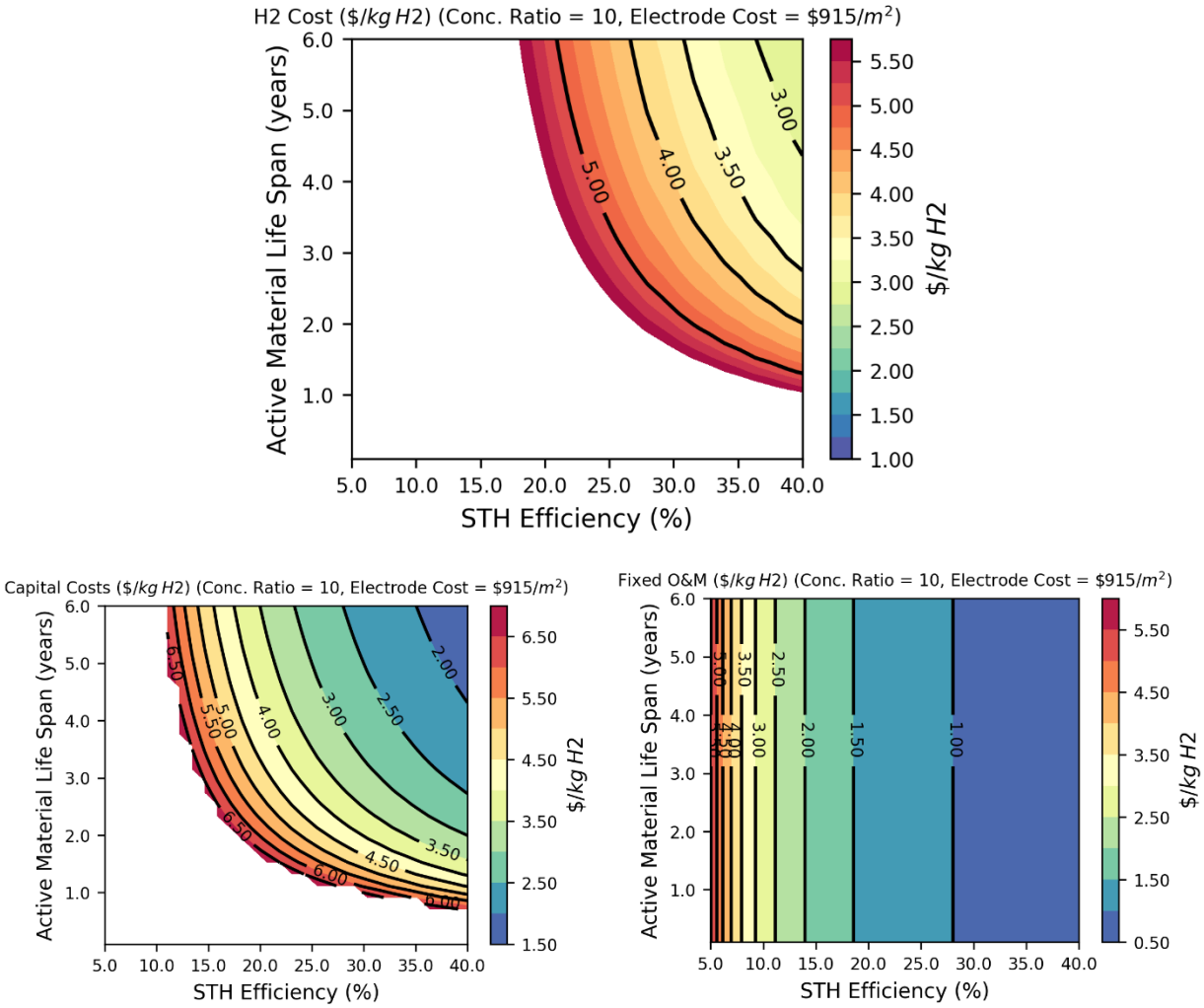


Figure 59 - PEC Type IV Scenario 1 with a concentration ratio of 10 and an electrode cost of $915/m^2. This scenario models a high electrode cost, low solar concentration ratio system that could be built today. Under these operating and cost conditions, it is not possible to achieve a $2/kg H_2.

