A General Techno-Economic Model for Evaluating Emerging Electrolytic Processes

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A general techno-economic model for evaluating emerging electrolytic processes

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Abstract

Increasing societal concern about carbon emissions and the concomitant emergence of inexpensive renewable resources provide growing impetus for the electrification of the chemical industry. While there have been notable recent advances in the science and engineering of electrolytic processes, there are comparatively few engineering economic studies that outline the technical specifications needed to approach feasibility. Here we introduce an open-source techno-economic framework to connect system performance and price goals to the constituent materials property sets with a goal of quantifying the economic potential of existing and conceptual electrolytic processes. To validate the outputs and demonstrate the versatility of this toolkit, we explore three contemporary electrolyses of varying technology readiness levels. Specifically, we first benchmark our model results against the Department of Energy hydrogen analysis model, then evaluate the impact of mass transport and catalyst performance on the electrochemical reduction of carbon dioxide, and chart a pathway to low-cost electrolytic production of phenol from guaiacol. As this model is based on generalized mass balances and electrochemical equations common to a number of electrochemical processes, it serves as an adaptable toolkit for researchers to evaluate new chemistries and reactor configurations as well as to back-translate system targets to interdependent materials-level property requirements.
1. Introduction

Electrochemical technologies are poised to play a pivotal role in enabling the use of low-cost intermittent renewable energy to transform the global energy economy. To date, most research efforts have focused on advancing electrochemical systems to serve as stationary energy storage, buffering power delivery from variable energy sources on the electric grid, and/or power sources, or displacing internal combustion engines in transportation.

However, efforts to develop electrochemical supplements to traditional chemical manufacturing have been less pronounced, despite the fact that the basic chemical and refining sectors were responsible for 26% of the global energy demand in 2012 and 32% of global carbon emissions in 2010. The ability of traditional manufacturing to take advantage of renewable electrons to replace emissions-intensive processes could begin reductions necessary in this sector to avoid the worst effects of climate change.

Historically, innovations in electrochemical processing have lagged thermochemical processing in large part due to the high quality of electrical energy as compared to thermal energy, leading to economic impetus to use cheap energy sources over more expensive or unreliable ones. However, as low-cost electricity shifts this paradigm opportunities emerge to explore the potential role of electrocatalytic processes, especially under conditions that challenge traditional processing strategies. Several possible advantages of electrochemical reactors include operation near ambient conditions, rapid dynamic response to start-up and shut-down, and innately tunable catalyst/electrode surface energy through varying potential all of which, if realized together, may enable reaction pathways that were not previously attainable.

Accordingly, the effective application of electrocatalysis may both improve the sustainability of extant manufacturing facilities and to unlock exciting new chemistries that could foster the materials and fuels of the future. Indeed, among others, there are growing bodies of work around electrolytic hydrogen production, the electrochemical reduction of carbon dioxide (CO₂) to a variety of useful products, and the electrocatalytic upgrading of biomass-derived
organics.\textsuperscript{[39–42]} While many of these scientific studies and resulting proof-of-principle devices hold promise, significant uncertainty remains around the ultimate performance, durability, and cost of these conceptual technologies.

As economic considerations drive technology adoption, the ability to connect system price targets to component performance parameters at an early stage is key to assessing concept feasibility, identifying technical obstacles, and, ultimately, allocating limited resources most effectively. Indeed, techno-economic models of varying levels of detail and sophistication have been put forth to assess the merits of specific electrochemical technologies.\textsuperscript{[26,43–52]} Most models focus on projecting prior performance data or reasonable predictions of improvement to a desired system price, rather than attempting to translate targeted system prices to required materials performance metrics, for a particular technology. In addition, to the best of our knowledge, no current model has been shown to be applicable to multiple electrocatalytic technologies, despite shared fundamental thermodynamic, kinetic, ohmic, and mass transfer processes described by the same set of constitutive relations. Accordingly, it is desirable to advance techno-economic models that are flexible enough to provide insight into the interdependence of cost-constraining variables thus illuminating pathways to cost reductions not apparent from the manipulations of individual variables. Such models should not only be widely applicable but should be generally accessible to the broader research community interested in electrochemical technology.

Herein, we present a general open-source MATLAB model for evaluating techno-economic feasibility of generic electrolytic reaction schemes. This framework can describe multiple chemistries, is flexible enough to inform correlations between performance-defining variables, and is freely available to the research community for critical assessment, iterative improvement, and application specific refinement. We intend this analytical framework to be useful in establishing design maps that identify performance benchmarks, highlight technical hurdles, and inform fundamental research into the chemistries, materials, and reactors needed for a
viable technology platform. This framework is designed for the purpose of providing the preliminary analysis that is necessary before performing a more detailed technical evaluation of any specific process, and as such we rely on several simplifying assumptions that will need to be carefully considered for the process chemistry of interest. To first demonstrate the accuracy of our model, we show agreement between our results and those reported by the hydrogen analysis model (H2A) of the United States Department of Energy (U.S. DOE) for the production of hydrogen via water electrolysis despite differences in model specificity and formulation. We next establish the model generality by evaluating CO₂ reduction, where the production of carbon monoxide (CO) presents a well-studied model system within which we test the impacts of mass transport and catalyst performance. Finally, we show the utility of the model in exploratory systems such as electrocatalytic hydrogenation by studying the tradeoff between electrocatalyst selectivity and reactor operating conditions in the hypothetical production of phenol from guaiacol. Ultimately, we aim for this toolkit to be broadly accessible to the research community so that they may use it to assess their own research interests.

