

**Final Technical Report**  
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**Feasibility Study of Hydrogen Production from Existing  
Nuclear Power Plants using Alkaline Electrolysis**

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## Executive Summary

The mid-range industrial market currently consumes 4.2 million metric tons of hydrogen per year and has an annual growth rate of 15%.<sup>1,2</sup> Industries in this range require between 100 and 1000 kilograms of hydrogen per day and comprise a wide range of operations such as food hydrogenation, electronic chip fabrication, metals processing and nuclear reactor chemistry modulation.

This market is currently served via delivered or piped in hydrogen, generated using methods such as steam methane reforming that produce significant CO<sub>2</sub> emissions because natural gas is required for the process. One way to produce hydrogen with no direct emissions is to electrolyze water. If the electricity source used to power the electrolyzer does not generate CO<sub>2</sub>, then the entire cycle of energy production and consumption can be free of greenhouse gas generation.

Hydrogen generated on demand for industrial or commercial applications has many advantages over delivered or piped in hydrogen due to regulatory and safety issues related to storage as well as significant infrastructure and delivery costs. GE has developed technology for an electrolyzer made primarily of advanced plastics which significantly lowers the cost of the stack module. On-site production of hydrogen on a nuclear site using electrolysis has the combined benefits of making hydrogen for needed processes (e.g., generator cooling) while making use of a very low cost source of electricity compared to fossil fuels.

To date, electrolysis has not been the preferred method for large-scale hydrogen production because it is more costly than reforming. Costs for electrolysis hydrogen are typically over \$8 per kilogram, while hydrogen made by large-scale steam methane reformers may cost less than \$2 per kilogram based on \$3.15/MMBtu cost of natural gas.<sup>3</sup> A key aspect of the Department of Energy's HFCIT (Hydrogen Fuel Cell Infrastructure Technology) program is to make the environmental benefits of electrolysis hydrogen possible by reducing the price of electrolysis hydrogen to under \$3 per kg.

An electrolysis-based hydrogen production system consists of the electrolyzer stack and the support system. The support system consists of equipment that control and monitor the input, output, and pressure of all resources required for production of H<sub>2</sub>. A significant portion of the overall capital costs of hydrogen electrolysis are in the stack itself. Also, since the stack tends to need replacing before the majority of system components, the capital cost of the stack has a greater effect on the overall hydrogen cost than the bulk of the support system. Therefore, the primary focus of this program was on reducing stack costs.

GE focused on lowering the materials, manufacturing, and assembly costs of the stack at every step, while insuring stack integrity and life targets were met. Numerous innovations in manifold design, electrodeposition, plastic joining, and methods to enable long-term stack integrity under elevated pressure have been

made as part of this project. A series of bench scale tests were completed for both the electrodeposition process and cell joining processes prior to assembly of the full-sized prototype stack. The full-sized stack bested the bench-scale performance at both ambient and at 15 bar pressure, and has performance comparable to stack efficiencies achieved by current state-of-the art electrolyzers, at a cost that is estimated to be 25-50% less than existing electrolyzers.

Another barrier to acceptance of electrolysis as the preferred hydrogen production method is the costs of the support system itself. The prototype system used to test the electrolyzer stack was developed under an earlier DoE program in which hydrogen was produced at ambient pressures. The system required considerable upgrades and a few design changes in order to achieve the 15 bar pressure required to meet program goals. This report details operation and maintenance experience of the electrolyzer system as well as suggested design changes for improved reliability. Another finding of the system study is the high cost of power electronics as a proportion of the total capital costs (which currently equate to about \$100-\$150/KW with volume pricing assumptions) as well as a need for system designs that reduce equipment exposure to the electrolyte.

Both the stack and system costs have been considered with a focus on utility customers, for whom capital cost is the primary driver of hydrogen cost. The GE research team quantified the performance targets necessary to meet the needs of these customers and developed a strategy to meet these needs at capital cost targets consistent with the hydrogen cost goals set by DOE. GE has achieved the goal of significantly reducing the capital cost of the stack and demonstrating performance that has near-term potential for successful commercialization. Going forward, the best opportunity to meet the 2012 DoE cost goals rely on significantly reducing the cost of power electronics and focusing on further improvement of system designs.

## **Project Goals**

The overall goal of the DOE HFCIT program is to accelerate the development and successful market introduction of hydrogen production, delivery and storage technologies.

The main objective of the project is to evaluate the feasibility of hydrogen production using alkaline electrolysis powered by nuclear energy at industrial scales of up to 1 kilogram of hydrogen per second. The determining factors of this evaluation will be cost competitiveness, regulatory considerations, and environmental impact. GE has partnered with NREL and Entergy to determine market requirements, evaluate electrolyzer technical performance, and generate reports and models to complete the study.

## **Technical Barriers**

This project addresses the following technical barriers from Hydrogen, Fuel Cells and Infrastructure Technologies Program Multi-Year Research, Development and Demonstration Plan:

- (G) Capital Cost
- (I) Grid Electricity Emissions

## **Technical Targets**

The goal of this project is to develop a low-cost alkaline electrolysis system. The relevant DOE hydrogen production targets are listed in Table 1. Note that the assumptions used in this analysis differs from the standard H2A forecourt analysis completed in earlier quarterly reports. The previous analyses included costs for compression, storage, and dispensing to vehicles. Since the current program is focused on industrial and commercial applications, the estimates for the “Project” portion have been calculated using the H2A model V1.08 changed to 1000 kg/day. Also, for the sake of clarity, an attempt has been made to separate the stack cost from the overall system costs since stack cost reduction was the goal of this program. A list of assumptions is as follows:

- Electricity cost: \$0.05/kWh for all cases (for comparison purposes). Differences in values reflect effect of stack efficiency on cost.
- Balance of system cost (\$500/kW) for the “Project” assumes volume pricing for power electronics.
- Operating and capital costs do not include dispensing to vehicles
- 1.9% inflation rate, 10% After Tax Real Internal Rate of Return, 100% Equity Financing, 10-year MACRS depreciation schedule, 20-year analysis period, 38.9% overall tax rate, and 15% working capital.
- Stack replacement every 10 years.
- Capital Costs are calculated using the Stack Capacity Factor

	Units	DOE 2006 Status	DOE 2012 Target	Project
Electrolyzer Cell Efficiency	% (LHV)	62%	69%	68.3%
Electrolyzer Stack and System Capacity Factor	Stack %	<b>70</b>	<b>70</b>	<b>90</b>
	System %	70	70	50
Electrolyzer Stack and System Energy Efficiency	%(LHV)	62%	66.9%	66.2%
	%(HHV)	73%	79%	78.2%
Electrolyzer Stack and System Capital Cost**	Stack \$/kg H2 (\$/kW)	\$0.49 (\$200/kW)		\$0.39 (\$150/kW)
	System Total \$/kg H2 (\$/kW)	\$1.08 (\$665/kW)	\$0.70 (\$400/kW)	\$0.82 (\$650/kW)
Electricity (System) cost	\$/kg H2	\$2.67	\$2.40	\$2.43
O&M	\$/kg H2	\$0.80	\$0.60	\$0.59
<b>Total Hydrogen Cost</b>	<b>\$/kg H2</b>	<b>\$4.56</b>	<b>\$3.70</b>	<b>\$3.84</b>

<http://www1.eere.energy.gov/hydrogenandfuelcells/mypp/pdfs/production.pdf>

\*\* Prior analysis conducted by the DoE and NREL assumed the stack would be replaced at an expense equal to 30% of the overall system cost. System cost is the assumed stack cost plus balance of plant (BOP).

## **Summary of Products**

Multiple products were generated in the course of this work including:

### **Technologies & Techniques**

- Designed and built a 10 x 2700 cm<sup>2</sup> cell demonstration plastic stack module for operation at 15 bar, which has verified material, manufacture, and assembly cost estimates.
- Demonstrated target efficiency of electrodes that utilize high surface area, low-cost in-situ electrodeposition process at the bench scale.
- Demonstrated efficiency of full-size demonstration stack at 15 bar pressure.
- Results of accelerated material and joining method tests indicate promise for use of polysulfone plastics for this application.

### **Publications**

Bourgeois, R; Swalla, D.R.; Ramsden, T.; "Low Cost Electrolyzer Technology for Industrial Hydrogen Markets", National Hydrogen Association Conference, March 31-April 4, 2008, Sacramento, Ca.

### **Patent Applications**

- Electrolysis System for Fertilizer Synthesis and Carbon Capture, 226133 DOE#S-112,474
- Electrolyzer Assembly Method and System, 226674, DOE #S-113,160
- Pressurized Electrolysis Stack with Thermal Expansion Capability, 227342 DOE #S-113, 342
- Methods and Systems for In-Situ Electroplating of Electrodes, 232538, DOE#S-116, 636
- Methods and Systems for Assembling Electrolyzer Stacks, 231431, DOE#S-117,052

### **Other Products**

- Determined electrolyzer capital costs
- Determined industrial and 1 kg per second commercial scale demonstration system designs and associated costs.

## Approach

Evaluate the feasibility of nuclear electricity and electrolysis for large-scale hydrogen generation by leveraging the joint experience of GE, Entergy, and NREL in low-cost electrolyzer stack technology, nuclear electricity markets, and modeling expertise, respectively.

Project tasks were set to meet this overall goal:

- 1) Research early adopter and long-term markets to determine customer requirements for an electrolyzer product.
- 2) Develop a technical strategy and create a preliminary design capable of meeting the customer needs and the DOE cost goals.
- 3) Engineer and analyze electrochemical cell materials and configurations capable of meeting the performance and cost goals
- 4) Continue development of a prototype electrolysis system capable of delivering 1 kg of hydrogen per hour and determine prototypical system reliability, operation, and maintenance costs.
- 5) Perform design and construction of a prototype pressurized stack module of at least 10 cells and study lifetime of stack under operation at maximum 15 bar pressure.
- 6) Produce information on viable system designs.
- 7) Determine suitability of materials and validate a stack lifetime that exceeds the design life target using accelerated life testing methods.

## 1. MARKET DEFINITION AND REQUIREMENTS

### 1.1 Electrolyzer Market

Increasing commercial interest in electrolyzers has developed due to a number of factors. Power utilities continue to seek cost effective methods to leverage power generation, transmission, and distribution assets during periods of low consumer demand. Also, steam methane reforming (SMR), which has traditionally been the accepted method for large-scale hydrogen production, becomes less attractive as natural gas prices continue to rise along with public pressure to reduce CO<sub>2</sub> emissions. Finally, the infrastructure, delivery, storage, and regulatory costs related to trucked in or piped in hydrogen make on-site production via electrolysis increasingly attractive to many in the mid-scale industrial market.

Reducing the initial capital cost of electrolysis is key to displacing other methods for hydrogen production. However, full acceptance of this technology and continued growth in these markets will only be realized if long-term operation and maintenance costs are also kept low. Based on the needs of the early adopter industrial customers, GE has developed a product prototype at the 1-5 kg/hr scale that embodies the technologies needed to manufacture modular building block stacks for larger systems.

Hydrogen market segments studied for this task are broken into three categories, as shown in Figure 1: Small (1-100 kg per day), medium (100-1000 kg per day) and large (1000+ kg per day). Nuclear plant water treatment for boiling water reactors (BWR) bridges the small and medium sized market, based on plant size.

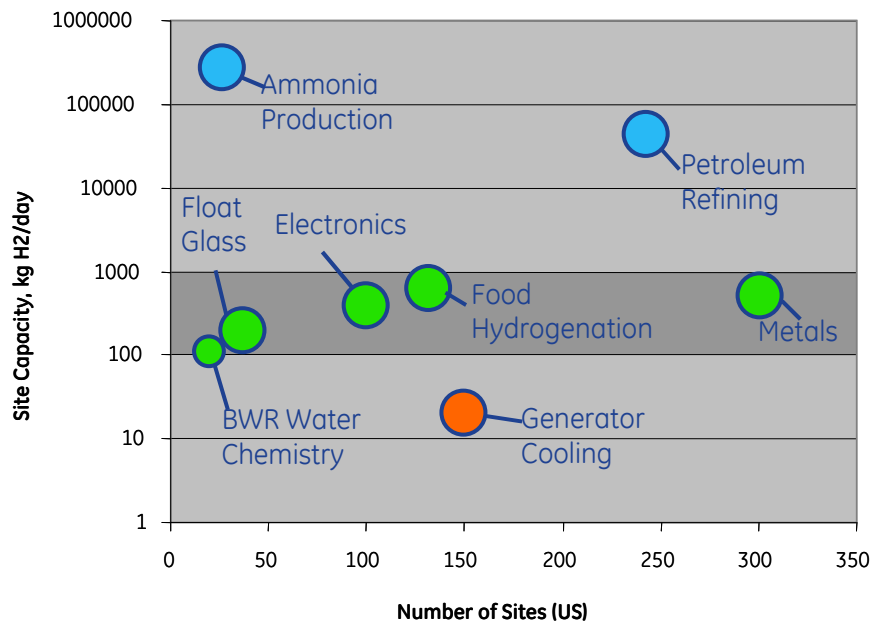


Figure 1: Industrial and commercial hydrogen markets<sup>4</sup>

### **1.1.1 Small Industrial Use (1– 100 kg/day)**

Electric utility use of hydrogen makes up the largest portion of small industrial users, and falls into two areas: Intergranular Stress Corrosion Cracking (IGSCC) mitigation for nuclear power plants and hydrogen cooling for generators. IGSCC is by far the larger of the two applications at approximately 100 kg H<sub>2</sub> per day versus approximately 2 kg H<sub>2</sub> per day for cooling use.

#### *IGSCC Remediation*

Hydrogen for IGSCC mitigation is limited to boiling water reactor (BWR) plants, and therefore applies to only 34 of the generating units in the US. There is considerable variation from site to site in IGSCC H<sub>2</sub> consumption. With the exception of the Cooper Nuclear Station in Nebraska, which produces hydrogen on site by electrolysis, hydrogen is delivered to IGSCC sites in liquid form and stored in tank farms.

IGSCC of piping and internals in the BWR is mitigated by application of Hydrogen Water Chemistry (HWC). Designers expect that the use of hydrogen for this purpose will be in the range of 40 to 340 kg H<sub>2</sub> /day. For example, the Entergy BWR units Pilgrim, Fitzpatrick, Vermont Yankee, Grand Gulf 1 and River Bend each use approximately 127 kg H<sub>2</sub> /day.