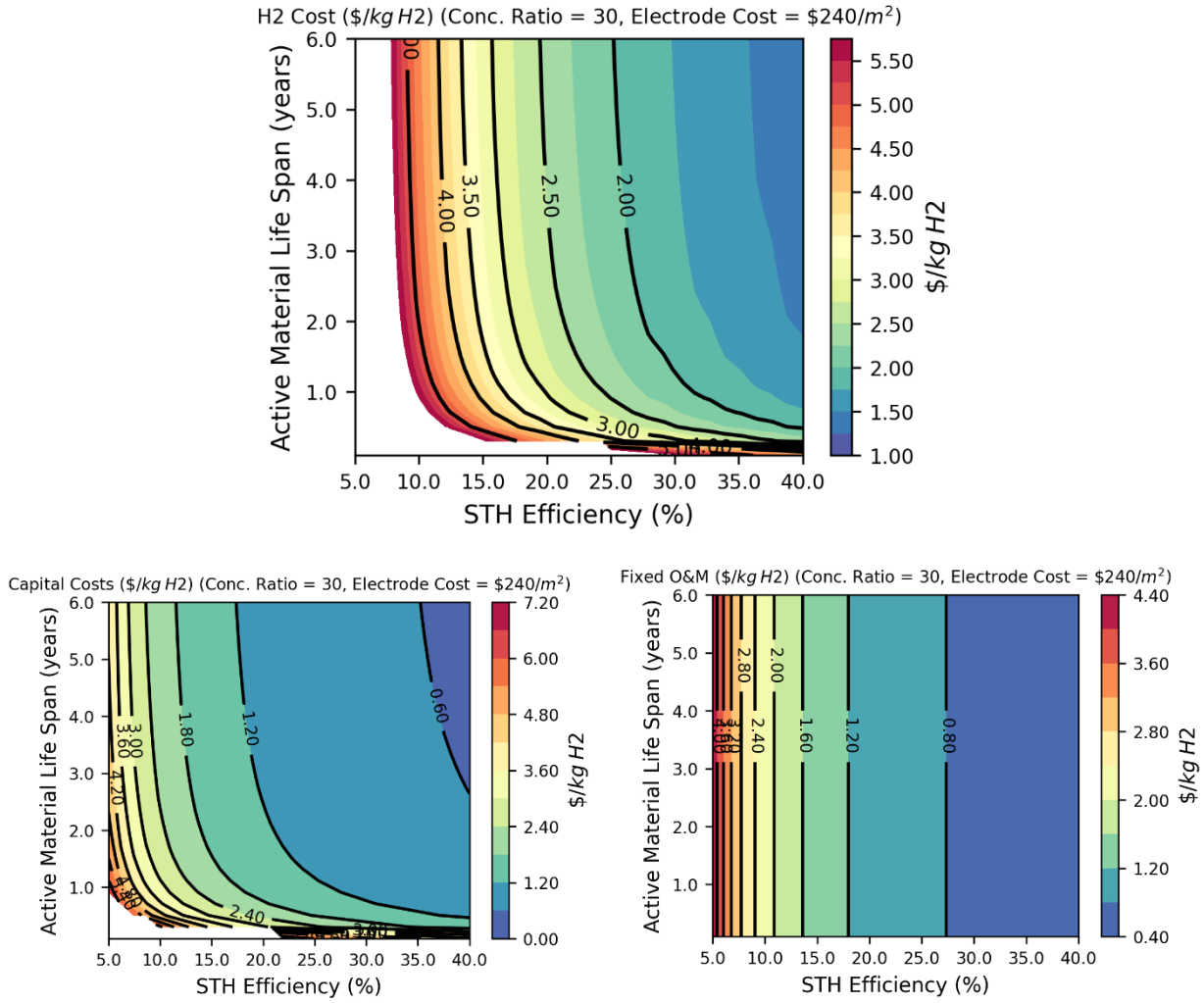


Figure 60 - PEC Type IV Scenario 2 with a concentration ratio of 30 and an electrode cost of $240/m^2. The scenario models a future low electrode cost, maximum concentration ratio case that may be possible in the future. There is a broad operating space to achieve a levelized hydrogen cost of $2/kg H_2.

The results for the scenario in Figure 59 suggest that a H2 cost of $2/kg is not currently achievable. However, the future scenario modeled in Figure 60 shows that with an STH Efficiency of >25% and a catalyst lifetime of >1 year, it may be possible to achieve the H_2 cost target of $2/kg.

9 Solar Thermochemical Hydrogen (STCH) Production

STCH production is a promising H_2 production technology that uses high-temperature heat from concentrated solar power to drive a series of chemical reactions to split water. Building on previous work, NREL examined the cost to produce H_2 by STCH. The results of the study are documented in NREL's report titled "Solar Thermochemical Hydrogen Production Supporting Boundary Study." The general chemical reaction pathway for STCH production is shown in Figure 61. NREL also conceptualized an actual production facility for STCH, originally rated for 100 tpd of H_2 (see Figure 62). This design is the basis for the TEA conducted by NREL with the assistance of SA.