2. Model Description

In this section, we describe the governing equations and a summary of the simplifying assumptions that are used to represent the electrolysis system. We accomplish the goals described in the Introduction by drawing inspiration from a number of previous techno-economic studies[43,53–56] on different electrochemical systems that sought to describe the connection between component materials and system characteristics. Incorporating these approaches, we are able to accurately simulate a wide range of chemistries and their economic feasibilities without requiring in depth knowledge about the specifics of each process. Although this model can map the feasible design spaces and establish interdependencies between key variables, it does not consider the technical intricacies associated with traversals of this space. We envision that once the design space and key tradeoffs are outlined, process-specific models can be used to refine the most-likely property sets for the associated chemistries. Importantly,
there are certain relationships that are beyond the scope of the current model, such as the particular dependence of the boundary layer thickness within porous electrodes on the velocity of the reaction mixture within the electrolysis stack. For the interested reader, this modeling toolkit with corresponding cost and performance parameters can be freely accessed on GitHub\cite{57} and a more detailed derivation is presented in the supporting information.

The process being modeled (Figure 1) can be broken down into several smaller subprocesses. The first block combines fresh and recycled reactant (R), solvent, and electrolyte. After pretreatment, the reaction mixture is introduced into the electrolytic stack, where R is converted to a desired product (P) or to undesired side product (W) in a parallel reaction. In addition to the parallel reaction network, any protic electrolyte solution in the presence of a cathode at a suitable reducing potential can undergo the parasitic hydrogen evolution reaction (HER), removing catalytically active hydrogen from the surface. The arrangement of the cells electrically in series but hydraulically in parallel within the stack was chosen as this configuration is most commonly used in industrial electrochemical applications.\cite{58}

\begin{figure}[h]
\centering
\includegraphics[width=\textwidth]{process_diagram.pdf}
\caption{An overview of the process modeled. Fresh and recycled reactant is introduced to a pretreatment block from the far left, before continuing to an electrolysis stack, where each cell is connected electrically in series and hydraulically in parallel. Within the stack, the reactants are electrochemically reduced to either desired or waste products. In addition, hydrogen gas can be generated from the reduction of the protic electrolyte solution on the cathode. The oxygen evolution reaction is assumed to occur on the anode though other electrochemical oxidation reactions could be considered. Finally, the crude reaction products enter a separations train, where the desired products are separated to be sold, unreacted material is recycled, and undesired products are wasted.}
\end{figure}
After the reaction mixture leaves the electrolysis stack, it enters the separations train, where unreacted material is separated from the products. For this analysis, we focus primarily on the electrolyzer stack operating at standard temperature and pressure, making simplifying assumptions about the overall process as well as contributions from the separations. First, the stack is assumed to operate at steady state and maintain constant production rate throughout its operating lifetime such that no dynamics, replacement, or component performance decay need be considered. Second, the stack is operated in a galvanostatic manner such that requisite reaction areas and electrochemical kinetic requirements can be directly calculated. Third, we assume the only spatial gradients that exist within each cell arise from mass transfer boundary layers such that heat transfer need not be considered and mass transfer can be simplified to steady-state unidirectional diffusion. Fourth, heats of reaction are ignored such that each cell is operated in an isothermal manner. Fifth, cell separators and membranes are assumed to operate with perfect selectivity, such that there is no product or reactant crossover between the cathodic and anodic chambers. Sixth, we assume that all separations costs can be described by Sherwood mass transfer correlations,\textsuperscript{[59,60]} recognizing that although such correlations have been used previously,\textsuperscript{[43]} they only provide order of magnitude estimations and likely underestimate the complex separations that accompany electrochemical processes. Seventh, we assume that the stack operates with perfect system efficiency, such that no losses arise due to having multiple cells in series due to shunt currents or other inefficiencies. Finally, cost factors for the capital, balance of plant, and additional costs can be estimated from similar electrochemical technologies (see supporting information for additional details on estimated economies of scale).\textsuperscript{[55,56]} While we recognize the limitations, making these assumptions allows us to construct a conservative yet generalizable model using the limited economic data available in the open literature. Using the above assumptions, we develop a framework of mass balances, thermodynamics, kinetics, and mass transfer relations that can be summarized by the annualized...
cost to the consumer shown in Equation 1, which is the basis for all determinations of economic feasibility.

\[ C = \frac{C_{\text{elec}} + C_{\pm} + C_{\text{BOP}} + C_{\text{sep}}}{\tau (1 - F_{\text{add}})} \frac{N_p M_p}{N_{\text{p}} M_{\text{p}}} \]  

(1)

This minimum cost to the consumer \( (C) \) is a combination of operating expenses – electricity \( (C_{\text{elec}}) \) and materials \( (C_{\pm}) \) – and capital costs associated with the electrolyzer stack \( (C_{\text{cap}}) \) and pumps, heat exchangers, and piping associated with the electrolysis stack or pretreatment in the balance of plant \( (C_{\text{BOP}}) \) each of which is normalized by the operating lifetime of the plant \( (\tau) \).