#### *Generator Cooling*

Many power generation units use H<sub>2</sub> to cool their generators. In addition to the 34 BWR units in the US there are 69 pressurized water reactor (PWR) nuclear power plant units and over 1200 fossil fuel generating units (2002 Economic Census). This is a very large number of users, but the use per unit for cooling is small. For BWR units, the H<sub>2</sub> used for cooling is “insignificant” relative to IGSCC. In the experience of power plant operators surveyed, cooling use is approximately 1-2 kg H<sub>2</sub> per day per generating unit.

The BWR sites, since they have liquid H<sub>2</sub> delivered to tank farms for IGSCC, use liquid H<sub>2</sub> for cooling also. Virtually all other sites are reported to have the H<sub>2</sub> delivered as compressed gas.

### **1.1.2 Medium Industrial Use (100 – 1000 kg/day)**

Examples of industries in this category are float glass production and metals processing. The metals industry is by far the larger user of H<sub>2</sub> with hundreds of installations in the United States.

#### *Float Glass*

About 90% of the world's flat glass is currently formed via the float method—which produces glass with extremely flat, parallel surfaces. In this process, a continuous ribbon of glass is floated on a bed of molten tin until it cools enough to be handled as a solid. The tin is prone to oxidation, which would form impurities in the glass. Also, an oxygen atmosphere can form a cloudy film on hot glass. Therefore, the atmosphere for this process is carefully controlled, with mostly N<sub>2</sub> and some H<sub>2</sub> as a scavenging agent to eliminate oxygen.

There are 23 float glass lines in North America. Each uses about 110 to 140 kg/day of  $H_2$ . The purity requirement is less than 5 ppm  $O_2$ . Most lines buy liquid hydrogen. Industry contacts indicated a willingness to evaluate electrolysis as an option. Key evaluation points would be cost of hydrogen and capital cost, size, and availability of the electrolyzer. Availability would need to be greater than 95%.

### *Metals*

Hydrogen is used in the metals industry in primary production (reduce metal oxide ore to elemental form) and in secondary processing (heat treating and annealing). The majority of  $H_2$  use in the metals industry is for secondary processing, and the fastest growing segment of the secondary processing is steel annealing. Reasoning that fast growth and new equipment would be the highest priority segment to enter, it is the annealing segment of the market that took focus for this study.

Annealing furnaces have traditionally run with atmospheres of about 10%  $H_2$  (ranging from about 5% to 70%) with the balance  $N_2$ . Increasingly, 100%  $H_2$  atmospheres are being adopted. Annealing furnaces may be continuous type for treating strip or sheet material, or they may be box furnaces for base or batch-type operation. Annealing furnaces are manufactured by companies like Ebner Furnaces, Inc., LOI Inc. Industrial Furnaces, and Rad-Con, Inc., and are supplied to large users like US Steel and the many smaller "heat treat and anneal" companies scattered around the US. According to industry experts there are about 6000 such operations installed around the world, with 500 to 1000 of them in the United States.

Most users of annealing and heat-treating furnaces buy liquid  $H_2$ , but some make their own  $H_2$  on site. The furnace manufacturer may specify, recommend, or buy and supply the  $H_2$  production equipment for those customers. HyRadix, H2gen and Proton were mentioned as common manufacturers of on site  $H_2$  generating equipment.

The typical annealing and heat-treating company might have several furnaces and use on the order of 800 kg/day of  $H_2$ . Purity was generally expressed in terms of dew point, with  $-40^\circ C$  to  $-80^\circ C$  mentioned commonly, with  $O_2 < 5$  ppm typical, although a range of  $O_2$  from 2 ppm to 10 ppm was also cited. Continuous annealing furnaces were said to use in the range of 50 to 400 kg/day  $H_2$ . Batch furnaces with 100%  $H_2$  atmosphere might use 0.25 kg to 0.5 kg per ton of steel. According to one furnace manufacturer the base cycle starts with a  $N_2$  purge followed by  $H_2$  fill rate of 2.4 kg/hr, then tapers off ("turns down") as 100%  $H_2$  atmosphere is approached.

### **1.1.3 Large Industrial Use (> 1000 kg/day)**

Ammonia production is one of the largest users of hydrogen, along with oil refining. A typical ammonia plant might produce about 1500 short tons per day of ammonia. There are several such plants scattered around the country.

Ammonia plants use a primary reformer to extract  $H_2$  by steam methane reformation (SMR) followed by a secondary reformer that captures  $N_2$  from air for synthesis of ammonia. This is a clever integration of chemical processes that makes the  $N_2$  nearly free. Older plants operate at a pressure of about 2200 to 2500 psi to aid in the synthesis. Newer, more energy efficient plants operate at 1200 to 1300 psi.

If an electrolyzer were to be used instead of SMR to produce  $H_2$  then some additional provision would need to be made for producing high purity  $N_2$ . For example, an air separation plant might be sited near the ammonia plant. The gases produced ( $H_2$  and  $N_2$ ) would need to be very high purity as concentrations of oxygen compounds greater than 1 or 2 ppm would poison the catalysts used in ammonia production. In concept this could be done, but the practicality of such a design would need to be looked at carefully. Evaluation of favorable economics would depend on replacement of the complete front end of the plant, not just delivering competitively priced  $H_2$ .

## 1.2 Cost of Hydrogen

The cost of produced hydrogen is primarily comprised of capital and maintenance costs. Currently available industrial electrolyzers in the 100-1000 kg per day range cost approximately \$4000/kW, or \$200,000 per kilogram per hour production capacity on the basis of 50 kWh/kg  $H_2$ .<sup>5</sup> The other significant element of the cost of electrolysis hydrogen is the cost of electric power.

If electricity is available at low cost, the sensitivity of total hydrogen cost to improvements in efficiency is not great. For instance, an increase in efficiency from 68% to 76% decreases the cost of hydrogen only about \$0.31/kg if the cost of electricity is \$0.05/kWh. In addition, there is a limit to the reduction in energy costs because the conversion efficiency cannot be greater than 100%. Therefore, we chose to concentrate our efforts on a dramatic reduction in capital cost.

In the current market, the cost of either the electrolyzer stack or the balance of system equipment alone is too high to reach DoE's targets for hydrogen cost (see Table 1: Technical Targets). GE has addressed these two costs in different ways. Our research indicates that, based on the relative entitlements of the stack and system capital costs, the best strategy consistent with our intended market is to reduce stack cost with a technology and design solution while taking advantage of economies of scale and volume to reduce system costs.

For electrolysis hydrogen to compete with captive reforming in the large-scale commercial market, the electrolysis plant would have to be immense, on the order of one kilogram per second production capacity. At this scale it may be reasonable to target a capital cost of \$400/kW. In addition, the large amounts of electricity required (180-200 MWe) are consistent with base-load nuclear power, a source of electricity with potential for very low cost. Assuming a capital cost of \$400/kW and electricity costing 1.5 cents per kWh, the cost of hydrogen at this scale would be much lower, as

shown in Table 2. These costs are less than \$2 per kg, which may be attractive for captive large-scale hydrogen production if natural gas feedstock costs rise. For instance, since 2000, the price of natural gas has steadily climbed from about \$3/MMBtu to a high of \$14/MMBtu in 2008.<sup>6</sup>

**Table 2: H2A Model Cost of Hydrogen (Capacity = 1000 kg/day). Internal rate of return (IRR) is 10%**

Cost of Electricity (Cents/kWh)	Capital Cost, \$/KW			
	\$4000	\$800	\$650	\$400
1.0	\$8.60	\$2.50	\$2.22	\$1.74
3.0	\$9.63	\$3.53	\$3.24	\$2.77
5.0	\$10.65	\$4.55	\$4.27	\$3.78
8.0	\$12.19	\$6.09	\$5.81	\$5.33
10.0	\$13.22	\$7.12	\$6.80	\$6.36
15.0	\$15.79	\$9.68	\$9.40	\$8.92

The assumptions used to build Table 2 are the same as those used in the Technical Targets section. Many of the basic assumptions often used in the H2A model pertaining to retail sale of hydrogen for vehicle use were changed to reflect the medium scale industrial market (between 100 and 1000 kg/day).

- basis is NREL's H2A Forecourt Modeling Tool v.1.0.8 (1,500 kg/day Electrolyzer case).<sup>7</sup>
- design capacity changed to 1000 kg/day (targeting medium scale industrial market).
- operation and maintenance costs are projected from actual costs obtained from operating electrolyzer at Cooper Nuclear Facility and project experience.
- electrolyzer production rate is 50 kWh/kg H<sub>2</sub> (LHV) (66% efficient).
- Capacity factor is 70%
- **internal rate of return (IRR) is 10%**
- stack replacement every 10 years

The analysis using NREL's H2A Forecourt Modeling Tool v.1.0.8 (modified to 1000 kg/day), and assuming industrial electricity rates of 5 cents per kilowatt-hour, a capital cost of \$4000/kW, and a production efficiency of 50 kWh/kg H<sub>2</sub> finds that hydrogen can be produced for about \$10 to \$11 per kg. At capital cost of \$800/kW and the same efficiency, however, the same model predicts costs of \$4 to \$5 per kg.

Limited comparisons between these values and the cost of delivered or piped in hydrogen to industrial facilities can be made at this time. This is due to the fact that suppliers treat pricing as sensitive information and won't discuss it except in a purchase agreement. These agreements typically also contain terms preventing the buyer from disclosing pricing. Early estimates indicated that liquid hydrogen is generally used by those with consumption in the range of 10k to 1,000k SCF/day at a

cost between \$6 and \$8 per thousand SCF. Expressed in kg that would be about 24 to 2400 kg/day at a price between \$2.50 and \$3.33 per kg.

It was reported in Q2FY2008 from information provided by Entergy Nuclear that the Pilgrim Nuclear Station used 32 scfm of delivered hydrogen in liquid form for IGSCC. The delivered cost of \$2.10/100 ft<sup>3</sup>/year for 16.82 million ft<sup>3</sup>/year of hydrogen amounts to \$353,203/year or about \$8.25/kg. If storage and personnel costs are added, the total cost for delivered hydrogen for the Pilgrim Nuclear Station came to \$596,885 or about \$14/kg. By comparison, the total cost of hydrogen generated at the same site using an 18-cell Stuart bipolar electrolyzer was \$158,022/year or \$3.69/kg. The reported electricity and O&M cost/yr are \$0.03 kWh and \$28K/yr, respectively. A ballpark estimate of the capital cost can be made by inputting the reported numbers into the NREL H2A model using the reported information. A capital cost of about \$800/KW, along with the reported electricity and O&M cost results in a hydrogen cost of about \$4.20/kg as calculated by the NREL H2A model. These costs appear quite attractive when compared to the cost of delivered hydrogen. Unfortunately, the operating experience of the overall EHWC (Electrolytic Hydrogen Water Chemistry) system at Pilgrim Station did not meet expectations. System availability, largely due to the hydrogen compression module (HCM), was extremely low. Furthermore, the location of the HCM was sited in a high radiation area, so maintenance during plant operation was restricted. The EHWC operated for only several years before being permanently shutdown and eventually abandoned.

The seven-cell Hydrogenics unipolar electrolyzer currently operating at the Cooper Station in Nebraska has a much better operating experience than the Pilgrim Nuclear Station. However, it is currently the only U.S. nuclear station with an operating electrolyzer. The performance history of Cooper's electrolyzer averages between 90% and 97% availability since beginning operation. The Cooper Station electrolyzer facility generates hydrogen at a rate of 7.5 SCFM (27 kg/day) or 3,942,000 SCF per year. The estimated cost for on-site hydrogen production is \$0.058/scf (\$22.80/kg) and \$0.166/scf (\$45.56/kg) for off-site purchased hydrogen, respectively. The electricity cost was not reported for this facility. The reported O&M cost was \$90K/year. The H2A model predicts the cost of hydrogen for this capacity to be about \$19.50/kg assuming the cost of electricity is \$0.05/kWh, a reported O&M cost (\$90K/yr) and assumed capital cost equal to \$4000/KW, which is comparable to the reported costs for this site.

The capital and O&M costs appear to differ significantly between the two nuclear sites, which makes comparison difficult. However, the cost benefit of on-site hydrogen over delivered hydrogen is clear. The cost of onsite generation is about half the cost of delivery and storage of hydrogen for these processes. Both sites reported the cost of onsite generation to be at least half the cost of delivery and storage of hydrogen for these processes.

### 1.3 Efficiency and Capacity of Current Electrolysis Technology

Hydrogen production rate and cost is directly related to the efficiency and capacity of the electrolyzer stack. Any improvements in reducing capital costs via material, manufacturing, or assembly methods, therefore, cannot compromise overall efficiency of the stack if long-term cost goals are to be met. However, the methods used to report efficiency and capacity sometimes differ or are unclear. When comparing efficiencies for different stack technologies, it is important to understand how they are derived.

The relationship between current and voltage in an electrochemical cell is a good indicator of its performance. According to Faraday's Law the heat of formation  $\Delta_f H_0$  of hydrogen can also be expressed as an electrochemical potential ("standard potential")

$$U_{00} = - \Delta_f H_0 / n_e F$$

with  $n_e = 2$  being the number of electrons participating in the conversion and  $F$  = the Faraday constant (96,485 Coulomb/mol)

However, only a fraction of the heat of formation  $\Delta_f H_0$  is available for reversible energy conversion. This fraction is given by the Gibbs Free Energy  $\Delta_f G_0 = 237 \text{ kJ mol}^{-1}$  for water at 25°C. Consequently, the theoretical voltage required to split water at 25°C by electrolysis is 1.23 Volts. The value under other conditions is described by the Nernst equation.

Therefore, the rate of hydrogen production is directly proportional to the current, and to produce hydrogen at rates of industrial interest, a higher voltage must be applied to the cell. The excess voltage is determined by the overpotentials of the anode and the cathode, the electrical resistance of the solution and the electrical resistance across the membrane. The electrode overpotentials can be estimated from Tafel plots, which describe the current / potential relationship for specific electrode materials.

The excess voltage increases the power consumption of the stack and reduces its efficiency. The efficiency can be determined by the ratio of the thermodynamic voltage to the voltage necessary to achieve the required production rate. A common point of debate concerns the use of the LHV (242 KJ/mol) or HHV (286 KJ/mol) to determine the efficiency. Using the HHV gives a standard potential of 1.48V. Therefore, the efficiency using the HHV for stack with a per cell voltage requirement of 1.6V is  $1.48\text{V}/1.8\text{V} = 82.2\%$  while the efficiency of the same cell using the Gibbs potential is  $1.23\text{V}/1.8\text{V} = 68.3\%$ . In this report, the Gibbs potential or LHV is used. Any data from the literature presented in HHV is converted to LHV for comparison. Table 2 lists the available data for currently available bipolar alkaline electrolyzers obtained from surveys conducted by NREL<sup>8</sup>. Data for PEM and unipolar electrolyzers can also be found in Ref(8). The per cell voltage by dividing the theoretical voltage (1.23V) by the LHV efficiency. Every 0.1V drop in cell voltage drops the stack efficiency

by about 4-5% (LHV). Unfortunately, there is not enough information provided to determine the current density for a given cell voltage. This market data, in addition to information regarding stack efficiencies gained from non-proprietary discussions with electrolyzer makers will be compared to GE's stack technology in Section 4.