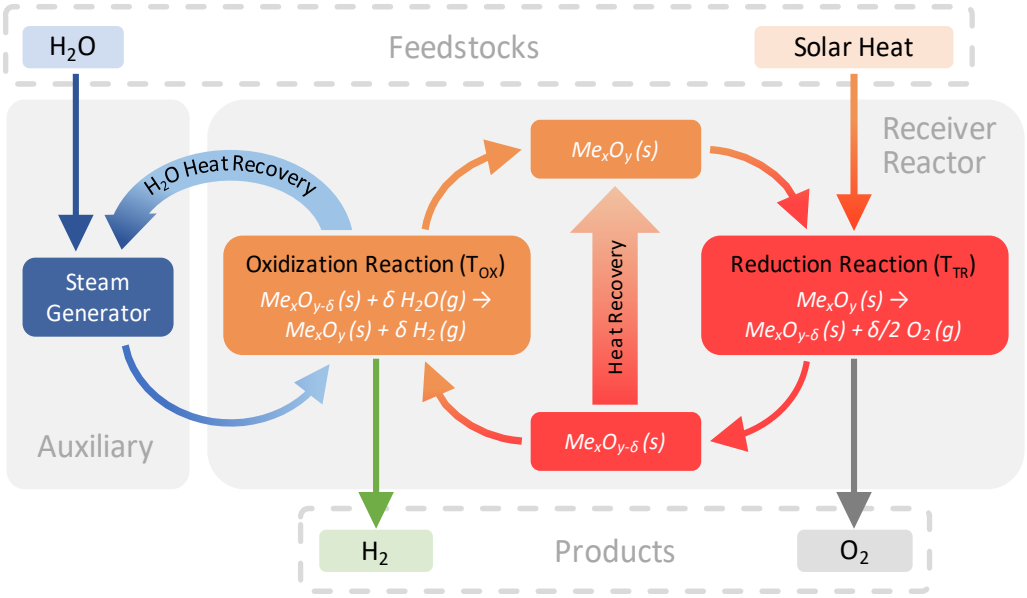


Figure 61 - General chemical reaction pathway for STCH

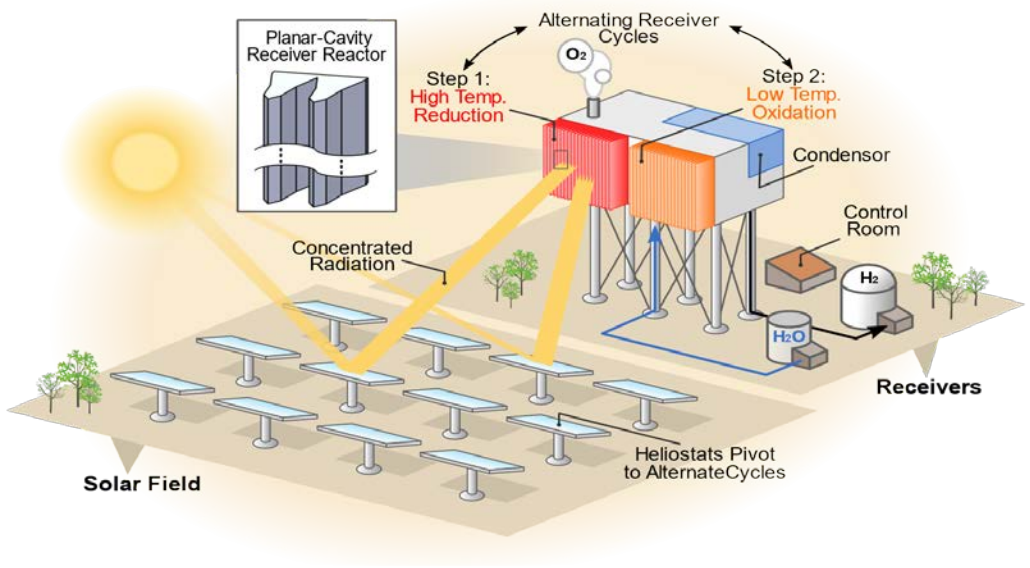


Figure 62 - Conceptualized production site for STCH

The system design parameters for STCH are shown in Table 32. The parameters are scaled from the original 100 tpd plant size conceptualized by NREL to a 14.1 tpd plant.

Table 32 - System design parameters for STCH H2A analysis

Parameter	Value	Notes
Daily Field Production Target	100 Ton H_2/day	DOE Target
Thermal Power to a Modular Plant	200 MWt	Peak power delivered to receiver aperture
Module Daily H_2 Production	14.9 Ton H_2/day	Assumed 21% STH efficiency, 90% capacity factor
Number of Modules in Field	7	Each 200 MWt plant produces 14.1 Ton H_2/day
Field size	~585 m radius	Determined by SolarPILOT model
Number of Heliostats per Plant	17,128	Heliostat size 4.25 m x 4.25 m
Tower height	130 m	Result of parametric Solar PILOT optimization
Tower cost	$4,371,475	Determined from literature as C=4.93*(130*h)^2.158 with 'h' in feet, then adjusted for inflation using CEPCI.
Annual solar field optical eff.	53%	Determined by Solar PILOT model
Solar Receiver Thermal Efficiency	80%	Referred to design at 900°C, but further optimization is possible to minimize thermal losses
Thermochemical efficiency	40 < 50 < 60%	Battleson, K. W., 1981, "Solar Power Design Guide: Solar Thermal Central Receiver Power System, A Source of Electricity and/or Process Heat," (April). Values correspond to Low, Baseline, High.
STH efficiency	17 < 21 < 25%	LHV basis Values correspond to Low, Baseline, High.
Electricity price (c/kWh)	7.76 ¢/kWh	Supplied by PV electric generation.
Annual Water Utility Usage	430.7 million gal	Assumed 11.8 gals/kgH_2

The calculated H_2 production cost using the H2A tool with the TEA inputs is $2.54/kg H_2. Figure 63 shows a sensitivity study for various parameters affecting H_2 production cost for STCH production. This study shows that the active material productivity (µmol H_2/g) is the key cost driver.

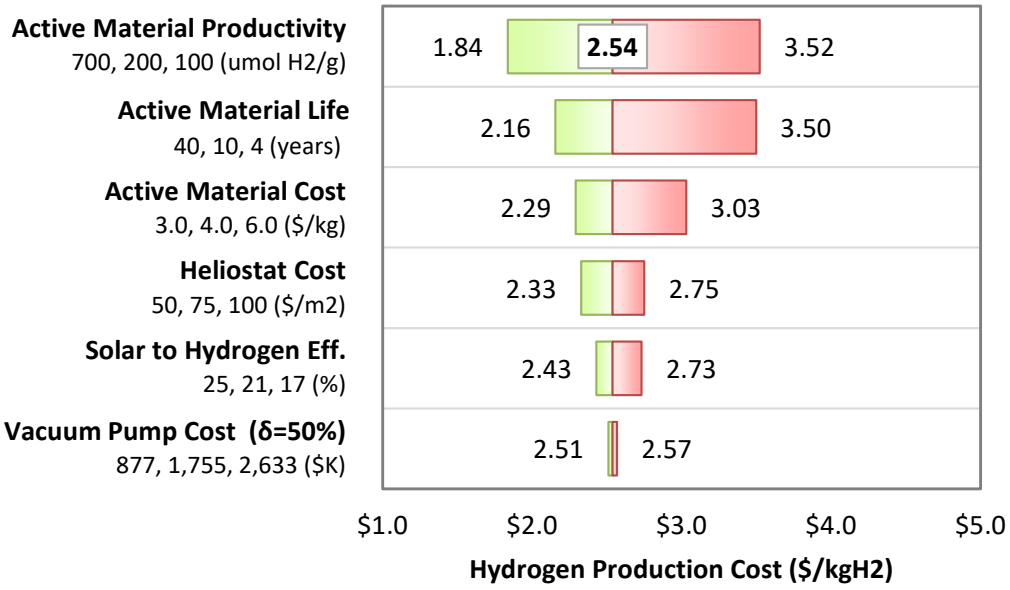


Figure 63 - Sensitivity study for various parameters affecting H_2 production cost for STCH production

10 Energy Transmission Cost Study

As public governments and industries strive to increase use of renewable energy resources, such as wind and solar, challenges related to storage and distribution of this intermittent energy are coming to the forefront. It is generally assumed that the electrical grid will be the primary means of transmitting renewable energy to consumers. However, the cost of electricity transmission can be high. SA conducted an energy transmission cost study, which is documented in a report titled "Relative Cost of Long-Distance Energy Transmission by Electricity vs. Gaseous and Liquid Fuels" and in a companion Journal article.[49] This study compares the cost of long-distance, bulk transport of electrical and chemical energy independent of production method or end-use.