The annualized separations costs \( (C_{\text{sep}}) \) are estimated from Sherwood correlations. The total costs contain additional economic factors such as labor costs, overhead, depreciation of any capital investments, and profit margins, which we assume to be a percentage of the total cost \( (F_{\text{add}}) \). This total cost is then normalized by the production rate \( (\dot{N}_p; \text{mole per second}) \) and molecular weight \( (M_p; \text{gram per mole}) \) to obtain the cost to the consumer in dollars per mass.

The economic feasibility or lack thereof can simply be made by comparison of the minimum cost to the consumer to current selling prices of the desired product. To calculate each individual contribution to this total, the output set assignment is generated first (Table S2). Of the 49 unknowns considered within the model, there are 15 equations relating their values by mass balances, constitutive relations, and component costs, resulting in 34 degrees of freedom through which the model may be parameterized. Equation 2 – 5 show the capital costs, balance of plant costs, electricity costs, and material costs, respectively.

\[ C_{\text{cap}} = A_c (F_{\text{cap}} + \rho_{\text{cat}} F_{\text{cat}}) \]  

(2)

\[ C_{\text{BOP}} = A_c F_{\text{BOP}} \]  

(3)

\[ C_{\text{elec}} = I V P_{\text{elec}} \]  

(4)

\[ C_{\pm} = \dot{N}_{R,f} M_R P_R + \dot{N}_{R,0} \frac{1}{m_R} P_{\text{solute}} (1 - \phi_S) + \frac{-I}{F} M_{\text{salt}} P_{\text{salt}} (1 - \phi_E) \]  

(5)
The capital cost equation is informed by the H2A model from the U.S. DOE\cite{55,56} and redox flow battery modeling by Dmello et al.\cite{53,54} where there is a fixed cost per electrolyzer area ($F_{\text{cap}}$), and the cost factor for the electrocatalyst is considered separately to be a function of the catalyst specific price ($F_{\text{cat}}$; per mass) and specific loading ($\rho_{\text{cat}}$; mass per area). To obtain a total installed capital cost, these cost factors were weighted by the requisite electrolyzer area, which can be obtained from thermodynamics, kinetics, ohmics, mass transfer and their constitutive relations (\textit{vide infra}). The balance of plant is assumed to have the same scaling; namely, as the electrolyzer stack area increases, so too must the cost of the associated hardware as determined by its own areal factor ($F_{\text{BOP}}$).\cite{55,56} Electricity cost contributions can be calculated using the present industrial electricity price ($P_{\text{elec}}$) and the necessary electrolyzer power draw, which can be determined from the product of the total current ($I$) and operating voltage ($V$). Finally, the material costs can be calculated as a sum of the necessary fresh reactant feed ($\dot{N}_{R,0}$), weighted by its molar mass ($M_R$) and price ($P_R$); the potentially required solvent as calculated by the reactant flow rate in the reactor inlet ($\dot{N}_{R,0}$) and its molality ($m_R$), the solvent price ($P_{\text{solv}}$) and fraction of solvent that is recoverable by recycling ($\phi_S$); and the amount of supporting electrolyte that may be necessary to add as a ratio to the amount of charge passed ($r$), the molar mass of the supporting electrolyte ($M_{\text{salt}}$), the price of the electrolyte ($P_{\text{salt}}$), and the fraction of recoverable electrolyte via separation ($\phi_E$). As the need for solvent or supporting electrolyte is reaction and device dependent, they can be incorporated or excluded as necessary. This is explicitly documented in the case studies presented here. All feed cost contributions, consisting of the reactant, solvent, and supporting electrolyte, can be directly calculated from material balances around either the electrolyzer itself or the entire process. To obtain these material balances, we first consider the amount of reactant that must enter the reactor ($\dot{N}_{R,0}$), which can be calculated according to the conversion ($\chi$), the total required current ($I$), the Faraday constant (F), the Faradaic efficiency ($\varepsilon_i$), and number of electrons ($\eta_i$) involved in each reaction, as shown in \textbf{Equation 6}.
\[
\dot{N}_{R,0} = -\frac{I}{\chi_F} \left( \frac{\varepsilon_P}{n_P} + \frac{\varepsilon_W}{n_W} \right)
\]  

(6)

In the use of Faraday’s law presented in Equation (6), we assume that the only three reactions that occur are the production of waste, desired product, and hydrogen, leading to a required unit sum of these Faradaic efficiencies. Although the reactant feed value is indirectly used to calculate the solvent requirements, as shown in the second term in Equation 5, the amount of fresh reactant \( \dot{N}_{R,f} \) can be reduced by recycling a known unreacted fraction \( \phi_R \). The mass balance around the entire facility can be used to compute this fresh reactant feed rate, as shown by Equation 7.