**Table 2: Electrolyzer Market Data**

HHV            39 kwh/kg  
LHV            33 kwh/kg

Manufacturer	Energy Req'mt (kwh/kg)	**Stack Energy Req'mt (kwh/kg)	Efficiency (% HHV)	Efficiency (%LHV)	Power input (KW)	H2 production (kg/hr)	Prod. Pressure (bar)
Stuart IMET 1000	53.4	51.9	75.1	63.6	288	5.4	20
Teledyne: EC 750	62.3	60.8	64.1	54.3	235	3.8	16
Norsk Hydro: Type 5040	53.4	51.9	75.1	63.6	2328	43.6	0.02

## 2. PRESSURIZED ELECTROLYZER STACK DESIGN AND ANALYSIS

Current and target costs for the electrolyzer stack and balance of plant have been specified. Our research indicates that the best strategy consistent with our intended market is to reduce stack cost with a technology and design solution while taking advantage of economies of scale and volume to reduce system costs.

A major component of electrolyzer stack cost is the stack structural materials and assembly labor. GE has invented a low cost electrolyzer stack which eliminates the cost and complexity of the traditional assembly of the cells with bolts and insulating gaskets. The GE stack module is constructed by assembling multiple cells into a single non-conductive plastic frame, which also provides for internal liquid and gas passages. The frame may be constructed from individual plastic components by various joining methods or in one piece by a molding or casting process. The life target goals for the plastic stack have been insured in part by enclosing it within a 15 bar capable pressure vessel that has been specially designed to limit the net tensile stress and creep deformation.

A low-cost, in-situ electrodeposition process has been developed to coat the active surface onto the electrodes. Both bench scale tests and tests conducted on the full-size prototype stack have confirmed the entitlement of our cell materials to reach the performance targets necessary to achieve stack cost goals.

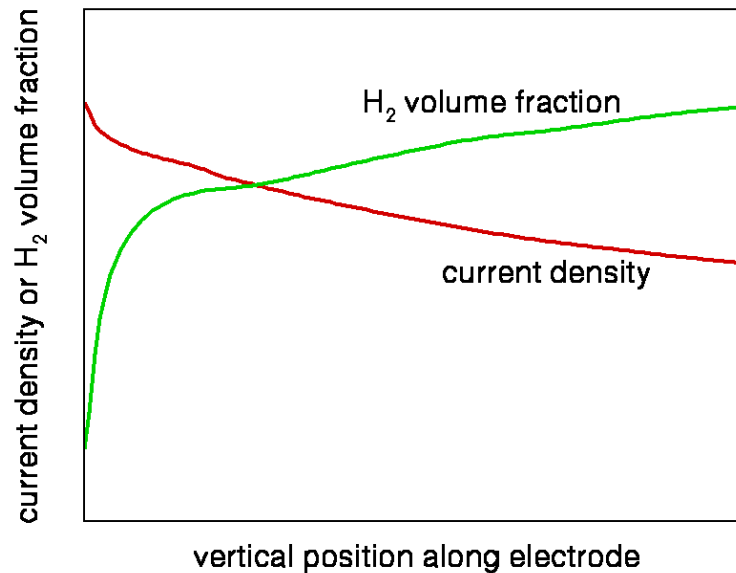
Numerous patents have been filed for design of the electrodes and pressure vessel, for the cell joining procedure, as well as the in-situ deposition process that is described in the following sections.

## 2.1 Cell design and catalyst deposition process

### Cell Geometry Configuration

For the electrolyzer to operate well, water and electrolyte must be distributed evenly along the electrode surfaces while oxygen and hydrogen gas is carried out of the cell. The electrode geometry chosen for the current program is largely based on a design developed in a 2004 DoE program. Results of the prior analysis are shown here for clarity.

Bubbles on the electrode reduce active reaction area, and the presence of bubbles also decreases the effective electric conductivity of the electrolyte by reducing the cross-sectional area of pure electrolyte available for current transport. This is a complicated and non-linear physical situation, which the model has successfully predicted. [Figure 2]

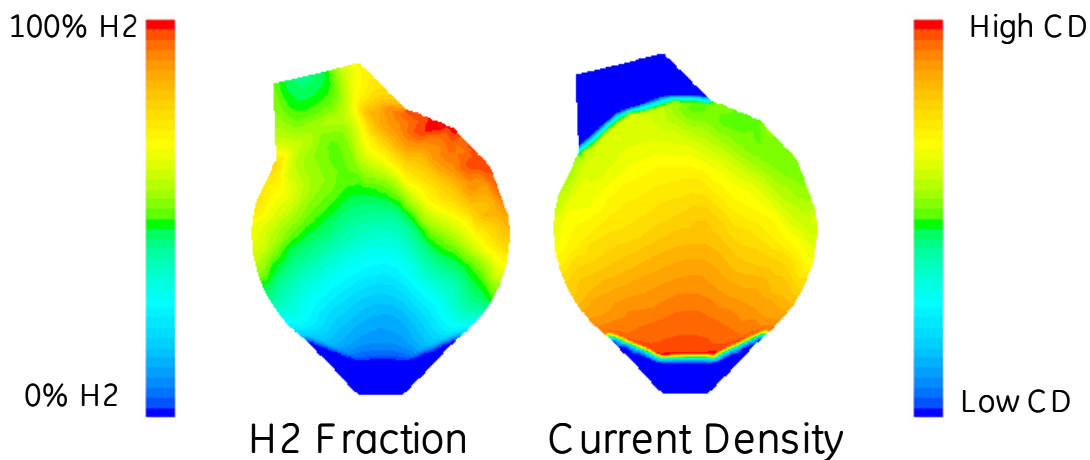


**Figure 2: Effect of evolved H<sub>2</sub> on local current density**

In 2004, analytical models were developed to predict both gas and liquid distributions over a circular cell face. This model was used to optimize the shape and geometry of the cells used in the current program to establish an entitlement in the megawatt size range needed for utility applications. Figure 3 shows the effect of bubble generation on current density in the baseline case of a large (1.5 m<sup>2</sup>) round cell. The maximum current density is near the entrance at the bottom of the cell, where the hydrogen concentration is lowest. The minimum current density is near the top of the cell on the side opposite the exit, where hydrogen bubbles have accumulated.

The performance of the cell is also tied to the temperature and flow rate of the electrolyte. Naturally, the ionic transport mechanism required to split hydrogen is more efficient at higher temperatures. The optimal flow rate for the electrolyzer

system needs to be high enough to insure that the hydrogen bubbles are effectively swept off of the electrode to avoid “masking” of the active surface, but low enough to insure that the gases have time to percolate out of the fluid in the separation tanks. Increasing the pressure of the electrolyte tends to decrease the bubble size, and hence increases the propensity for bubble masking and time required to achieve full separation of the gas from the fluid. Therefore, the desired operating pressure must be taken into account in the cell and system design.



**Figure 3: Cell CFD Results**

#### Electrode Materials

GE electrode technology applies a high effective surface area, nickel-based coating to the base metal bipolar plate for high performance at low cost. In 2005, we achieved our target performance using a wire-arc spray deposition technique. In the current program, an electrodeposition process was developed to further reduce costs without compromising performance. Electrodeposition has several potential advantages over wire-arc. It allows the use of thinner bipolar plates without warping, and is capable of coating a three dimensional electrode surface instead of a simple flat plate. This permits much closer spacing of the electrodes for reduced ohmic losses and higher efficiency at a given hydrogen production rate. In the current entitlement, the surface area was enhanced by welding a simple mesh to both sides of the bi-polar plate.

Nickel and nickel-based alloys can be readily deposited electrolytically from aqueous solutions comprising soluble ionic species of the desired metals together with other agents that control the solution conductivity, pH, and surface tension of the liquid against solid substrates. When an electric potential is applied across two electrodes

immersed in the solution, nickel ions (and any other metal ions present) will be reduced to the metallic state at the negative (cathode) face of the electrode.

The quality, i.e., the physical state and mechanical properties, of the deposited layer is a function of many parameters, including the condition of the substrate onto which the metal is to be deposited, the concentration of the ionic species present in solution, the temperature, and the deposition rate, to name a few. In the case of a co-deposit of two metals, e.g. nickel and zinc, critical to the quality of the deposit is the ratio of the two metal ions in solution, for it is one of the parameters that determines the relative amount of the two metals in the metallic deposit.

The effect of varying these parameters on performance have been studied under the current program. Other possible catalysts, such as cobalt, were also studied. Optimal performance was obtained by maximizing the active surface area, which was achieved at the macroscale by coating on a mesh, and at the microscale, by leaving a pure, stable, porous layer of material on the negative (cathode) face of the electrode. Details of the bench scale and full-size stack performance tests are included in Section 4.

#### Diaphragm Development

A literature study identified several non-asbestos diaphragm materials for study. Ceramic materials such as those based on nickel oxides or metal titanates have low specific resistance but tend to be unstable as they reduce in the presence of hydrogen. Challenges for polymer membranes include poor wettability, the risk of puncture if contacted by a metal electrode, and the possibility for gas bubble crossover through pores. We tested various polymer diaphragms in operating cells to determine impedance and in pressure hold tests to determine the minimum differential pressure (aka “bubble point”) before gas crossover. This diaphragm was used successfully in the bench scale electrolyzer system test, meeting cell performance requirements and demonstrating low gas crossover as measured by hydrogen purity measurements. The bubble point is the pressure differential that results in bubble passage through the membrane.

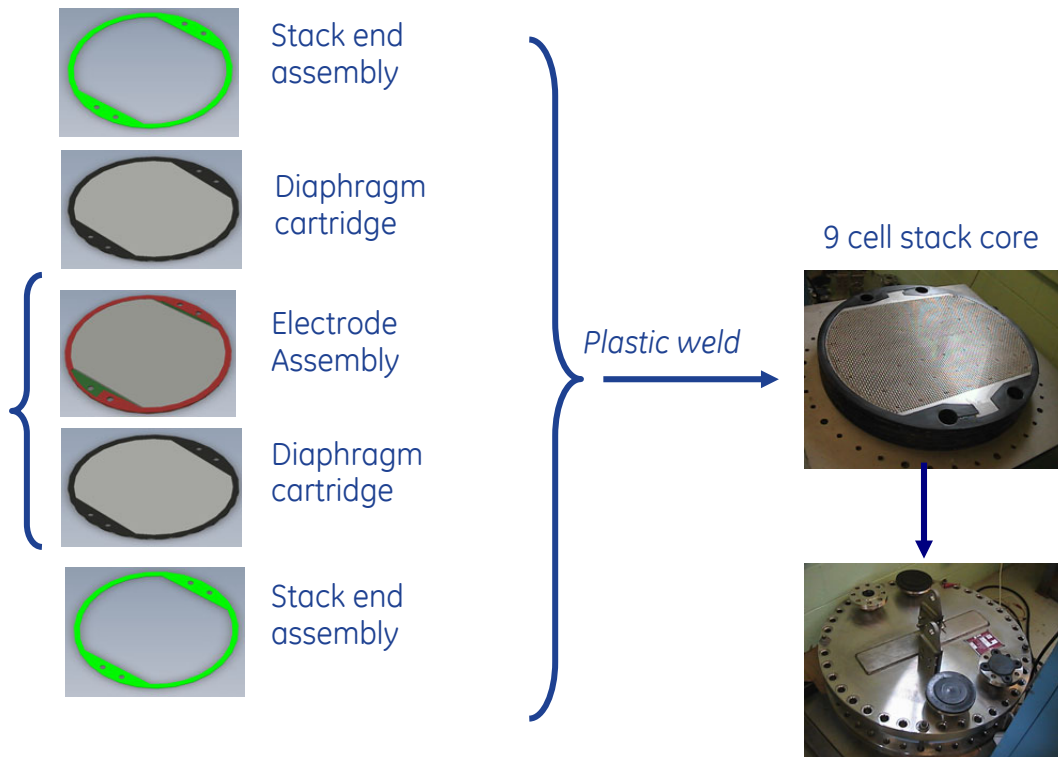
The prototype stack developed for this program has 10 cells. Each cell has one electrode and one diaphragm. The design and assembly of the cells is illustrated in Figure 4. The diaphragm and electrodes are manufactured as independent cartridges, then joined together to make the full plastic stack. The diaphragm material is captured between two plates that have circular flow passages for the inlet and outlet flow of the electrolyte from the stack.

The metal bi-polar plate is joined to an electrode spacer plate. Joined to each side of this spacer plate are plastic plates that direct the electrolyte flow over the surface of the electrode.

The pressure vessel end plates act as a current collector, so each end plate is an electrode and a “half-cell”. Each end plate has mesh welded to it, so to avoid having

the mesh hit the first diaphragm, an assembly for the stack end is added, which completes the “half-cell”. The diaphragm and electrode cartridges are then alternately joined in series, ending with the stack end assembly and capped by the pressure vessel end plate. Note that when assembling the stack, special care must be taken to label which end is the inlet or outlet ports as well as the stack orientation (cathode vs anode side) relative to the system installation.

Special tooling was developed in order to place the plastic stack into the pressure vessel without placing undue stress on the cell joints or damaging the membrane. This tooling was also used as part of the cell joining procedure.



**Figure 4: Stack assembly procedure**

## 2.2 Cell joining procedure

The plastic stack is constructed from thin, injection-molded plates, welded together. Since there are many complicated internal passages, we have selected a wire welding method for the demonstration stack. This is a robust method, with the path of the wire determining the exact location of the weld. Our patent includes other line welding techniques to achieve the same result. The wire welding method was designed in consultation with manufacturing engineers to insure that it can be easily automated and repeatable at production volumes necessary to meet overall cost goals.

Small-scale coupon tests were executed to determine the parameters required to achieve a repeatable weld, as well as methods to place the wire and optimal time

period. These tests were repeated on the full-sized plates to insure that the weld procedure could be successfully scaled up. Details of these tests are included in Section 3.2.