Three energy transmission methods were analyzed: Electrical Transmission Lines, Liquid Pipelines, and Gas Pipelines. The capital cost was estimated based on existing cost models but normalized to our specifications. The cost was compared for 1,000 miles of transmission and on an even basis. The data is presented as $/mile (traditional) as well as $/(mile-MW) and $/MWh. The models included capital expense (CapEx) for materials, labor, Right-Of-Way, pumping/compression stations, and miscellaneous expenses. We developed total costs for transmitting energy. Capital cost is sometimes erroneously presented as a transmission cost. A few studies suggest that a set percentage of the total transmission cost is the capital cost. Transmission cost should include capital cost and operating cost. We included costs of pumping and compressor stations for pipelines as well as transmission line losses for electrical lines. Costs for electricity production, fuel production, and fuel conversion are not included.

For Electrical Transmission Lines, the modeled parameters include aluminum core steel reinforced lines on a new lattice structure. 500 kV high-voltage direct current (HVDC) lines are modeled with 2 substation locations. Terrain estimates are broken up evenly between 8 types, ranging from flat ground to wetland and mountain terrain. Similarly, Right-Of-Way costs are broken up into 12 zones, evenly distributed among each zone for a representative model. Capital costs and resistive losses are based on *Capital Costs for Transmission and Substations* (2014). For Gas Pipelines and Liquid Pipelines, pipeline cost models are taken from literature (Rui et al) and pipeline models are derived from *Oil and Gas Journal* data. Data is for onshore, natural gas pipelines from 1992-2008. No reliable cost data was found for liquid pipelines. Following common practice, the same cost models used for gas pipelines were also used for liquid pipelines. Models were optimized for lowest cost (by selecting optimal pumping station spacing). Capital costs and operating power requirements were considered. The Power requirements were costed at ¢5/kWh.

In most studies, cost metrics compare electrical and pipeline cost on $/mile basis. This does not account for capacity. It only represents capital cost and usually shows electrical and pipeline capital cost on a similar order of magnitude. By comparing the capital cost on a $/(mile-MW) basis, the capacity of the transmission method is included. Amortizing the capital cost with an annual operation cost allows for a

[49] "Cost of long-distance energy transmission by different carriers", Daniel DeSantis, Brian D. James, Cassidy Houchins, Genevieve Saur, Maxim Lyubovsky, iScience, Volume 24, Issue 12, December 19, 2021. https://doi.org/10.1016/j.isci.2021.103495

comparison of total transmission cost in $/MWh. Thus, we compare transmission methods on $/mile, $/(mile-MW), and $/MWh bases. In the results, costs for transmission methods are usually broken down to $/mile. It is more useful to consider the cost per distance per capacity. Production method and conversion costs are not considered in this analysis. Inclusion of those costs could change relative ranking of the options.

Table 33 shows the system parameters for each transmission method. Table 34 shows the cost parameters for the energy transmission calculations. Table 35 shows the results for each transmission method.

Table 33 - System parameters for each transmission method

Transmission Method	Liquid Pipeline			Gas Pipeline	
Energy Carrier	Crude Oil	Methanol	Ethanol	Nat Gas	Hydrogen
Pipe diameter (in)	36	36	36	36	36
Flow velocity (m/s)	3.7	3.9	3.9	18	18
Pressure Drop (bar/mile)	2.5	2.5	2.5	0.67	0.19
Pump/compressor load (MW/station)	29	30	30	39	18
Pipeline Operating Power (MW/1000 mi)	715	757	758	464	162

Table 34 - Cost parameters for the energy transmission calculations

Interest (Discount) Rate	Misc. Costs per year	Maintenance Costs per Year	Operating Expenses	Corporate Tax Rate	Capital Recovery Factor	Equipment Lifetime (Amort. Period)
8%	5% of CapEx	5% of CapEx	0.5% of Pipeline Cost	26.6%	~12%	Pipelines: 33 yrs Elect. Line: 60 yrs

Table 35 - Results for each transmission method

Transmission Method	Electrical	Liquid Pipeline			Gas Pipeline	
Energy Carrier	HVDC	Crude Oil	Methanol	Ethanol	Nat Gas	Hydrogen
Flow (amps, kg/s)	6,000	1,969	1,863	1,859	368.9	69.54
Rated Capacity (MW)	2,656	91,941	37,435	50,116	17,391	8,360
Capital Cost ($M/mile)	$3.9M	$1.47M	$1.92M	$1.92M	$1.69M	$1.38M
Operating Power: Rated Capacity	12.9%	0.78%	2.02%	1.51%	2.67%	1.94%
Capital Cost ($/(mile-MW))	$1,467	$16	$51	$38	$97	$166
Transmission Cost ($/MWh)	$41.50	$0.77	$2.2	$1.7	$3.7	$5.0

Figure 64 shows the fully amortized energy transmission costs ($/MWh per 1,000 miles) over the lifetime of the transmission method. In all cases of energy transmission, pipelines prove to be the most cost-effective method to transmit energy across long distances. The results indicate that the cost of new electrical transmission per delivered MWh can be up to eight times higher than for H_2, about eleven times higher than for natural gas, and twenty to fifty times higher than for liquid fuels. These energy transmission cost differences are important but often not recognized, and ultimately must be considered in light of the entire energy production/transmission/distribution pathway cost.