\[
\dot{N}_{R,f} = \dot{N}_{R,0} \left( 1 - \phi_R \left( 1 - \chi \right) \right)
\]  

(7)

The reactor area, which is the critical factor for determining capital costs, can be computed as the ratio of the total current requirement \( I \) to the applied current density \( j \). The total current that is passed can be computed from the fixed production rate \( \dot{N}_P \) and Faradaic efficiency \( \varepsilon_P \) by Faraday’s law, as shown in Equation 8.

\[
A_c = \frac{I}{j} = -\frac{\dot{N}_P n_P F}{j \varepsilon_P}
\]  

(8)

The final cost contribution to be defined is the electrical power draw \( IV \) as shown in Equation 4. As the total applied current has already been computed, the only remaining piece is the cell voltage \( V \). Based on the assumptions of stack efficiency, the total current and cell voltage can be used as a direct proxy for cell current and stack voltage, as the number of cells cancel from the power calculation. The total voltage drop across the cell can be described by the summation of the open circuit (thermodynamic) voltage \( V_0 \), and the overpotentials due to kinetics \( \eta_{\text{kin}} \), ohmics \( \eta_{\text{ohm}} \), and mass transfer \( \eta_{\text{mt}} \) \(-V = V_0 + \eta_{\text{kin}} + \eta_{\text{ohm}} + \eta_{\text{mt}}\). The thermodynamic voltage can be computed from the difference in standard reduction potentials of the two half reactions \( V^0 \), and reactant concentrations according to the Nernst equation, shown in Equation 9. As shown in the supporting information, we assume unit activity coefficient for all reacting species.
\[ V_0 = V^\theta + \frac{RT}{nPF} \ln \left( \frac{1-x}{x} \right) \]  
(9)

The ohmic overpotential is perhaps the most readily calculated as the only requisite information is the ionic conductivity of the medium (\(\sigma\)) and the distance separating the two electrodes (\(L_c\)), from which Ohm’s law can be used, \(\eta_{ohm} = j \frac{L_c}{\sigma}\). The kinetic overpotentials can be calculated from any appropriate electrokinetic theory, but for this implementation of the model we elect to use Butler-Volmer kinetics, which is parameterized by a charge transfer coefficient (\(\alpha\)) and an exchange current density (\(j_0\)) as shown in Equation 10.

\[ j = j_0 \left( \exp \left( -\frac{\alpha n_{\text{kin}} nPF}{RT} \right) - \exp \left( \frac{(1-\alpha) n_{\text{kin}} nPF}{RT} \right) \right) \]  
(10)

Finally, considering the complex topographies of most real electrodes, we begin the analysis of mass transfer by considering the steady state diffusion of species described by a molecular diffusion coefficient (\(D\)) to a planar electrode across a boundary layer of certain thickness (\(\delta\)). Accordingly, the limiting current density (\(j_{lim}\)) and corresponding mass transfer overpotentials can be computed using Equation 11.

\[ j_{lim} = -\frac{nPFDCR_f(1-x)}{\delta}, \eta_{mt} = \frac{RT}{nPF} \ln \left( 1 - \frac{j}{j_{lim}} \right) \]  
(11)

As stated earlier, the primary focus of this model is the physical description of the reaction and transport processes within the electrolyzer itself. However, the relative cost and performance of the separations can significantly impact reactor performance or operating conditions, thus it is necessary to consider separations in any reactor model, even if only from an order of magnitude perspective. In the Sherwood plot shown in Dahmus and Gutowski,[59] there are three classes of materials presented that are all found to fit different lines of the form \(P_{sep} = k_p/w_i\), where \(P_{sep}\) is the separation price, \(w_i\) is the initial concentration, and \(k_p\) is the separation factor in dollars per kilogram of total mixture passing through the separations train. By fitting these data, the authors were able to obtain separation factors of $1 \text{ kg}^{-1}$ of initial mixture for separating biologics, $0.01 \text{ kg}^{-1}$ of initial mixture for separating metals, and $0.001 \text{ kg}^{-1}$ of initial mixture.
for separating gases. The separations costs can then be roughly approximated as the product of these factors and the total mass flow rate through the separations train.

To this point, there have been no assumptions of specific materials properties that are used within the electrolytic stack or as feed or product. Instead, we rely on the specification of independent model parameters to identify these cases. As with all economic models, parameter uncertainty can have significant implications on one’s ability to draw conclusions, which can be especially daunting in the case of evaluating new technologies. However, by examining the trends that may be established through these types of models with standard sensitivity analysis (vide infra) or by using a more complex Monte Carlo approach (Figure S2-S3), researchers can establish bounds upon which they may be able to draw their own conclusions.

For researchers to evaluate chemistries of their own interest, the model can be run by modifying any subset of these model parameters and constructing an instance of the “EconomicCase” object available on GitHub. The model is run through the construction process, which can be verified by examining the “cost” property of the constructed object. Through this approach, we generated a simple toolkit for understanding the physical bounds of the economically viable operating space. The ability to identify the shape of this space allows a more direct translation of system price targets to requisite materials performance targets, illuminating the most pressing technical challenges.