### 2.3 Pressure vessel design and analysis

The ASME certified, 15 bar capable pressure vessel used to house the plastic stack is shown in Figure 5. The current design was completed and ASME tested by Troy Boiler Works (Troy, NY) in collaboration with GE engineers. The front end plate has inlet and outlet ports located at the lower and upper portions of the vessel, respectively. While either end of the stack can act as the cathode or anode, having the front end, negatively charged plate bolted to the system eliminates the need to electrically insulate the stack from the system as long as the system itself is grounded.

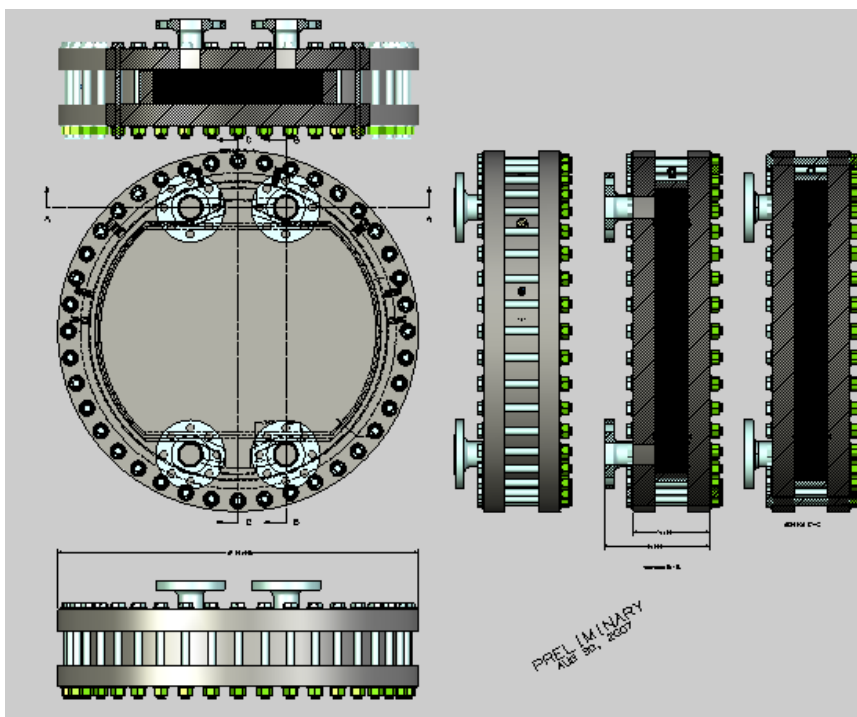


Figure 5: Electrolyzer stack pressure vessel

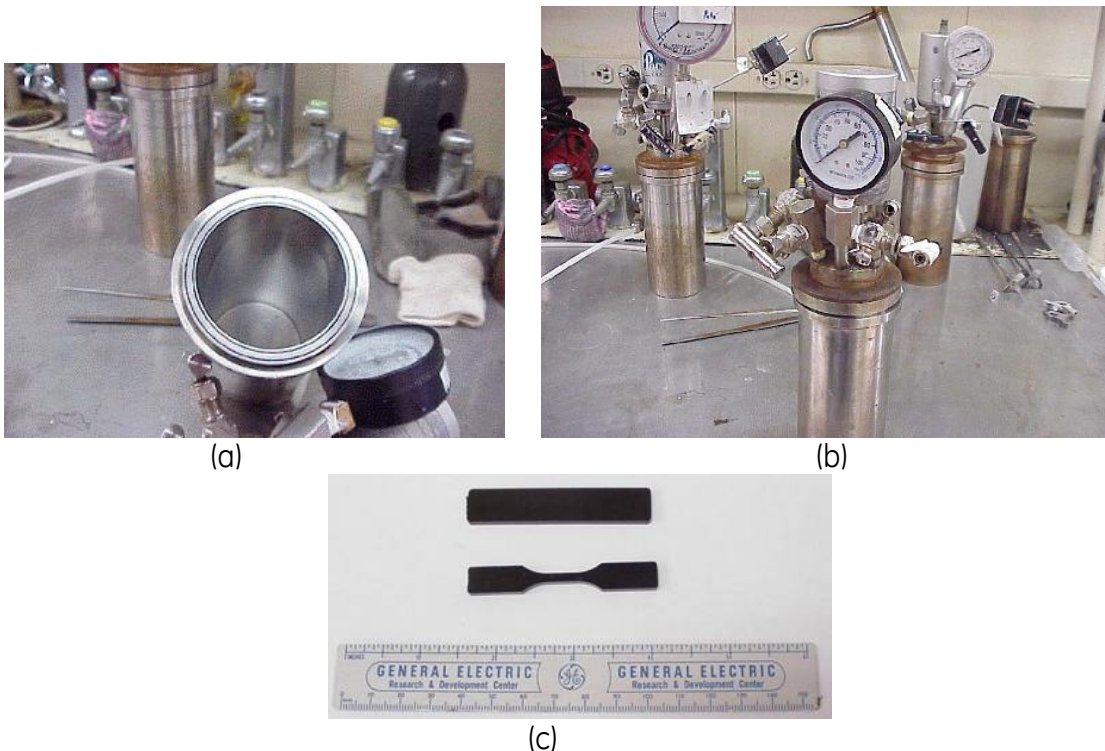
## 3. ACCELERATED LIFE AND STACK INTEGRITY TESTING

### 3.1 Plastics oxidation and KOH exposure tests

Oxidation tends to degrade the strength of plastic over time. Electrolysis produces high-pressure oxygen and other oxidative species such as ozone. We have performed accelerated testing of the plastic used in the electrolyzer to ensure that it will meet the goal of a 10-year stack life, which is required to keep amortized capital cost on target.

Two experiments comprise this task: accelerated aging of plastic in high pressure oxygen, and a second experiment comparing plastic exposed to concentrated hot KOH, oxygen and electrolysis products to plastic exposed to concentrated KOH and

oxygen alone, all at ambient pressure. In the Q12008, we concluded that the absence of an effect by the electrolysis products indicated that the experiment done in high-pressure oxygen alone is the best indicator of plastic life in a high-pressure electrolyzer. Therefore, KOH exposure tests were discontinued after the first quarter.



**Figure 6: (a,b) High pressure Parr vessel and (c) tensile and SEN specimens.**

Accelerated testing in high pressure oxygen and elevated temperature, which is currently equivalent to 179 weeks (3.44 years) at a design pressure and temperature of 15 bar and 80C, respectively, has now been completed on samples of Noryl® EN265 and a modified formulation of Noryl® EN265 with enhanced antioxidant package. Both Noryl® formulations show complete loss of yield strength in tensile samples, and both have lost all ductility and are now quite brittle. Testing of Single-Edge Notched (SEN) 3-point bend samples of the same materials after 179 equivalent weeks indicate a complete loss of both ductility and the yield strength. Images of the tensile and SEN samples as well as the Parr vessel used to accelerate the oxidation process are shown in Figures 6 (a,b,c).

Identical tests on tensile and SEN (3-point bend) samples of Udel® and Radel® specimens subjected to high pressure oxygen equivalent to 179 and 201 weeks (3.44 and 3.86 years), respectively, at a design pressure of 15 bar and 80C, indicate that both materials retain 100% of their initial yield strength in both stress states. Radel® specimens retain 100% of their initial ductility in both stress states, while Udel® samples retain 73% of their initial ductility in tension, while retaining 100% in bending.

GE is continuing this experiment, with samples being pulled for testing every 3 to 4 months. The high temperature oxygen exposure tests show that Noryl® will be sufficient for the prototype test plan of this project. However, in order to use Noryl in an electrolysis stack at a sustained pressure of 15 bar and temperature of 80C for periods longer than ca. 40 weeks, the surface must be protected from oxidation by a barrier layer.

A summary of these results are shown in Table 3.

**Table 3: Yield Strength and Ductility Retention in Tensile and Bending for Candidate Stack Materials Following Exposure to High Pressure High Temperature Oxygen**

3-Point Bend NI Material	Equiv Years 15 bar Exposure	%Yield Retained	%Energy Retained	%Extension Retained
Udel P1700	3.86	109.00	109.00	100.00
Radel R-5000	3.86	116.86	102.95	94.03
Noryl EN265	3.44	0.00	0.01	0.00
Noryl EN265 mod	3.44	0.00	0.02	0.00

Tensile Material	Equiv Years 15 bar Exposure	%Yield Retained	%Energy Retained	%Extension Retained
Udel P1700	3.44	101.71	73.60	77.47
Radel R-5000	3.44	97.71	122.27	78.22
Noryl EN265	3.44	1.20	0.00	0.00
Noryl EN265 mod	3.44	1.47	0.00	0.00

Legend: Green = acceptable, Yellow = marginal, Red = Unacceptable

It should be noted that the amorphous, glassy structure of the candidate plastic materials tends to inhibit gas diffusion more than some other polymers. However, long term, continuous operation of the electrolyzer will likely involve some migration of the gases through the plastic. The diffusion distance (X) as a function of time (t) for a given diffusion constant (D) is approximated as:

$$X = \sqrt{2Dt}$$

The diffusion coefficient for gases in polymers are on the order of 10e-6 to 10e-8 cm<sup>2</sup>/sec, depending on molecular size. The width of the full-size plastic plate is about one inch, which equates to a minimum gas diffusion time of about one year if operated continuously. Therefore, the annulus area between the plastic stack and the pressure vessel cylinder shell should periodically be monitored for traces of H<sub>2</sub> or O<sub>2</sub>.

### 3.2 Cell joining tests

A series of wedge breaking tests, as shown in 7(a), were conducted on small-scale, one-inch wide coupons as well as on sections cut from the full-sized plate. The initial

small-scale tests were conducted to finalize the weld process parameters and develop a transfer function for the current, voltage, and wire length needed to weld the full-sized plate. In these tests, it was confirmed that the weld was as strong as the plastic, as shown in Figure 7(b). Finite element analysis further substantiated the test results (Figure 8). The area in red indicates stresses above the yield strength of this material very near the wire when the plates are pushed apart 1/8" (the width of the wedge-breaking bar).

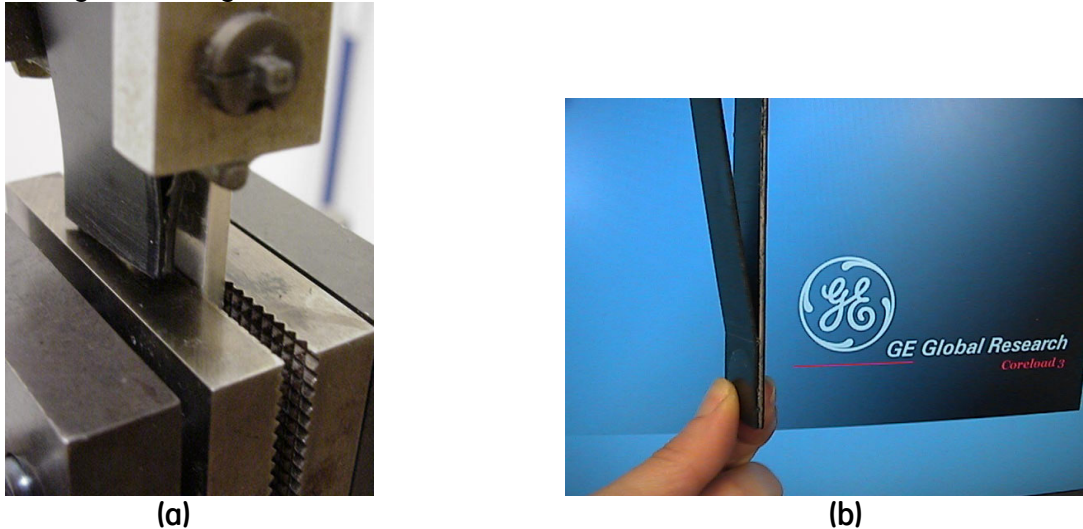


Figure 7: (a) Welded Noryl sample in wedge breaking test; (b) results after test.

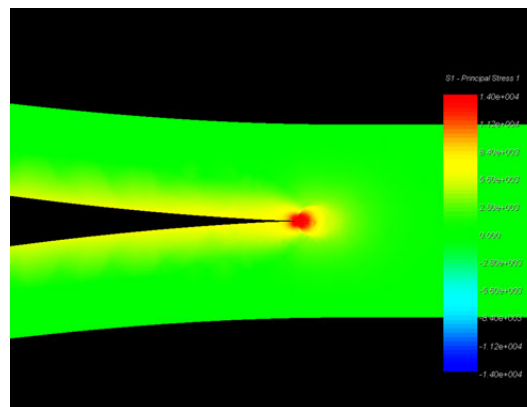


Figure 8: Weld specimen finite element analysis.

In addition to these wedge-breaking tests, a series of small-scale stack hydrotests were completed to determine the hermeticity of the welds. Five circular plates with an outer diameter and inner diameter of 5 inches and 3.5 inches, respectively were welded together. Each of the plates are the same thickness as the plates used on the full-sized stack.

These tests were conducted with three different axial support scenarios: a) welded plastic end plates b) bolted, compliant plywood endplate c) bolted, rigid metal endplate. Table 4 shows the maximum pressure achieved and length of time tested for each set of tests.

**Table 4: Effect of endplate constraint on small-scale stack holding pressure and time.**

Endplate description	Maximum pressure (psi)	Time at pressure
Welded plastic	15	5 min
Compliant, bolted endplate	180	10 min
Rigid metal endplate	350	48 hrs (test stopped)

Clearly, the endplate stiffness and axial constraint has a measurable effect on the strength of the weld. From these tests, we concluded that the life target of the plastic stack might only be achieved using a solid metal endplate designed for minimal deflection in order to reduce the probability of the peel effect compromising the weld integrity.

An additional small-scale pressure test showed that a pressure up to 100 psi could be achieved using the rigid metal endplate and without any welds keeping the plates together. From this, we concluded that the compliance of the plastic itself is sufficient to form a hermetic seal at lower pressures if the endplates are relatively rigid. A patent has been filed to capture this embodiment for lower pressure applications. This embodiment has the advantage of allowing replacement or inspection of individual diaphragm or electrode cartridges in addition to lowering assembly costs.