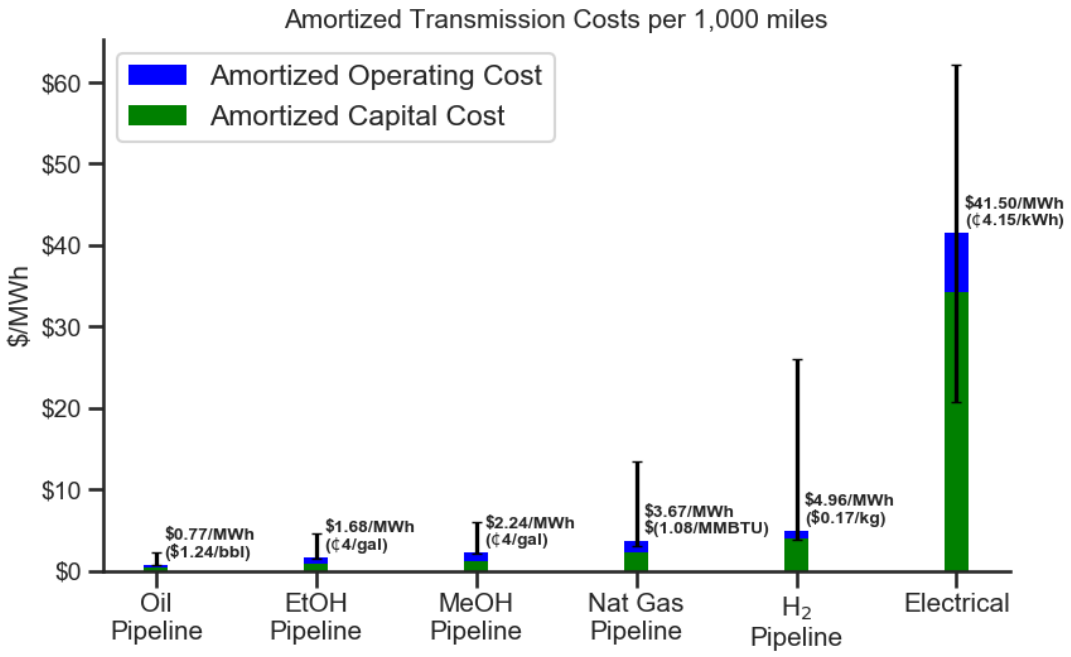


Figure 64 - Levelized cost of energy transmission over 1,000 miles in $/MWh of delivered energy. The labels over the bars show the cost of delivery in the price units specific for each energy carrier. Pipeline error bars represent uncertainty in the cost based on the construction location of the pipeline. Uncertainty in the electrical line is +/- 50% of the amortized energy transmission cost.

While the capital costs of construction of electrical transmission lines and the pipelines are about the same (between about $1.5M and $4M per mile (and may vary significantly depending on the particular project size, location, topography, financing options, etc.), the energy-carrying capacity of the electric wires is much lower than for gaseous and liquid pipelines. Multiple electrical transmission lines would have to be built to transport the equivalent amount of energy as a single high-capacity pipeline. Furthermore, operating energy losses are much greater in electrical transmission lines than in chemical fuel transportation. These two factors make electrical transmission the most expensive transmission method among the studied energy carriers.

This work suggests that renewable energy transmission via gaseous or liquid carriers is a more efficient and less expensive method for energy transmission compared to electricity. The cost of wind or solar conversion into electricity or chemical carriers is not considered in this study. While the first is well defined by wind turbines and solar photovoltaics practiced on a large scale, the technologies for renewables conversion into gas and liquids, e.g., through water electrolysis and gas-to-liquid processes, are not as mature and the cost is less certain. Deployment of technologies for the conversion of renewable fuel on a large scale should bring the cost down in the same way that costs decreased for wind turbines and solar photovoltaics, potentially making the renewable energy pathway cost-competitive with fossil fuels. Combined with much lower transmission cost, this will benefit overall utilization of renewable energy. Renewable gaseous and liquid fuels also provide an advantage over electrical transmission in that they allow for long-term, cost-effective, and large-scale energy storage. Such storage may solve the renewable energy curtailment problem characteristic for renewable electricity.

11 H_2 Price For Competitive LCOE

SA conducted a study on the H_2 price for a competitive Levelized Cost Of Electricity (LCOE). To determine the price of H_2 needed to achieve a competitive cost for electricity produced from H_2, a simplified LCOE model was created and applied to five electricity production cases:

- Natural Gas Combined Cycle (NGCC): the baseline scenario to establish a market competitive electricity price
- NGCC with Carbon Capture and Sequestration (NGCC-CCS): the above baseline case but with capture of the flue gas carbon emissions
- Advanced Gaseous Combined Cycle (AGCC): similar to the NGCC but operating with H_2 fuel
- Solid Oxide Fuel Cell (SOFC): modeled as a SOFC operating on H_2 fuel
- SOFC Combined Cycle (SOFC-CC): modeled as a SOFC operating on H_2 fuel with a bottoming cycle for increased efficiency

Nominal values for capital costs are estimated for each case, but values are also varied parametrically to illustrate cost sensitivity. System fuel-to-electricity plant efficiencies are estimated for each production pathway. All four production options have an LCOE equal to the NGCC reference value from the simplified cost model. This was achieved by varying the H_2 fuel price so that the target LCOE was attained. The result is the determination of the H_2 price needed in each case to achieve a market competitive LCOE (i.e., an LCOE equal to the NGCC LCOE).

National Energy Technology Laboratory (NETL) conducted a study in 2015 to determine the construction and operating costs of various fossil fuel power plants.[50] This study included cost and performance parameters and an LCOE analysis for two power plants: an NGCC with CCS and an NGCC without CCS. SA created a simplified cost model from the data provided in the NETL report and is able to duplicate the NETL LCOE results to an acceptable level (<2.5% variation in either case study). The input assumptions and results of SA's simplified cost model for the NETL NGCC case studies (production only LCOE) are shown in Table 36.

[50] T. Fout, "Cost and Performance Baseline for Fossil Energy Plants," NETL, DOE/NETL-2015/1723, Jul. 2015.