3. Results and Discussion

As described in the previous section, we established a general framework for evaluating the economic feasibility of reductive electrocatalytic reactions. While a full parametric sweep of the 34-dimensional operating space is impractical, it is nevertheless useful to both validate and demonstrate the current model against both established and hypothetical technologies, with a focus on pertinent sets of interdependent cost-constraining variables. Instead of attempting to visualize the full operating space, we project into 1- or 2-dimensional space and examine the interdependencies between subsets of relevant variables and parameters. To that end, we report
the usage of the model through three case studies: water electrolysis to generate hydrogen (Scheme 1a), CO production through reduction of CO$_2$ (Scheme 1b) and phenol production through the electrocatalytic hydrogenation of guaiacol (Scheme 1c).

**Scheme 1.** Simplified representations of electrolytic production of (a) hydrogen from water, (b) carbon monoxide from carbon dioxide, and (c) phenol from guaiacol via a 2-electron reduction with a possible unwanted 6-electron over-reduction to methoxycyclohexanol.

### 3.1. Comparison of current work to U.S. DOE H2A model

Water electrolysis to generate molecular hydrogen is emerging as a promising approach to improve the sustainability of the chemical industry.$^{[27]}$ Hydrogenations, which underpin many important chemical processes, rely on steam methane reforming, or other such reforming reactions, followed by compression in capital intensive turbines to generate the necessary supply of high-pressure hydrogen.$^{[61]}$ Water electrolyzers have the potential to harness renewable electricity to generate pressurized hydrogen feeds avoiding additional capital intensive processing steps.$^{[27]}$ However, high electricity costs and expensive catalytic materials have thus far limited technology adoption in favor of the significantly cheaper hydrocarbon reforming processes. Advancing new catalysts, engineering improved reactor designs, realizing economies of scale, and decreasing electricity costs will each contribute to increasing the cost-
competitiveness of electrolytic hydrogen generation.\cite{55,56}

To evaluate the accuracy of the present model, we benchmark our results for water electrolysis against the established H2A model\cite{55,56} at two production scales (1500 kg day\(^{-1}\) forecourt to 50000 kg day\(^{-1}\) central), and show that despite the lack of chemical specificity, the current framework can accurately predict cost behavior (Figure 2). First, in both cases, the cost of electricity dominates the price of hydrogen electrolysis. This scaling of the electricity cost is largely due to the low molecular weight of hydrogen, necessitating high specific power for economic viability. This is motivating research efforts towards increasing the power density while decreasing the overall power consumption of water electrolyzers.\cite{62} The additional costs, balance of plant, and capital cost factors were all matched to the hydrogen cost parameters for this study, leading to nearly identical behavior between the two models. In both models, the feed is treated as process water, which makes up a negligible fraction of the total hydrogen cost. Finally, it is worthwhile to consider the effects of electrolysis cost across different scales. From the H2A data, there appear to be no significant economies of scale, which subsequently translates to the current model as the capital cost factors having little sensitivity to the product throughput. However, it remains unclear the extent to which economies of scale reduce production costs as few electrochemical systems are manufactured at the scale of their thermochemical analogues.\cite{63}
Figure 2. Direct comparison between the U.S. DOE H2A results and the proposed model’s results. (a) At two scales, the model shows quantitative agreement, with slight differences arising from accounting for electrolyte costs. (b) Sensitivity of the model to parameters as reported by the current model and by the U.S. DOE.

Although the total costs appear to appropriately verify the model, it is perhaps more important that the model predicts the correct response to variations in parameter values as shown in Figure 2 (b). A more complete treatment of the model sensitivity by using Monte Carlo simulations is shown in Figure S2. While there is general agreement across multiple parameters, a key discrepancy in model sensitivities is the treatment of system lifetime. Upon deeper inspection of the H2A model, aggressive salvage of the electrolytic system is anticipated. However, as electrochemical systems at scale have not yet been deployed from which salvage values can be extracted, we opt to examine the extreme case of obtaining no salvage value from the capital purchase. In this limiting case, we may overestimate the impact of system lifetime, but based on the lack of clarity in the H2A model, this approach provides the most conservative estimate of the process economics.
3.2. The impact of mass transfer and catalyst performance on CO₂ reduction

One of the most studied electrolytic reactions over the last decade is the electrochemical reduction of CO₂, which has potential value as a method for carbon-neutral fuel production or waste stream valorization, depending on the electrocatalyst used and electrolysis conditions. Concerted efforts on catalyst development have led to the generation of two-electron reduction product such as CO and formate at Faradic efficiencies that approach and surpass 95% as well as steady advances in selective generation of more deeply reduced products.[26] However, a number of scientific and engineering challenges remain for the commercialization of CO₂ reduction;[64–67] for example, the successful incorporation of highly engineered catalysts into high surface area electrodes that are stable over the reactor lifetime.[65] Despite these hurdles, multiple start-ups are taking innovative approaches to developing economically feasible and large-scale CO₂ reduction.[34]

Although well-founded CO₂ reduction economic models exist in the literature,[26,43,68] several important questions persist for the general design space for CO₂ electrolysis. First, from the suite of possible products that can be made from CO₂ reduction, which is the most promising? Second, what feed source of the CO₂ should be used for electrochemical reduction? Third, what are the tradeoffs in catalyst material choice in the importance of activity and selectivity? Using our model, we can begin to address these questions at a high-level, outlining the shape of the design space for these devices and developing the intuition necessary to better design future electrolytic processes.