## **4. ELECTROLYZER PERFORMANCE**

### **4.1 Electrodeposition testing and performance results**

#### Bench scale tests

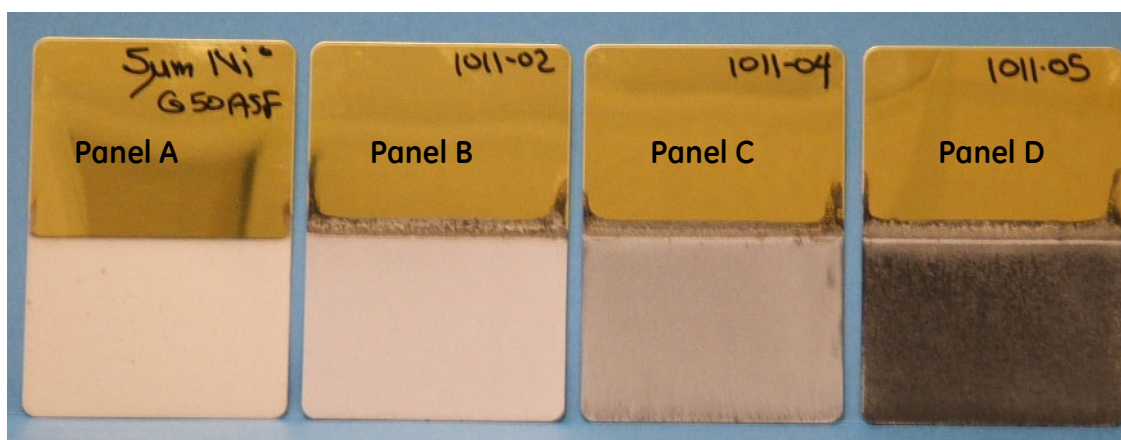
The ideal electrodeposition process required to achieve optimal performance on the full-sized, 10 cell stack was achieved by following a rigorous testing process. Numerous electrodeposition tests were conducted first on coupons, then proceeding to a bench-scale single cell device, then a full-size single cell device before proceeding to in-situ deposition on the full-sized stack. From this procedure, the performance was determined to be affected by the following variables. Specific tests used to evaluate these variables are also listed.

- Plating solution chemistry (coupons, bench-scale)
- Coating thickness (coupons, bench-scale)
- Adhesion (coupons, bench-scale, full-size single cell)
- Surface area (coupons, bench-scale)
- Coating effectiveness/coverage (bench-scale, full-size single cell)
- Flow rate and current density during deposition (bench-scale)
- Cleanliness of surface before coating (full-size single cell)

Results of these tests will now be discussed. The precise chemistry of the electrodeposition solution, as well determination of the optimal coating thickness and resulting adhesion were first tested using a series of brass coupons (Figure 9).

In the laboratory, two electroplating solutions were formulated—one to deposit pure nickel and another to deposit nickel-zinc. In order to understand how the presence of zinc ion affects the deposit when being co-deposited with nickel, the concentration of zinc ion was varied while holding the nickel ion concentration constant in the second formulation. Standard electrolytic, polished brass panels (10 cm x 7 cm) were used as substrates for a series of experiments where first a layer of about 5  $\mu\text{m}$  of pure nickel was deposited on half of the panel. The panel was then transferred, without rinsing or drying, into the second plating solution and plated for a time to achieve a thickness of a NiZn co-deposit.

Figure 9 shows a macroscopic view of the plated coupons. The nickel base layer (panel A) has a matte finish, and it is known from other work done with this particular plating formulation that the grain size of the deposit is on the order of about 5  $\mu\text{m}$  with a slightly rough, but level surface finish. When zinc ion is added into the plating formulation, and a layer of NiZn is deposited, the surface, while still level with a matte finish becomes slightly rougher (Panel B). When the  $\text{Zn}^{+2}$  level is increased, the deposit becomes rougher still, and under microscopic viewing shows the beginning of nodules comprising nickel and zinc (Panel C). When the metal layer from the second plating formulation is deposited in a greater thickness, the surface structures of the deposit show an increase in surface area, evident under microscopic viewing and by the fact that the macroscopic appearance of the coating is darker (Panel D).



**Figure 9. A series of brass panels plated on the lower half with varying amounts of Zn in a Ni-Zn deposit**

The surface area for reaction is further increased by chemically removing the zinc metal, leaving a pure, stable, porous nickel layer on top of the dense nickel layer to adhere to the supporting substrate.

After the balance between coating thickness and adhesion was achieved on the brass coupons, the procedure was transitioned to the bench-scale single cell test device for performance testing and in-situ deposition process development. This device, shown in Figure 10, has a cell area equal to about 160 cm<sup>2</sup>.



**Figure 10: Bench-scale single cell test device**

A performance comparison between a flat plate electrode and an electrode with a simple mesh bonded on were completed, as shown in Figure 10(b). Note that both the electrode and mesh materials are the same used on the full-sized demonstration stack.

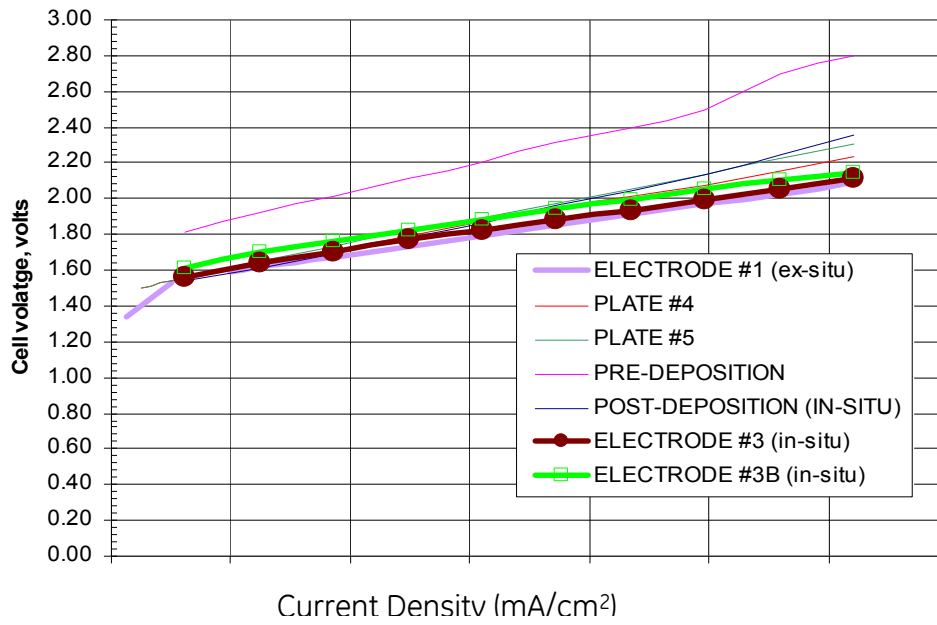
Both in-situ and ex-situ deposition techniques perform measurably better than an uncoated plate. However, only marginal improvements in performance have been achieved with either in-situ or ex-situ techniques with the mesh welded onto the bi-polar plated compared to the bi-polar plate alone, as shown in Figure 11(a).

Measurable improvements in performance appear to be feasible if a significant increase in Ni-Zn coating thickness is achieved. Figure 11(b) show results from an early set of tests in which a cell voltage 1.6 V was obtained. However, in this case, the electroplate thickness was enough to obstruct fluid flow between the electrode and diaphragm in addition to adhesion that is less than desired. Therefore, a balance between active surface area and fluid flow volume is desired during the deposition process.

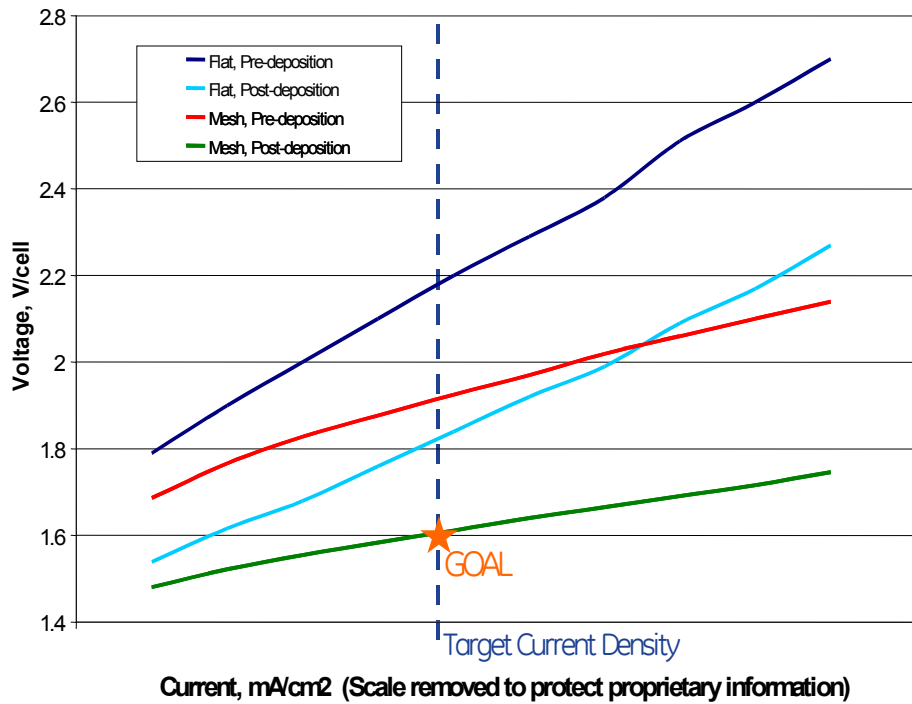
There are a number of risks associated with ex-situ deposition and handling of micron scale nickel-based surfaces. Given the fact that small size single cell experiments indicated no measurable performance difference between in-situ and ex-situ electrodes, it was therefore determined that deposition on the full-size electrodes should ideally occur in-situ. Work then proceeded on determination of the ideal flow rate and current density needed to achieve optimal performance and coating effectiveness.

Figure 12 illustrates the Ni-Zn deposition results for experiment #3B. While the coating adhered well in the darker areas, it is clear that the solution did not thoroughly coat the upper left corner. In these tests, the plating solution flowed from the lower left corner to upper right, opposite corner. The less optimal square shape likely resulted in more of the plating solution adhering nearest the flow path. It is also possible that the deposition chemicals may have been exhausted before reaching the upper corner. Interestingly, despite uneven coating distribution, the performance was not measurably affected, and as mentioned above, the adhesion was acceptable.

### Alkaline Electrolyzer Cell Tests



(a)



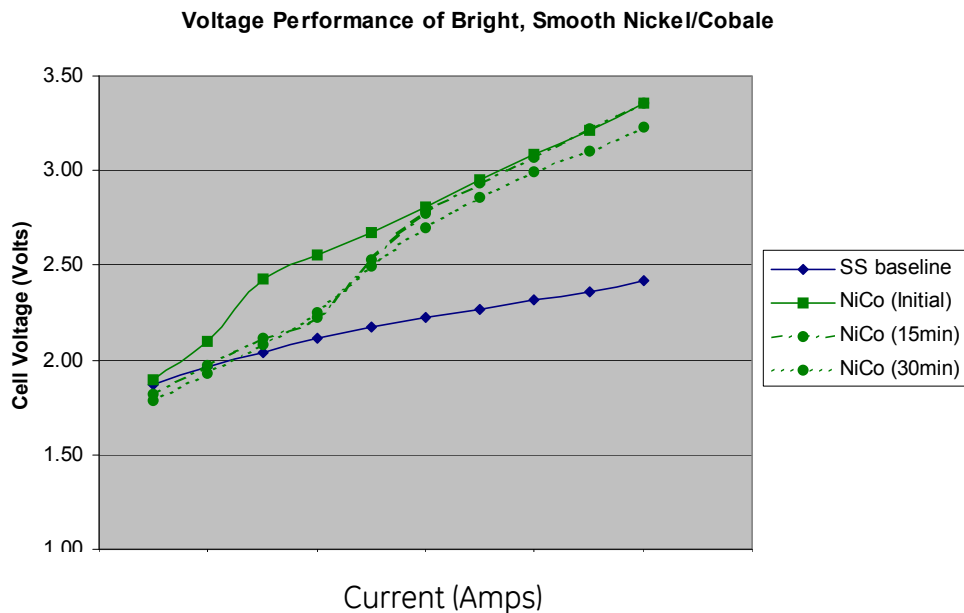
(b)

Figure 11: Single cell bench scale tests showing (a) effect of in-situ vs ex-situ deposition on performance and (b) effect of mesh vs flat plate on performance.



**Figure 12: (a) Image of bench-scale experiment #3B showing coating effectiveness and (b) single-cell device illustrating orientation of flow passages.**

At this point in the testing, the performance appeared to reach a “plateau” despite efforts to increase surface area and improve coating distribution. Therefore, two alternate plating chemistries, nickel-cobalt and nickel, respectively, were tested to compare with Ni-Zn. Extensive experience with Ni-Co plating processes indicated some promise for achieving improved performance using this chemistry. Furthermore, literature states that a smooth Ni-Co surface should catalyze the hydrolysis of water as efficiently as the etched Nickel/Zinc. The Ni-Co cathode was plated ex-situ, then assembled in the bench scale test cell device. Performance tests of this cell are shown in Figure 13.



**Figure 13: Performance results of Ni-Co electroplating experiments conducted on bench-scale test device.**

While the shiny (i.e. low surface area) Nickel/Cobalt performed worse than the stainless steel baseline, the plating coverage was excellent. Achieving a high surface area coating is clearly critical to electrode performance. Additional research considered out of scope for the current project would be needed to determine the appropriate Nickel/Cobalt plating conditions to provide surface characteristics similar to that of Nickel/Zinc.

A test to determine the effectiveness of plating with nickel alone was conducted to potentially eliminate the need to chemically etch the Zinc from the Nickel/Zinc surface. No performance improvement compared to the baseline stainless steel plate was noted, however surface coverage was acceptable.

#### Full-sized single cell tests

A series of deposition experiments were conducted on a full-sized single cell test rig as a final risk mitigation step prior to plating on the prototype 10-cell stack. Figures 14 and 15 are pictures of the electrode surface after plating in the full-size single cell deposition rig for the first and second set of experiments, respectively. Improved coating adhesion was achieved in the second plating experiment by in-situ soaking of the electrode in a mild nitric acid solution prior to plating. This mild nitric acid solution was determined to have no negative effect on the diaphragm integrity.

Voltage taps were placed on either side of the full-size cell during plating to insure that we have an adequate power supply to electrodeposit all cells simultaneously in the 10 cell prototype electrolyzer module.