Table 36 - SA's simplified cost model for the NETL NGCC case studies (production only LCOE)

NGCC Power Plant Cost Analysis from NETL Report		NGCC without CCS	NGCC with CCS
NETL NGCC Power	MW_{net}	630	559
NETL Installed Construction Cost	$2016/kW	$955	$2,062
Total NETL Installed Construction Cost	2016$	$601,650,000	$1,152,660,822.92
Annual Payment	$/yr	$67,789,150	$129,872,678
NG-Turbine Efficiency	% LHV	38.10%	38.10%
Bottoming Cycle Efficiency	%LHV	39.10%	43.50%
NGCC Efficiency	%LHV	57%	51%
NETL Natural Gas Feed Flow (includes eff)	kg/hr	84,134	84,134
NETL NG Cost	$/mmBTU	$6.50	$6.50
NETL NG Cost	$/kg	$0.26	$0.26
Annual NG Cost	$/yr	$191,623,598.40	$191,623,598.40
Annual Operating Labor	$/kW/yr	3.81	5.43
Maintenance Labor	$/kW/yr	5.98	10.97
Admin & Support Labor	$/kW/yr	2.45	4.10
Taxes and Insurance	$/kW/yr	14.51	31.40
Total	$/kW/yr	26.74	51.90
Total Annual Fixed O&M	$/yr	$16,848,203	$29,010,121
Maintenance Materials	$/MWh	1.20	2.20
Variable Operating Costs	$/MWh	1.76	4.20
Total Variable Cost per year	$/yr	$8,273,618	$17,488,279
Total Maintenance Materials Cost per year	$/yr	$5,650,126	$9,176,493
Total Annual Payments	$/yr	$290,184,695	$377,171,170
Annual Electricity Generation	kWh/yr	4,690,980,000	4,162,314,000
NETL Report LCOE	$/kWh	$0.0611	$0.0925

The simplified cost model uses a scalable capital cost to determine the total construction cost of a theoretical power plant. Users supply both the desired capital cost ($/kW) for the plant construction cost and the desired plant power (MW), allowing for parametric studies on the effect of capital cost vs. the cost of fuel. Estimated system efficiencies[51] are used to calculate annual fuel consumption, based on a user-supplied power input to the cost model. The annual cost of fuel is calculated from a user-defined fuel price multiplied by the fuel required for plant operation. Fixed operating and maintenance costs are based on an annual percentage of capital cost, based on the type of power generation equipment selected. Annual payments to repay plant capital cost are assessed via a capital recovery factor of 0.113, based on an after-tax return of 11%[52] and a 40-year plant lifetime. Annual expenses are totaled and divided by the annual net electrical energy produced to determine the LCOE for the production of electricity.

The simplified LCOE model is used to determine the H_2 price ($/kg) value which achieves an LCOE equivalent to the baseline natural gas plants (NGCC with CCS and NGCC without CCS).[53] This final fuel price establishes a target H_2 price that would allow for competitive electricity generation.

[51] All efficiencies in this report are defined as net system efficiency, $\frac{Net\ Electricity\ Out}{Fuel\ LHV}$.

[52] NETL reports a tax rate of 38%.

[53] NETL used a baseline natural gas price of 2016$6.86/GJ (2016$6.50/mmBTU). That value is maintained here in the analysis although recent 2018 natural gas prices have been as low as 2016$2.82/GJ (2016$2.67/mmBTU).

NGCC power plants have high efficiencies (modeled as 57%) and low capital cost (modeled as $955/kW). This combination makes the production of electricity fairly inexpensive from NGCC plants. To match this low electricity cost with H_2-fueled power plants, the cost of H_2 must be low. Figure 65 and Figure 66 show the H_2 cost required to achieve parity with the NETL case study for NGCC without CCS and NGCC with CCS, respectively. The lines are representative of the parametric study conducted in which the H_2 price is found based on the total installed cost of the power plant in question. The data points on each line represent the nominal capital cost for a prospective plant, and thus also indicate the most likely cost of H_2 for that technology.

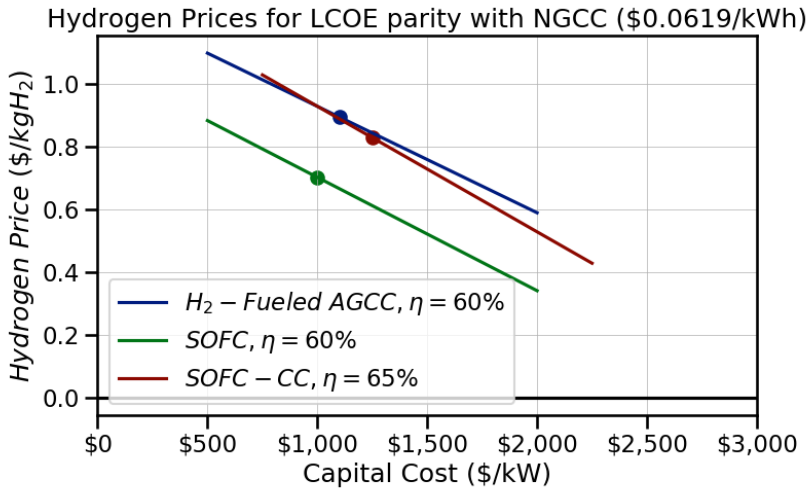


Figure 65 - H_2 cost to achieve LCOE parity with the NETL case study for NGCC without CCS. The point on each line marks the most likely cost-point for each analyzed technology.

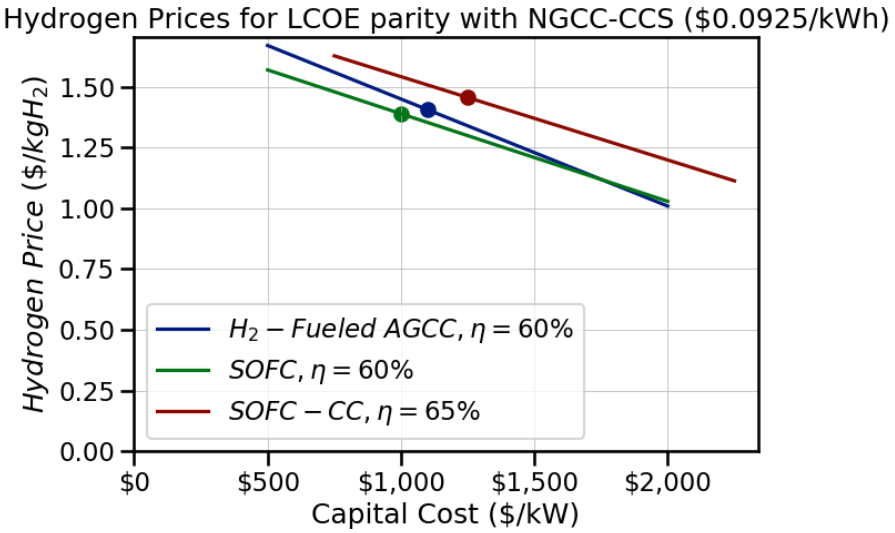


Figure 66 - H_2 cost to achieve LCOE parity with the NETL case study for NGCC with CCS. The point on each line marks the most likely cost-point for each analyzed technology.