The cost to the consumer of CO, formate, methane (CH₄), ethylene (C₂H₄), and ethanol (C₂H₅OH) based on the model predictions shown in Figure 3 suggest the economic feasibility of CO production from CO₂. In these estimates, gaseous CO, methane, and ethylene were assumed to be generated without supporting electrolyte or solvent in the reactant or product streams, while liquid formate and ethanol were generated in the presence of a carrier solvent, but no electrolyte, that could remove them from the electrolysis reactor. From these cost
breakdowns, the number of electrons necessary to produce CH\textsubscript{4}, C\textsubscript{2}H\textsubscript{4}, and C\textsubscript{2}H\textsubscript{5}OH makes electrical costs a significant contribution to the total cost. Consequently, there are correspondingly high capital and balance of plant costs. Although the electric and capital requirements for both CO and formic acid appear promising, the added difficulty in the liquid separation adds significant cost to formic acid production. To estimate the separations cost, we use separation factors of $0.003$ kg\textsuperscript{-1} of mixture for all gas separations and $0.4$ kg\textsuperscript{-1} of mixture for all liquid separations. While data for non-biologic liquid products was not found in published Sherwood charts, we assume these separations would be about one order of magnitude less expensive. The cost of these separations was verified by more rigorous analysis of the distillation train using ASPEN Plus simulations (see supporting information). In each of these estimated separation factors, the factor of either 3 or 4 arises from the number of products (reactant, product, waste, and potentially solvent) that must be separated from the crude reaction stream. For this analysis, we only consider separations of products in the same physical phase, treating all secondary products in that phase as a single waste stream. We note that for these separations estimates, the separation cost is directly correlated to the separation factor indicating that any change in the difficulty of the separation impacts the separation cost in the same way. We compare these costs to the consumer with the average selling prices as reported by Verma et al. and show quantitatively that CO can be produced in an economically feasible manner with current material performance. However, there is still a significant gap to overcome to synthesize methane (CH\textsubscript{4}), ethylene (C\textsubscript{2}H\textsubscript{4}), and ethanol (C\textsubscript{2}H\textsubscript{5}OH) either due to the natural abundance of the desired product or inexpensive thermochemical alternatives for its production. In addition, due to the significant current requirements, even if electricity prices were to decrease from $0.0612$ kWh\textsuperscript{-1} to $0.03$ kWh\textsuperscript{-1} and separations costs could be drastically decreased, the capital and balance of plant requirements still make these products economically infeasible with present technology. Note that we do not consider carbon taxes in this analysis, but recognize that additional incentives to process CO\textsubscript{2} to more valuable materials could offset
some of these economic penalties. In addition, here we only consider direct electrochemical transformations as opposed to more sophisticated tandem electrochemical/thermochemical systems. Such tandem systems could be easily studied using these types of general economic frameworks, if desired. For the remainder of this section, we focus on CO production as it is potentially profitable, as others have previously noted.[26,43]

**Figure 3.** The cost breakdown of CO\(_2\) reduction technology to a variety of different products including CO, formate, methane (CH\(_4\)), ethylene (C\(_2\)H\(_4\)), and ethanol (C\(_2\)H\(_5\)OH). Each bar contains cost components from electrical power, separations, electrolyte, additions to the total cost, balance of plant, and capital investment. The bars on the left use the parameters described in Table S2, with an electricity price of $0.0612 kWh\(^{-1}\), while the hashed bars use an electricity price of $0.03 kWh\(^{-1}\) based on predictions of future electricity prices.\(^{[25]}\)