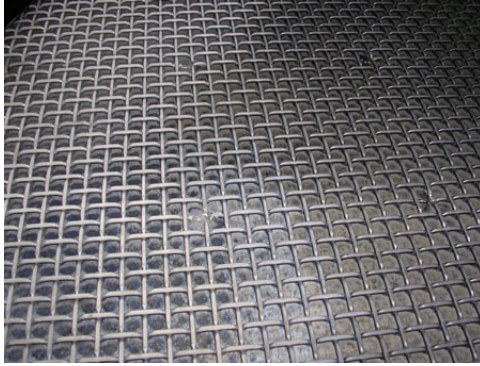


(a)



(b)

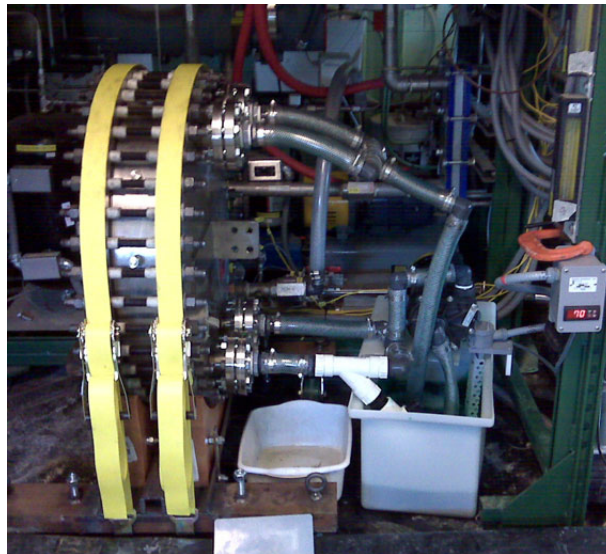
**Figure 14: (a) Image showing localized peeling of electrodeposited surface; (b) image showing uniform coating of electrode surface.**



**Figure 15: (a) Image showing well-adhered electrode coating after cleaning with nitric acid; (b) uniform coating of electrode surface for this experiment using the full-size deposition rig.**

## 4.2 Full-sized 10-cell stack performance

The in-site deposition process described above was used to apply the active surface to the full-size, 10 cell stack (Figure 16). Applying the coating this way has many benefits. There is no handling of the coated electrode, which maintains the integrity of the coating and is safer for personnel. The process also conserves valuable floor space in a manufacturing environment and, since all plates are coated simultaneously, is quicker and cheaper than plating each electrode ex-situ.

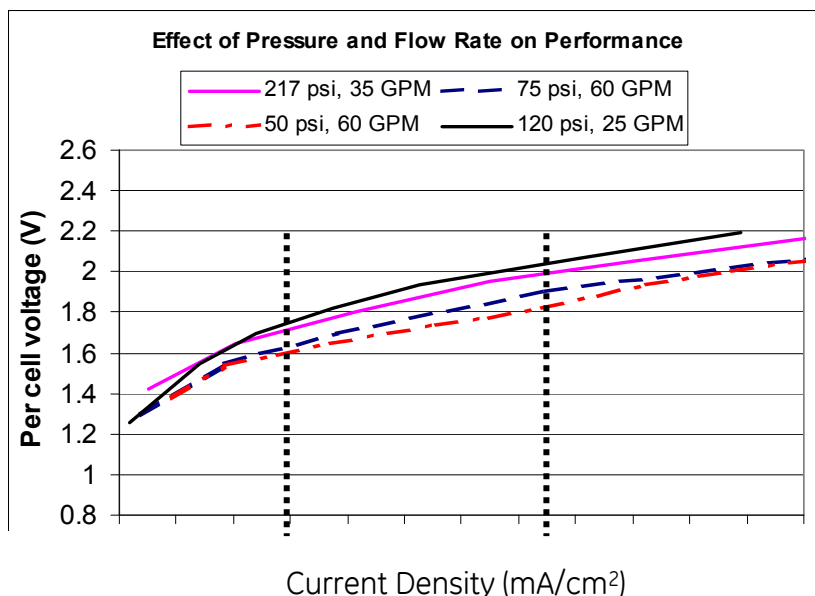


**Figure 16: Image showing in-situ electrodeposition process on 10-cell prototype electrolyzer stack.**

After the coating was applied, a series of tests were completed to determine the performance of the stack at different electrolyte temperatures, flow-rates, and applied current. The results are also shown for a range of pressures, since the performance of the stack at high pressure appeared to be limited by the ability of the system to remain stable at lower flow rates necessary to achieve full separation of the gases.

The effect of pressure and flow rate on performance is illustrated in Figure 17. From this plot, it appears that the performance decreases as the flow rate is reduced from 60 GPM at test pressures between 50 and 75 psi down to 25 GPM at a pressure of 120 psi. All tests were conducted at 80C.

Lowering the flow rate tends to increase the tendency for bubble masking of the electrode surface, thereby decreasing the surface area for reaction with the electrolyte. However, Installation of the mass spectrometer indicated a propensity for gas crossover at the higher flow rates. This is one indicator of gas entrainment most likely due to inadequate gas separation. The faster the electrolyte moves through the separation tanks, the less time the gases have to percolate out of the fluid. If the gas crossover is significant enough, a potential safety problem occurs due to mixing of hazardous levels of H<sub>2</sub> and O<sub>2</sub> in the separation tanks. Perhaps not surprisingly, the optimal performance for the stack was achieved at pressures between 50 and 75 psi at a flow rate of 60 GPM, which are the conditions for which the system and separation tanks were sized.



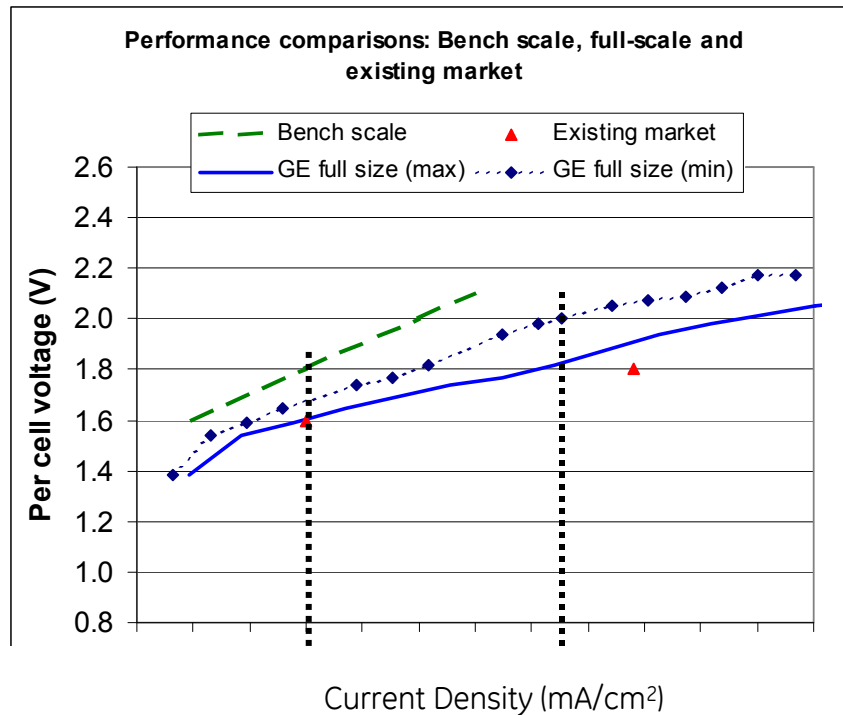
**Figure 17: Effect of pressure and flow rate on performance for the full-sized prototype stack**

At the lower current density, the cell voltage varied from 1.6V to about 1.7V. The maximum efficiency for this set of tests was 72.3% or 45.6 kWh/kg (LHV). It is more likely that an electrolyzer used for production would operate at a current density that is closer to the higher target shown. At this level, the cell voltage of 1.8V provided an efficiency of 68.3% (LHV) with a corresponding H<sub>2</sub> production rate of 0.26 Kgph.

Figure 18 compares performance obtained on the single cell bench-scale test device and GE's full-size 10-cell prototype stack in comparison with electrolyzers currently available on the market. Note that the data on this chart for existing electrolyzers was obtained from non-proprietary discussions with electrolyzer makers. Interestingly, the performance quotes obtained in these discussions appear to be somewhat higher than the performance estimates obtained by NREL (see Table 2). More discussion between GE's stack performance and costs compared to market data is included in Section 7.

The performance of the 10-cell stack (plotted as "GE full-size (max)" in Figure 18) is significantly better than the performance achieved on the bench scale at any current density, which is mostly attributed to the circular shape of the full-size electrode compared to the square shape of the bench scale electrode. Shunt current calculations completed prior to the design of the manifolds indicated very low losses due to the small gap between cells. However, in the unlikely event that the shunt current losses were as high as 30% (plotted as "GE full-size (min)" in Figure 18) the performance of the 10 cell stack is still superior to the bench scale. This further substantiates the conclusion that cell shape is a bigger factor in rationalizing the bench-scale versus full-scale improvement.

The performance of GE's prototype electrolyzer meets that of existing electrolyzers at the lower target current density, which translates to an efficiency of 72.3% or 45.6 kWh/kg (LHV).



**Figure 18: Comparison of GE full-size and bench scale stack performance with existing market.**

## 5. ELECTROLYZER STACK CAPITAL COSTS

Capital costs for an electrolyzer stack using the technology developed in the program may be estimated based on the cost to construct the prototype 10-cell pressurized plastic stack.

### Plastic

To save tooling costs, the plastic stack core for the 10-cell stack was constructed from injection molded blank plates into which flow passages and other features were machined. In full production, dedicated tooling would be used to create the same part in one piece. Therefore, fewer welds and less assembly labor will be required.

Molded plastic costs are estimated to be roughly double the cost of the plastic raw material. For our 10-cell stack with approximately 2500 cm<sup>2</sup> of active area, less than 1.2 kg of plastic is required to make each cell. High-grade industrial plastics such as Udel™ and Radel™ are available at approximately \$12-\$13 per kg. However, molding and processing costs will increase the cost of the finished plastic stack parts by two times the raw material costs or more, depending on labor rates.<sup>9</sup>

### Electrodes

The electrodes in the GE 10-cell stack are made from wire mesh welded to stainless steel plates. A high-effective area nickel based surface is applied to the stainless steel substrate using GE's electrodeposition process. Specifics of the

deposition method are the subject of pending GE patent applications. Since electrodeposition is a very low-stress process, the electrodes may be made of thin section stamped metal to further reduce costs.

### Pressure Vessel

For a pressurized stack, the metal pressure vessel will be a significant cost. Most of the mass and expense of a pressure vessel is in the end covers, and additional length can be added to a vessel at low cost. Therefore it is advantageous to make the electrolyzer stacks as long as possible and minimize the number of pressure vessels. The length of the stack may be limited by the available voltage of the power supply, flow distribution concerns, and finally production limits on the pressure vessel itself. In practice, alkaline electrolyzer stacks of several hundred cells and pressure vessels several meters long are not uncommon. The pressure vessel for the GE 10-cell stack cost approximately \$28,000. We expect that in volume production the pressure vessel cost for a stack with similar cell area will be \$15000-\$20000.<sup>10</sup> A 400-cell stack of the same design as the 10-cell will fit in a pressure vessel approximately 13 feet (4m) long. The current entitlement design, with minor modifications, can manage 200 cells, with a length of about four foot long. Longer stacks may be accommodated by incorporating thermal expansion features included in patented designs.

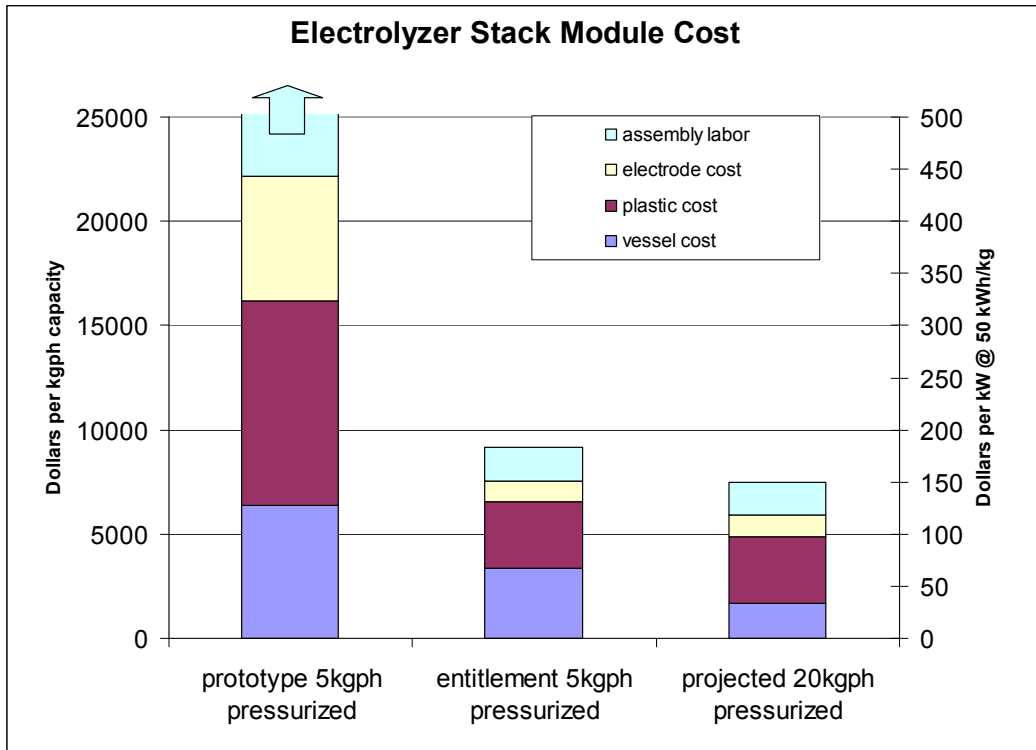
Note that the pressure vessel is re-usable, and capable of achieving a 20 yr lifetime, and perhaps longer, which is common for most pressure vessels. Therefore, capital costs and amortization schedules for the stack could assume a 10 yr life for the plastic stack to be installed in a re-usable pressure vessel that can be scaled up or down as capacity requirements dictate.

### Assembly

Assembly labor is required to join the plastic cartridges together to form an electrolyzer stack. For the GE 10-cell stack, as described previously, individual molded plates were joined to form the electrode and diaphragm cartridges. A full production design would incorporate several of these parts into one mold. Specific techniques to join the cells in a high volume process are the subject of pending GE patent applications.

### Cost Estimate Based On 10-Cell Demonstration Stack

A summary of the projected electrolyzer stack module costs based on the construction of the 10-cell demonstration stack is shown in Figure 19. Labor costs in Figure 19 are assumed to be \$40 per hour. This cost is based on semi-skilled, US factory work labor, fully-burdened overhead and benefits add-ons.<sup>11</sup>



**Figure 19: Electrolyzer Stack Cost**

The costs for a prototype 5 kilogram per hour stack module are based on a direct scaleup of the cost to produce the 10-cell demonstration stack. This stack module is assumed to be 200 cells in a single pressure vessel. Labor costs are high due to the non-automated plastic joining method, which involved 40 hours of labor per cell. Plastic costs are based on the costs for the small production parts used in the demonstration stack. The pressure vessel is rated to 15 bar.