The results show that the baseline electricity LCOE (production only) is $0.0619/kWh for NGCC without CCS and $0.0925/kWh for NGCC with CCS, in $2016. The NGCC plant has a very good efficiency (57% electrical) due to the use of a bottoming cycle and a low capital cost ($955/kW). The H_2-fueled AGCC plant has a slightly higher efficiency (60%) and a slightly higher capital cost (estimated at $1,100/kW though a strong reference value is not currently available). The H_2-fueled SOFC plant has a nearly equivalent electrical efficiency (60%) and cost ($1,000/kW) to those of the H_2-fueled AGCC plant. Both SOFC values are set at DOE Target levels.

The H_2 production only price (at the projected nominal capital cost) for each technology to achieve LCOE parity with the baseline NETL LCOE for both NGCC without CCS and NGCC with CCS is shown in Table 37. These H_2 prices are lower than the DOE targets for alternative fuel vehicles and will be a challenge to achieve. The initial cost analysis suggests that H_2 prices must be less than $1.50/kg H_2 to achieve a competitive LCOE when compared to NGCC with CCS. For NGCC without CCS, the H_2 price to achieve a comparative LCOE drops below $1.00/kg H_2. The H_2-fueled AGCC is observed to be similar to the fuel cell systems in efficiency and capital cost, and thus requires a similar H_2 cost. However, the H_2-fueled AGCC is substantially lower in maintenance costs (due to no stack replacement costs) and requires a large power plant size (>500 MW) to achieve its high efficiency.

Table 37 - H_2 price required to achieve LCOE parity with NGCC power plants. Costs for H_2-fueled plants are based on nominal capital cost estimates for each technology.

	LCOE parity with NGCC without CCS ($0.0619/kWh$_{electric}$)	**LCOE parity with NGCC with CCS ($0.0925/kWh$_{electric}$)**
H_2-fueled AGCC	$0.89/kg H_2	$1.41/kg H_2
H_2-fueled SOFC	$0.70/kg H_2	$1.39/kg H_2
H_2-fueled SOFC-CC	$0.83/kg H_2	$1.46/kg H_2

12 Cost Results Comparison

Comparisons between the cost results for PEM Electrolysis, Solid Oxide Electrolysis, AEM Electrolysis, PEC Production, and STCH Production are shown in Table 38 and Table 39. Of the currently commercialized electrolysis pathways, SOE appears to have the lowest potential cost at $4/kg. AEM electrolysis is not yet complete, however, after initial review, there is reason to suspect a potentially lower system cost than PEM or SOE electrolysis systems. The water-splitting methods (PEC Type II, PEC Type IV, and STCH) have a lower technology readiness level and have yet to be validated in an industrial setting.

Note that the H_2 cost results from all of the electrolysis systems are highly dependent on the assumed price of electricity. The DOE and industry are actively pursuing methods and operating scenarios to allow <$0.03/kWh electricity which would reduce the $/kg$H_2$ costs dramatically compared to the values shown. Additionally, technology advancement continues for all systems and the performance estimates for future systems (particularly for the emerging technology PEC and STCH systems) have a considerable degree of uncertainty regardless of our attempts to select defensible parameter values. Consequently, this cost comparison should be viewed as informational to guide R&D decisions but not to make definitive assessments of a concept's potential worth.

Table 38 - Comparison between the cost results for Electrolysis Methods: PEM, SOE and AEM

Technology	**PEM Electrolysis**		**Solid Oxide Electrolysis**		**AEM Electrolysis**	
Case	**Current**	**Future**	**Current**	**Future**	**Current**	**Future**
TRL	9	7	7	5	7	5
System Cost (2016$/kW) **[Fabrication Rate, System Size]**	$460 [10 MW/yr, 50 MT H_2/day]	$233 [2,000 MW/yr, 50 MT H_2/day]	$522 [700 MW/yr, 50 MT H_2/day]]	$357 [700 MW/yr, 50 MT H_2/day]	Not Yet Available	
H_2 Production Cost (2016$/kg H_2)	$4.96 [Base, Central]	$4.48 [Base, Central]	$4.16 (Base, Central)	$3.89 [Base, Central]	Not Yet Available	

Table 39 - Comparison between the cost results for Water Splitting Methods: PEC and STCH

Technology	**PEC Type 2 Water Splitting**		**PEC Type 4 Water Splitting**		**STCH Water Splitting**	
Case	**Current**	**Future**	**Current**	**Future**	**Current**	**Future**
TRL	4	4	4	2	4	2
H_2 Production Cost ($/kg H_2)	>$5	<$2	>$6	<$2	$2.54	

13 Research Outlook, Cost Challenges, and Recommendations

13.1 Proton Exchange Membrane Electrolysis

The PEM electrolysis hydrogen production method is mainly limited by the efficiency and cost of electricity to bring down the cost of hydrogen. Initial capital cost of the stack is driven by the relatively recent spike in Iridium pricing and precious metal-coated Titanium separator plates. As in all of these hydrogen production technologies, the hydrogen compressor often dominates the BoP component cost. A key research challenge is the ability to reduce precious metal loadings while maintaining performance and durability.

13.2 Solid Oxide Electrolysis

SOE will be a technology competitor to PEM and alkaline electrolysis in the future, however, currently low manufacturing volumes and high capital cost will prevent SOE from dominating the market in the near term. SOE has superior performance and electrical efficiency than PEM that helps drive down the cost of hydrogen at the same price of electricity. There are challenges to ramping up in manufacturing production to bring down the capital cost through economies of scale. The ability to operate lower in temperature will be beneficial to the cost of materials but also can negatively impact performance. Striking that balance will be important from a capital cost point of view, however, the ability to improve efficiency may have a greater benefit to the cost of hydrogen. While stack cost reduction is desired, the SOE BOP costs represent a large fraction of total system cost. BOP cost reduction efforts perhaps can best be achieved by system simplification and operating condition optimization.