The feed concentration required for an economically viable CO\(_2\) reduction process determines the feasibility of different feedstocks as well as the scale of any necessary preconcentration steps. Also important to process operation is the conversion per pass, as this value has a significant impact on the composition of the reactor exit stream, and direct implications for the difficulty of the separations. To a rough approximation, Sherwood plots have been used in prior literature for calculating the cost of separations technologies, each of which scales with the mixture throughput.\(^{[43,59,60]}\) In this case, as per-pass conversion decreases, both the separations costs and reactant concentrations increase, as a low conversion indicates that there is more unreacted material in the bulk fluid which necessitates more material passing through the
separations train. Our model suggests that minimum reactor feed CO\textsubscript{2} concentrations of ca. 18% v/v are necessary, with separation costs becoming insignificant at conversions greater than ca. 40% (Figure 4). This is in agreement with the work published by Pletcher, who postulated that CO\textsubscript{2} electrolysis would likely need significant pre-concentration before it could be reacted in an electrolyzer.\cite{68} At less than ca. 18% and high current densities, the bulk concentration is too low to support access of the CO\textsubscript{2} gas to the electrocatalyst within the electrode causing mass transfer limitations which necessitate larger reactors and increased cost. Enabling efficient operation at low feed concentrations would require advances in electrode design or manipulation of mass transport characteristics to either increase the effective diffusion coefficient or decrease the boundary layer thickness. Similarly, there are diminishing returns on any per-pass conversions above ca. 40%, as, at this point, the costs of separations are negligible compared to more significant contributions such as electrical power and reactor capital. In addition, operating at higher conversions necessitates higher purity feeds, as the bulk concentration will deplete to the point of slowing mass transfer processes. Notably, we do not account for any impacts that chemical impurities in the feedstocks that may have on either reactor or separations performance in this analysis.
Figure 4. The impact of mass transfer and separations on the cost of reducing CO$_2$ can be significant. As the single pass conversion inside the electrolyzer changes, there is a sharp mass transfer limit that cannot be surpassed, which effectively limits the feed concentration of CO$_2$ to the unit. The shaded contours represent different per pass conversions ($\chi$).

One of the most pressing questions for CO$_2$ electrolysis surrounds the tradeoffs between catalyst productivity and selectivity in a single material. To wit, for a particular catalyst, what is the necessary activity and selectivity to achieve particular system cost targets? Here, to explicitly evaluate the trade-offs that exist, we set cost to the consumer and assess required selectivity at any given activity. Figure 5 shows contours of iso-cost with the requisite electrolyzer current draw and efficiency needed to achieve the specified cost targets. As the total cost to the consumer is increased, the range of acceptable materials widens as these can have lower activity and selectivity.

Figure 5. Design space for catalysts in CO$_2$ reduction to CO. As the catalyst selectivity decreases, activity must increase to maintain productivity thus necessitating higher current densities. Mass transfer limitations set a lower bound on the allowable faradaic efficiencies. As the total cost to the consumer decreases, the requirements on catalyst activity and efficiency get more stringent, as indicated by increasing allowable minimum faradaic efficiencies. The shaded contours represent different product costs.

The interplay between the applied current density (i.e., reactor operating condition) and electrocatalyst Faradaic efficiency can be observed along a contour. As the Faradaic efficiency decreases, the total current requirement must increase to keep the production rate fixed, which
necessitates higher applied current densities to prevent increases in capital requirements and electrical power costs. Mass transfer rates of CO$_2$ to the electrode surface set an upper limit of achievable current density, as the reactant surface concentrations approaches zero. This, in turn, establishes a minimum acceptable Faradaic efficiency, which adjusts with the minimum allowable cost to the consumer. It is also important to note the contraction of the design space that occurs from decreasing cost to the consumer, which both increases the minimum faradaic efficiency and minimum current density that must be achieved in any reactor. Promisingly, however, we see that the required current density and selectivity have been achieved in synthetically accessible catalytic materials and advanced reactor systems, albeit for shorter operating times and in smaller formats than ultimately necessary.$^{[31]}$

### 3.3. Investigation of catalyst and reactor tradeoffs in guaiacol electrocatalytic hydrogenation

Although less studied than H$_2$ evolution and CO$_2$ reduction reactions, electrocatalytic hydrogenations (ECHs) are a growing area of electrolyses that hold promise for the sustainable production of more complex chemicals and fuels. ECH differs from electrolytic hydrogen production in that hydrogen adatoms that are electrochemically generated in situ from protic media react with the electro-active unsaturated reactant. Several proof-of-principle examples of such processes include ECH on sugar derived platform chemicals,$^{[69–73]}$ biomass pyrolysis products,$^{[39,41,74,75]}$ and even unsaturated hydrocarbon gases.$^{[76]}$ Owing to the dearth of techno-economic studies beyond large scale life-cycle analyses$^{[77]}$ as compared to other electrochemical technologies, we articulate general performance benchmarks representative of cost-effective ECH using the production of phenol from guaiacol, a popular model of depolymerized lignin found in pyrolysis oil,$^{[40,75]}$ as an example transformation. The molecular structure of guaiacol is nearly identical to that of phenol, only containing an additional methoxy group at the ortho position to the hydroxyl group. As phenol is a highly valued material with significant market sizes, the ability to electrochemically cleave the methoxy group from the
guaiacol, without hydrogenating the aromatic ring, would be highly desirable. Despite demonstrations indicating partial reaction of guaiacol to form phenol,[40,41,75] multiple questions remain regarding optimal reactor geometry and the requisite catalyst performance for cost-competitive processing. Therefore, to identify the operating space for electrolyzers and electrocatalysts, we examine the iso-efficiency curves shown in Figure 6, assuming process water as a solvent and supporting electrolyte to aid ionic conductivity. In these curves, the Faradaic efficiency towards the HER, where the hydrogen adatoms on the surface combine to generate molecular hydrogen gas rather than enter the organic reaction network, is varied within a particular color value, while the different colors identify distinct selectivities between the cleavage of the methoxy group via hydrogenolysis and the saturation of the aromatic ring. For example, within all purple lines the total cost to the consumer at different applied current densities can be observed at a 12% selectivity within the organic network. As the saturation value of these lines decreases (i.e., becomes more gray), the total cost to the consumer decreases as the electrocatalyst becomes better at suppressing HER and therefore more effectively using the provided electricity. By examining different organic network selectivities, which are represented by different color values in Figure 6, we can directly observe the interplay between the three relevant catalyst variables of organic reaction selectivity, HER faradaic efficiency, and electrocatalyst activity.
Figure 6. The sensitivity of phenol production cost to catalyst selectivity and activity. Plots (a) – (d) show different product selectivities indicated by different colored lines, while lines of different intensity within each color scheme show the impact of parasitic hydrogen evolution on the selling price. As lines move towards the bottom right hand of the charts, the product efficiency increases while the hydrogen evolution reaction (HER) efficiency decreases.