Entitlement costs for the same pressure vessel are estimated in the next column. The pressure vessel cost is reduced by high volume production. The cost of plastic is estimated at an entitlement cost of two times the raw material cost for high-grade industrial plastic, plus a setup cost per cell for the automated joining process. Electrode costs assume a high volume process for stamped metal. Labor costs in the automated joining process are shown assuming one hour of work per cell. This includes final plastic part preparation and stacking. Labor costs may be lower in large scale production, and this represents an opportunity to drive stack module costs down further.

Finally, costs are projected for a 20 kilogram per hour stack module. This stack of 800 cells is in two pressure vessels. Labor and material costs are otherwise the same as for the 5 kg per hour case.

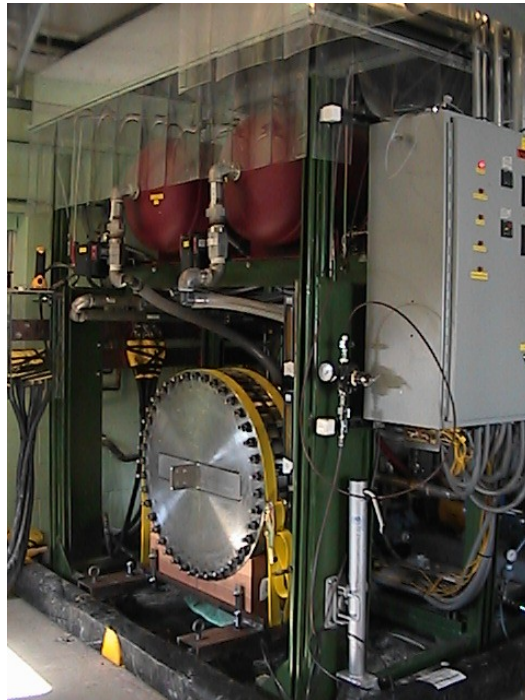
These assessments include only the stack module itself. The complete system will include circulating pumps, separator tanks, the power rectifier, and other control and process hardware. Balance of system costs will range from a few hundred dollars per kW for large systems to thousands of dollars per kW for smaller systems of approximately 5 kg per hour. System cost estimates are included in Section 7.

## **6. PROTOTYPE SYSTEM TESTING (1 KGPH)**

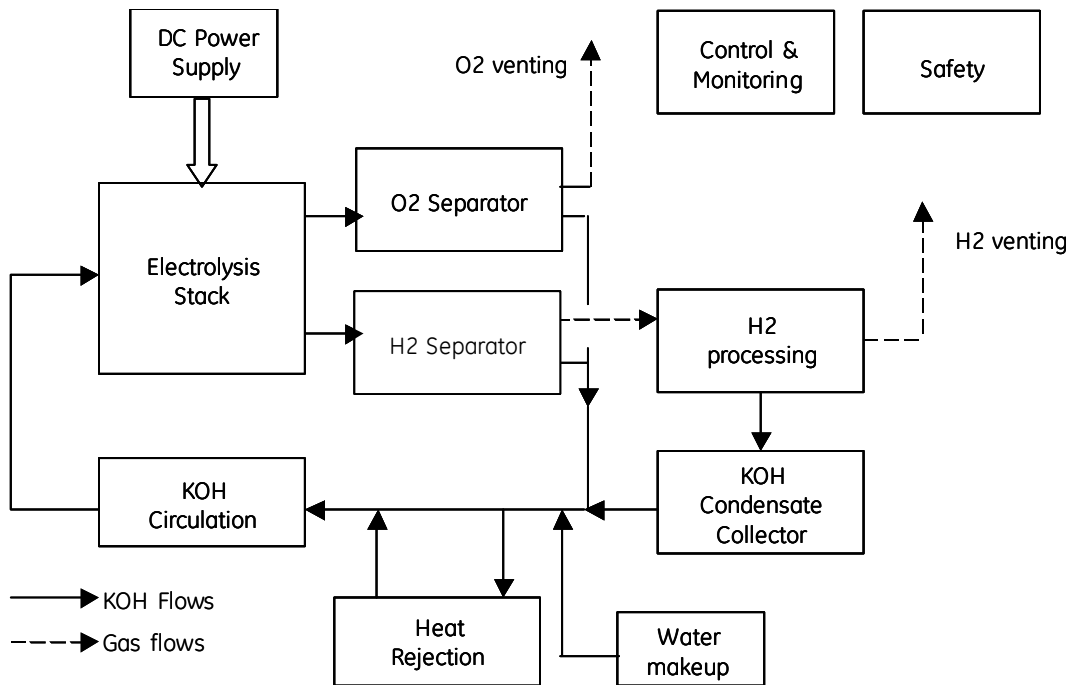
### **6.1 Operation and maintenance history**

An electrolyzer system capable of producing 1 kgph at ambient pressure was designed and built during a 2005 DoE program.

This system has been upgraded to generate hydrogen at 15 bar pressure using the prototype stack (Figure 20) developed as part of the current program. A simple schematic of the system is included here for clarity (Figure 21).



**Figure 20: Full-sized prototype stack and 1 kgph capable electrolyzer system**



**Figure 21: Electrolyzer System Schematic**

The system had been sitting idle since the last set of tests were run, which was about 1 or 2 years. Therefore, a complete inspection of the system and re-commissioning of the electrical system, all mechanical systems, and controls were conducted to identify items needing replacement or refurbishment.

Some evidence of corrosion on exposed steel supports was evident. The stainless piping and components (a few of which are mentioned below) were tested to insure safe operation at their rated pressures. The system was degreased with “Simple Green<sup>®</sup>” prior to operating at elevated pressure to mitigate risks due to high pressure O<sub>2</sub>.

The following is a list of components that were replaced prior to taking the system up to 15 bar pressure.

- Back pressure regulators
- Heat exchanger
- Liquid level sensors
- O<sub>2</sub> sensor and sensor cartridge
- All of the 1 ½” pipe fittings
- Pressure transducers
- Pressure relief valves
- N<sub>2</sub> purge inlet solenoids
- Pressure relief spring

The electrolyzer system was first commissioned up to a pressure of 50 psi to gain experience and troubleshoot any system related problems that may arise prior to operation at 15 bar pressure. After equipment replacement and additional upgrades were completed, the system was leak checked with water and N<sub>2</sub> at 230 psi and room temperature. Note that all leak checks were done without the stack in the system. A set of by-pass pipes were put in place during these checks. No leaks in any of the equipment were detected, and so system testing was allowed to proceed.

Ultimately, the reliability of the heat exchanger, level sensors, and gas sensors posed challenges that prevented continuous operation of the stack. Issues with the level sensors and gas sensors were successfully resolved through minor design and installation changes, which will be discussed shortly. Failure of the heat exchangers likely occurred to the high-wettability and corrosive nature of KOH, even though they were supposedly designed to manage KOH at the pressures and temperatures we operated. System design suggestions to increase reliability and safety or reduce costs have been highlighted in bold.

#### *Heat exchanger:*

During the first set of tests, the heat exchanger developed a significant leak during operation. Also, the low liquid level sensors in the separation tanks were giving false low readings, both of which prevented operation at that time.

The heat exchanger had been rated for 150-psi operation at 230°F for KOH, and had been leak checked to 50 psi in water prior to full testing, so it was unclear why it failed. A post-mortem inspection of the heat exchanger indicated the cause of leaking to be gasket failure between two of the plates. This heat exchanger had sat idle in the system for 1-2 years, so it is likely the gaskets suffered from dry rot or from long term exposure to KOH. The replacement heat exchanger, rated for 229 psi at 200°F in KOH, was on hand and available for immediate installation. After installation of the new heat exchanger, the system was leak checked and run continuously without further incident at pressures less than 150 psi.

However, the new heat exchanger ultimately failed after testing for about an hour at 15 bar (217 psi) and 80C (176 F), which is less than the rated value. It was determined that the cooling system would not be needed for this set of tests, so the valves going to the heat exchanger were shut-off, and testing at elevated pressure continued.

**System design improvements should include a cooling system that is not exposed to KOH.** This may be achieved by convection via fans that blow cooler air over the surface of the piping. Any evaporative cooling methods, such as a cooling tower, may not be feasible because of the propensity of KOH to turn to potassium carbonate when exposed to air.

#### *Gas sensor operation:*

System commissioning steps taken before operation indicated that the O<sub>2</sub> sensor cartridge needed replacement after having sat idle in the system for an extended

period. Typically the sensor would only see  $N_2$  during start up and shut down and very pure  $H_2$  during operation. If the system is not operating for more than a few days, the sensor should be removed and placed under  $N_2$ . Furthermore, any exposure to air should be followed by a re-calibration prior to operation to insure proper functioning.

After testing for a few days, the entire  $O_2$  sensor unit module failed and needed to be replaced. The cause of failure was later determined to be caused by KOH vapor in the gas path. This unit was originally installed after the backpressure regulator, which means that there was no indication of the combustibility of the gases in the separator tanks until the gases were being released. In addition, there was no way to determine if the system experienced gas-crossover since there were no  $H_2$  sensors on the  $O_2$  side of the gas system.

Before operation at pressures greater than 50 psi, the  $O_2$  sensor was moved before the backpressure regulator to facilitate measurement out of the  $H_2$  tank in advance of a gas release. A mass spectrometer was also added to detect the amount of  $H_2$  in the  $O_2$  tank and as a back up detection method for the existing  $O_2$  sensor. The piping to the mass spectrometer passed through an ice-bath to condense out any KOH vapor that might have accumulated in the system and to protect the mass spec. The  $O_2$  sensor was tied into the emergency safety stop for the building, and would shut down hydrogen production if the  $O_2$  levels went higher than 50% of the combustible gas limit. The combustible gas limit for  $O_2$  in  $H_2$  is 4% both at ambient and elevated pressure, so the E-stop was set to trip at 2%. **All piping going to sensors should include a condensate trap or heater to eliminate water or KOH vapor contamination.**

**It is strongly suggested that any system that is operated long-term be fitted with permanent  $H_2$  sensor controls** that are tied into the E-stop system, similar to the  $O_2$  sensors.  $H_2$  sensors are long-lead time items, and so the mass spec was used as a temporary solution in order to continue testing.

#### *Level sensor function*

The level sensors experienced problems during the first set of tests at pressures less than 50 psi. Proper level sensor function was restored after the rectifier was grounded at the negative terminal.

These same sensors seemed to develop periodic problems that necessitated numerous removal and re-installation steps. It is difficult to maintain a leak-tight seal necessary to operate at high pressures without inadvertently over-tightening the sensors, making them inoperable.

If the level sensors trip, signal is lost, or the backpressure regulators fail, the system controls are designed to close the valves between the electrolyzer outlet and separation tanks. However, the time necessary to fully shut the flow may not be sufficient to prevent KOH from escaping into the gas lines and ultimately, out of the

building, causing a safety issue for surrounding personnel. **Long term or unmonitored operation should include a feature for capturing any possible KOH escaping out of the gas lines.** We added a plastic tank at the O<sub>2</sub> gas condensate catch tank to manage this risk.

#### *Backpressure regulators:*

Backpressure regulators are optimized to function for a range of pressures and flow-rates. Ideally, the gas should be released fairly continuously and in small bursts once the pressure in the separation tank is achieved. This is needed both for accurate H<sub>2</sub> production readings as well as for safety. High velocity O<sub>2</sub> (greater than about 25 m/s) can be a significant safety risk.

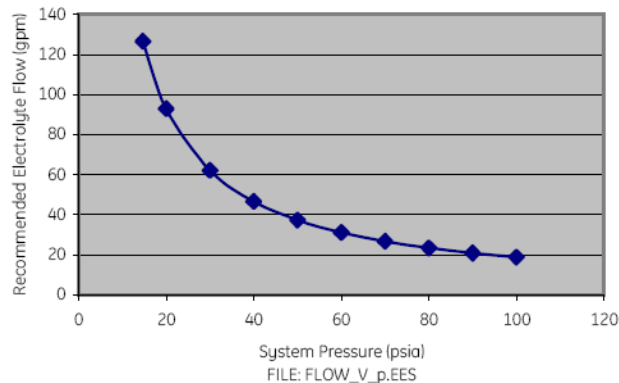
The backpressure regulator initially installed into the system was designed for low pressure operation (<50 psi). While they were capable of holding higher pressures, the release rate was not stable and exceeded the H<sub>2</sub> flow sensor capability. For these tests, the H<sub>2</sub> production rate was estimated using the ideal gas law ( $PV = nRT$ ), and noting the pressure change in the separation tanks during the release time. This assumes that the volume of the gas in the tanks remain constant.

A backpressure regulator capable of managing 15 bar pressure and release rates as high as 20 l/min was installed before operation at elevated pressure. This backpressure regulator performed reasonably well at this pressure, and no operation problems were experienced during testing.

#### *Separation tanks*

The separation tanks showed no indication of long term damage due to KOH exposure. The system commissioning step appeared to be successful in clearing the tanks and piping of any rust or other contaminants that may have accumulated during the 1-2 year period the system was not operating.

The design completed in 2005 assumed a maximum pressure less than 100 psi for optimal separation in the tanks that were installed. Calculations for operation at higher pressure using the existing separation tanks indicated we could achieve adequate separation if the flow rate were kept below 30 gpm or so (Figure 22). As mentioned previously, it appears that full separation of the gases was not achieved. Setting the flow rate below 30 gpm appeared to reduce the gas crossover and improve gas separation, however, the system could not be kept stable at the required flow rate for a period long enough to safely conduct a hydrogen production test. At the lower flow rates, the heater would accumulate steam and create a vapor lock. This condition quickly leads to a pressure imbalance in the stack and system, driving the fluid levels in the tank to go out of balance. This condition creates additional stress on the pump as well. After this last series of tests, the KOH pump failed as well. Since the system does not allow for the flow rate to be lowered below about 30 gpm, the only other solution would be to increase the size of the separation tanks.



**Figure 22: Plot of system flow rate (gpm) versus system pressure (psi)**

The existing design has a liquid surface area of  $0.5854 \text{ m}^2$  ( $6.4 \text{ ft}^2$ ) assuming a bubble radius of  $0.1 \text{ mm}$  (2 mils). If the bubble size at 15 bar (217 psi) is roughly half the bubble size at 100 psi, the required surface area for separation is  $1.57 \text{ m}^2$  ( $16.9 \text{ ft}^2$ ).