13.3 Anion Exchange Membrane Electrolysis

AEM electrolyzer systems are promising for their potential of using non-PGM catalysts and coatings, low membrane cost, and stainless steel and nickel components (i.e., titanium is not required). AEM is currently at a lower TRL than PEM and thus will benefit from further research to improve performance and durability. Most recent R&D effort has focused on pure water systems rather than liquid electrolyte (KOH) enhanced systems. However, both should be considered as AEM operation with KOH added to the water feed currently dramatically improves performance and durability, at seemingly a modest expense associate with the KOH. There is confidence from the AEM community that AEM could reach PEM performance and require lower capital cost in the future. A key challenge will be to improve durability to the level of PEM systems while limiting or removing precious metal catalysts and coatings.

13.4 Photoelectrochemical Hydrogen Production

PEC has multiple substantial challenges to overcome to be competitive with other hydrogen production technologies. For both PEC Type II and Type IV, the main deficiency is a lack of a set of PEC materials that can simultaneously achieve modest durability (>1 year) and modest STH performance (>5%). Current material systems can achieve one but not the other. Development and analysis of system concepts are useful for understanding how the PEC nanoparticles or electrodes will scale-up into complete, cost-effective systems. However, PEC material set identification remains the top R&D focus. Improvement in

efficiency will have a tremendous impact on initial capital cost as it impacts the total panel or bed area required.

13.4.1 PEC Type II (Nanoparticle Suspensions)

As a low TRL hydrogen production pathway, PEC Type II performance needs to be experimentally validated using the system geometry proposed in this study. In particular, the baggie-on-raceway design assumes achievement of rapid diffusion of ions between the hydrogen production and oxygen production regions, which is key to system effectiveness. Alternative designs will need to be considered if the basic geometry does not support efficient solar to hydrogen efficiency.

13.4.2 PEC Type IV (Panel Electrodes)

As a low TRL hydrogen production pathway, PEC Type IV receiver panel performance needs to be experimentally validated using the complete system geometry expected to be implemented in the field. Multiple receiver configurations and operational scenarios are possible, with the use of concentrated solar to reduce the required PEC panel area being a promising option. Since a higher temperature and pressure system will result from the proposed 30:1 concentration ratio, economic and feasible design of the receiver will need special attention. This is an engineering rather than scientific challenge but impacts system performance, durability, and basic economic feasibility. In addition, effective control systems will need to be developed to maintain appropriate liquid water levels in the receiver to ensure ion transport. Alternative designs will need to be considered if the basic geometry does not support efficient solar to hydrogen efficiency.

13.5 Solar Thermochemical Hydrogen Production

STCH production utilizing a power tower (heliostat solar collection field concentrating on a central tower focal point) holds challenges in both the STCH reaction material selection and the design of the system (i.e., solar receiver, gas handling, heat exchanger). Similar to PEC, a set of active materials achieving all threshold targets has not yet been identified by researchers, although this is an active area of research. Key material aspects that will impact the cost of H_2 the most are hydrogen productivity (µmol H_2/g), cost ($/kg active material), and service life (years at operational cycle time). Additionally, improving STCH efficiency through lower material reduction temperatures, high-efficiency solar receiver designs, improving reaction rates inside reactors with vacuum or purge gases, and improvement of thermal efficiency through heat recuperation and better heat exchanger design should further be considered to make STCH a more competitive hydrogen production technology.

14 Appendix A: Contract Publications and Presentations

(The listing below does not include quarterly technical status reports.)

1. James, B.D., DeSantis, D.A. Update to HPTT: Technical and Economic Analysis of Biofermentation of Corn Stover, Presentation to Hydrogen Production Technical Team (HPTT), November 15th, 2016.
2. Brian James, and Daniel DeSantis. "Analysis of Advanced H2 Production and Delivery Pathways," presented at the DOE Annual Merit Review, Washington D.C., June 2018.
3. Brian D. James, Daniel A. DeSantis "Recent case study updates for water electrolysis and a case study of long-distance energy transmission," presented to the HPTT. April 2, 2019.
4. Brian D. James, Genevieve Saur, and Daniel A. DeSantis, "Analysis of Advanced H2 Production Pathways," presented at the Department of Energy Annual Merit Review Meeting, April 30, 2019, Washington, D.C.
5. Daniel DeSantis and Brian James, "Analysis of Advanced H_2 Production & Delivery Pathways," presented to the H2A Tech Team. January 8, 2020.
6. David Peterson, James Vickers, and Dan DeSantis, "Hydrogen Production Cost From PEM Electrolysis - 2019," DOE Hydrogen and Fuel Cells Program Record (19009) - DRAFT to DOE. February 3, 2020. https://www.hydrogen.energy.gov/pdfs/19009_h2_production_cost_pem_electrolysis_2019.pdf
7. Daniel DeSantis and Brian James, "H2IQ: Analysis of Advanced Hydrogen Production and Delivery Pathways." March 25, 2020. https://www.energy.gov/eere/fuelcells/2020-fuel-cell-technologies-office-webinar-archives
8. David Peterson, James Vickers, and Daniel DeSantis, "Hydrogen Production Cost From High Temperature Electrolysis - 2020," DOE Hydrogen and Fuel Cells Program Record (20006) - DRAFT to DOE. September 29, 2020. https://www.hydrogen.energy.gov/pdfs/20006-production-cost-high-temperature-electrolysis.pdf
9. Badgett, A., Ruth, M., James, B., Pivovar, B., "Techno-economic Analysis (TEA) Methods Identifying Cost Reduction Potential for Water Electrolysis Systems", Current Opinion in Chemical Engineering, Volume 33, September 2021. https://doi.org/10.1016/j.coche.2021.100714
10. Daniel DeSantis, Brian D. James, Cassidy Houchins, Genevieve Saur, Maxim Lyubovsky, "Cost of long-distance energy transmission by different carriers", iScience, Volume 24, Issue 12, December 19, 2021. https://doi.org/10.1016/j.isci.2021.103495