**Operation at higher pressure using this system design should be conducted using separation tanks that have a liquid surface area approximately two to three times larger than the existing design.**

#### *Stack pressure monitoring*

A pressure gauge was installed on the perimeter of the electrolyzer stack module to monitor the pressure of the fluid inserted between the stack and pressure vessel cylinder. The pressure did not exceed 50 psi during any of the tests, which seemed to indicate that the welds were holding at the higher pressures. Otherwise, the pressure would likely have risen to the same pressure as the stack internals. **Ideally, the pressure of the fluid in this space should be tied to the system** to insure that the internal and external pressure of the stack are in equilibrium. This provides the maximum protection against creep and weld failure. An additional sensor to detect  $\text{H}_2$  or  $\text{O}_2$  may also be added to detect leaks in the welds or long-term migration of gas through the plastic itself.

#### Maintenance procedure for running long term

Any surfaces that get contaminated with liquid KOH need to be cleaned with DI water to prevent the formation of potassium carbonate crystals.

All safety systems should be checked and calibrated periodically.

All gas sensors, temperature sensors, and flow meters should be calibrated every 6 months.

## 6.1 Suggested system design modifications for operating at ambient and elevated pressure (15 bar)

Based on the operation and maintenance history of the system, the following design changes are proposed for long-term operation of the electrolyzer system.

- System design improvements should include a cooling and pumping system that is not exposed to KOH.
- Sensors that detect combustible levels of H<sub>2</sub> in the O<sub>2</sub> tank and O<sub>2</sub> in the H<sub>2</sub> tank should be a permanent feature, and tied into a building's emergency stop controls.
- Long term or unmonitored operation should include a feature for capturing any possible KOH escaping out of the gas lines should the liquid level sensors or backpressure regulators fail.
- All piping going to sensors should include a condensate trap or heater to eliminate water or KOH vapor contamination.
- Electrolyzer systems for 15 bar production should be able to operate at a low flow rate (< 20 gpm) or separation tanks with a liquid surface area approximately 2X-3X times larger than the existing design.
- The pressure of the fluid maintained in the space between the plastic stack and the pressure vessel shell should be tied to the system to insure the internal and external pressure of the plastic stack are in equilibrium.

## 7. MEDIUM AND LARGE SCALE SYSTEM CONCEPTUAL DESIGN AND COSTS

The operation and maintenance history as well as the costs of the electrolyzer currently operating at Cooper Nuclear Station, in Nebraska, and electrolyzer formerly operating at the Pilgrim Nuclear Station, near Plymouth Ma, have been thoroughly discussed in past quarterly reports, and so will not be discussed in detail here.

### *System costs*

Tables 5, 6, and 7 list the capital costs for the stack and balance of plant (BOP) for systems capable of hydrogen delivery at 1 kgps, 1000 kg/day, and 100 kg/day, respectively. The 1 kgps capacity would be a commercial scale industrial facility, and the 1000 kg/day and 100 kg/day providing the upper and lower capacity bounds for the medium range industrial capacity facility (refer to Figure 1). The costs for the stack are based on a per cell cost for a fully realized design and automation facility (about \$10,000/kgph), and assuming that each pressure vessel will hold a minimum of 200 cells. The stack cost is based on the number of cells needed to make the desired capacity, and so tends to increase if the stack is operated at lower current densities. The balance of plant cost is based strictly on the energy required to produce the desired capacity, and so tends to increase with required current density. The efficiency of the GE electrolyzer stack has been factored into the costs as well.

Experience building and testing the 1kgph system at GE indicates a system capital cost for the lower capacities to be on the order of \$800/KW (not including the cost of the stack). A significant portion of this cost is in the AC/DC rectifier and other power electronics, which is estimated to cost between \$150-\$200/KW. Recent discussions with NREL <sup>12</sup> indicate rectifier costs to be between \$100-\$140/KW assuming volume pricing, so the 1 kgps capacity system costs were calculated assuming a BOP equal to \$500/KW. Official results of NREL's most recent survey of electrolysis manufacturers were not available at the time this report was written.

The optimal operating current for the facility would be the one that had the best balance of stack cost and BOP for the desired capacity. In each case, the lowest overall cost for a given capacity is at 1000 amps, operating at an efficiency of at least 68% (LHV) or better.

**Table 5: Capital costs for 1 kgps electrolyzer system. BOP = \$500/KW**

Applied amps	# cells	Total KW	Stack cost		Total cost (K)
			(K)	BOP cost (K)	
500	257093	212102	\$60,931	\$106,051	\$166,982
1000	141791	258059	\$33,604	\$129,030	\$162,634
1500	103876	311628	\$24,619	\$155,814	\$180,433

**Table 6: Capital costs for 1000 kg/day electrolyzer system. BOP = \$800/KW**

Applied Amps	# cells	Total KW	Stack cost		Total cost (K)
			(K)	BOP cost (K)	
500	2828	2333	\$670	\$1,866	\$2,537
1000	1560	2839	\$370	\$2,271	\$2,641
1500	1143	3428	\$271	\$2,742	\$3,013

**Table 7: Capital costs for 100 kg/day electrolyzer system. BOP = \$800/KW**

Applied Amps	# cells	Total KW	Stack cost		Total cost (K)
			(K)	BOP cost (K)	
500	283	233	\$67	\$187	\$254
1000	156	284	\$37	\$227	\$264
1500	114	343	\$27	\$274	\$301

Comparison between data provided by NREL (See Table 2) for existing electrolyzers on the market indicate that the efficiency for a given power input (KW) exceeds the performance of the Stuart IMET 1000 (now Hydrogenics, Toronto, ON) and the Teledyne EC750 in the 100 kg/day capacity range and meets the performance of the Norsk Hydro Type 5040 in the 1000 kg/day capacity range. Information regarding costs of these systems was not available for comparison.

#### *Conceptual design*

As a general rule, power conversion losses are minimized at higher voltages. Therefore it will be desirable for a single rectifier to provide power to enough electrolyzer cells to total at least 1000 VDC. At approximately 1.8 volts per cell, this

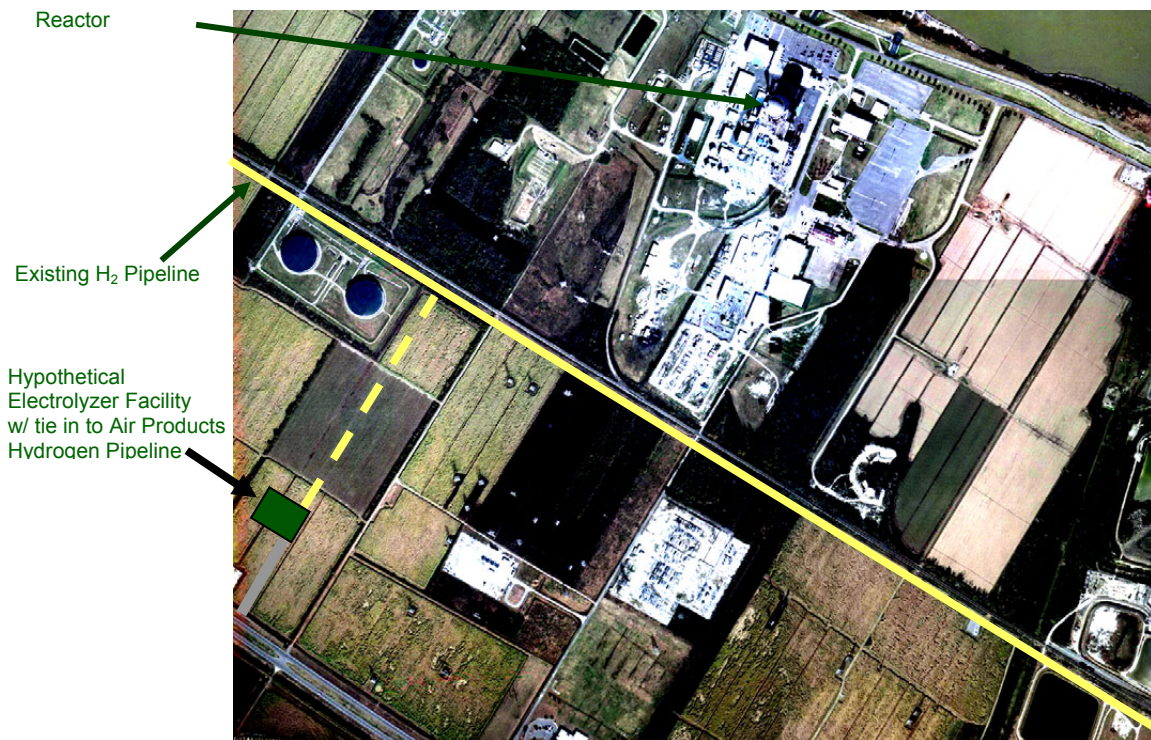
suggests electrolyzer power blocks each serving more than five hundred cells. Due to manifold losses, practical stack modules may be limited to 100-200 cells each. Therefore the optimal arrangement of stack modules will be several 100-200 cell modules powered from the same rectifier in electrical series. Based on the tables above, a facility requiring 1 kgps would need about 710 electrolyzer modules with 200 cells each.

For the purposes of the feasibility study we are considering the case of power blocks consisting of four 200 cell stack modules connected in electrical series to the same power rectifier. Each stack module is sized to draw a maximum of 1500 amps and operate at a cell voltage of 1.8V, requiring 540 kW per module or 2160 kW per power block. The rectifier for each power block would provide 1500 A at 1440 VDC. Approximately 178 such power blocks would be required to produce the total of 1 kilogram of hydrogen per second.

At a production rate of approximately 48 kWh per kg of hydrogen, the electrolyzer required to produce a kilogram per second of hydrogen would consume approximately 174 MW of electric power. At a rate of 9.2 liters of water consumed per kg of hydrogen produced, approximately 33 metric tons or 7000 gallons of water per hour would also be required.

At least half an acre of the facility would be required to house the electrolyzers alone. The liquid surface area for the separation tanks requires about 17 ft<sup>2</sup> per electrolyzer for operation at 15 bar pressure. This is about the same footprint as the 200 cell electrolyzer. Floor space can therefore be conserved if the separation tanks are mounted directly over each electrolyzer. The piping, pumps, heat exchangers, rectifiers, water processing, and other large equipment would occupy a space many times that of the module itself. Entergy has estimated this facility would occupy approximately seven acres of land.

It is anticipated that a facility of this size and complexity will need routine and unobstructed access for operations and maintenance. Figure 23 is an aerial view of the Waterford 3 site with a proposed location for the electrolyzer facility depicted in green. The size of the green rectangle approximates a land area of 7 acres. The distance from the facility to the Waterford 3 reactor building is just over one mile.



**Figure 23: Aerial View of Waterford 3**

The solid yellow line in this figure indicates an existing Air Products hydrogen pipeline that traverses the Waterford property. This pipeline is 12  $\frac{3}{4}$  inches in diameter and is buried at a minimum depth of 5 ft while on Waterford property. The dashed yellow line in the figure represents a hypothetical tie-in to the existing hydrogen pipeline. Current hydrogen flow rate through the existing line is estimated at approximately 37 million SCF/day (87,470 kg/day or 1.01 kg/sec). The proposed electrolyzer facility's hydrogen production rate of 1 kg / sec would roughly double the flow rate in the existing pipeline should the entire output of the facility be injected. Estimates also suggest that a small increase in pressure (<50PSI) in the existing pipeline will result from the added injection. Although more rigorous analyses would certainly be required before actually giving serious consideration to such a project, the cursory assessment suggests that this conceptual tie-in arrangement would be plausible. (No judgments were made regarding the downstream capability to receive the additional product or to handle the modest increase in pressure.)

#### *Regulatory Considerations*

An injection rate of 1 kg H<sub>2</sub>/ sec will result in transient storage in the tie-in line. However, it is possible that the actual quantity of hydrogen stored at such a facility might be relatively small since the output is injected directly into the existing pipeline. There is still a considerable volume of hydrogen contained in the tie-in line and an

increased volume in existing line (due to the injection) that would have to be considered in the safety evaluation.

The evaluation of the existing hydrogen pipeline at Waterford 3 is based on a comparative analysis between the hydrogen pipeline and other previously analyzed explosive events. This comparative analysis determined that the hydrogen pipeline has a much lower blast impact (~1/3) and is bounded by the natural gas pipeline traversing the site. Logically, it was concluded that the consequences of a potential break and explosion in the hydrogen pipeline are less severe than the consequences of a break and explosion in the natural gas pipeline. The safety analysis on the natural gas pipeline indicated that a maximum overpressure of 1.6 PSI would result in the worst case explosive event scenario. Even in this worst case scenario, this overpressure is itself substantially less severe than the overpressure or blast impact from a postulated explosion of a Liquefied Petroleum Gas (LPG) tanker truck traversing the highway in front of the plant which could result in an overpressure of 3 PSI.

In the hypothetical scenario presented, there is a considerably greater distance between the electrolyzer facility and the nuclear plant than in either of the bounding events in the existing safety analyses. It should also be noted that the blast impact from the existing hydrogen pipeline is substantially lower than in either of the bounding events. A detailed analysis is beyond the scope of this study, however, it seems reasonable to assume that the stored hydrogen in the electrolyzer facility tie-in line and the added volume that would result in the existing hydrogen pipeline would likely not provide enough blast energy to exceed the 3 PSI bounding condition in the existing safety analysis. Although a more rigorous analysis would be required before embarking on such a project, the cursory evaluation suggests little, if any, regulatory impact.

#### *Regulatory Compliance Impact*

For the hypothetical scenario presented, the regulatory compliance assessment impact is considered minimal. Although the discussion is conceptual in nature and relatively straightforward, the analysis process is consistent with the precedent already established at the Waterford 3 station. Additionally, the discussion does identify the likely areas of concern (blast impact on safety related structures) that would have to be more extensively analyzed were this project actually contemplated. To perform such an evaluation, however, electrolyzer facility design details would be required and exact siting location would have to be established. Once these prerequisites were available, the evaluation would likely require a 2 – 3 man month effort. Such an effort would include the detailed analyses as well as the changes to nuclear plant documentation, as appropriate. Should the evaluation determine that physical changes to the nuclear plant are required or that an NRC review is required, the effort and costs will increase accordingly.

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- <sup>10</sup> Based on pressure vessel quote from John Okonski, Jr, Troy Boiler Works, 2800 7th Avenue, Troy, NY 12180.
- <sup>11</sup> Discussion with Neil Johnson, Senior Engineer, GE Global Research, February, 2007.
- <sup>12</sup> Discussion with Todd Ramsden, NREL, December 10, 2008.