These design space maps show qualitatively that it is more important to have a electrocatalyst that can minimize the HER than to have one that is selective within the organic network. There is a minimum reaction selectivity, ca. 24%, in terms of electronic currents within the organic network that must be maintained to drive the production costs below about $1.30 kg\(^{-1}\), which is approximately the global spot price in 2018.\cite{78} Indeed, product selectivities above ca. 50% do not appear to offer significant economic gains, while HER Faradaic efficiency above ca. 20% at 24% product selectivity precludes economic feasibility even in the case of highly active
catalysts. Interestingly, the minimum electrocatalyst activity can be identified from the case of perfect selectivity towards our desired product. Given the aforementioned spot price, this would necessitate a current density of nearly 150 mA cm\(^{-2}\), which may be feasible based on prior work in non-aqueous or mixed media.\(^{79}\) While these representations provide valuable insight into relationships between different materials properties, it is more difficult to outline likely roadmaps to economic viability through sequential advances in this fashion. Instead, we employ a waterfall analysis shown in Figure 7 to highlight the effects of cumulative technology improvements derived from the previous sensitivity analyses. The data for the base case is obtained from experiments performed by Saffron and co-workers in unoptimized H-type electroanalytical cells.\(^{40,75}\) The authors reported phenol and hydrogen Faradaic efficiencies of 6% and 75%, respectively. Based on the difficulty of measuring electrochemically active surface area, only the total currents were reported, but by approximating the geometric area of the electrode, we estimate the applied current density to be near 50 mA cm\(^{-2}\). With these parameters and estimates of other parameters that were not explicitly measured (Table S4), we generate the base case shown on the far left of Figure 7. The first improvement, based on the sensitivity analysis shown in Figure 6, is an increase in the current density from 50 mA cm\(^{-2}\) to 200 mA cm\(^{-2}\), potentially enabled by advances in catalyst material, electrode fabrication, or cell format, resulting in a cost decrease of 18.53 $ kg\(^{-1}\). Next, we assume enhancements in catalyst selectivity such that the Faradaic efficiency of hydrogen generation decreases from 75% to 20%, which further reduces the cost to 6.93 $ kg\(^{-1}\). Finally, we assume the catalyst selectivity within the organic reaction network increases such that Faradaic efficiency of the final product is 75%, a significant increase from the 6% reported,\(^{40,75}\) leading to a final phenol production cost of 0.42 $ kg\(^{-1}\). However, should such an improvement proven unattainable, a target product Faradaic efficiency of ca. 24% would enable production at equivalent to the current phenol spot price, as observed in Figure 6.
Figure 7. Waterfall chart detailing the possible engineering improvements that could be made to the production of phenol electrochemically. Given improvements in catalyst activity and selectivity, and reducing the propensity for hydrogen evolution, phenol could theoretically be produced at prices well below its current spot price.

These analyses help to illustrate pathways to economical production of valuable chemicals via electrocatalytic technologies, but do not indicate that the technology yet exists to perform such electrolyses at scale. To that end, a significant amount of work is still needed before the benefits of such technologies can be realized. Based on the presented analysis, developing catalytic materials, electrodes, and reactors that can support high current density, suppress HER, and remain selective to the desired product remain of paramount importance. However, this particular pathway to economic feasibility is simply a demonstration and others may emerge with technical advances. These new pathways may be mapped in a similar manner. Finally, in this analysis, for the sake of clarity, we exclude the influence of separations, but note that they will impact the overall feasibility of electrocatalytic hydrogenation processes.

4. Conclusion

Here we introduced a general model of thermodynamics, kinetics, ohmics, and mass transfer for evaluating the economic feasibility of electrocatalytic production of a wide variety of materials. First, we established the accuracy of the present model by comparing the U.S. DOE
H2A model results to the predictions that our model make on the electrocatalytic production of hydrogen. Next, we examined the model outputs of the electrochemical reduction of CO₂ to CO, specifically with regard to the tradeoffs in mass transfer and catalyst performance. Finally, we showed a theoretical pathway to low cost electrocatalytic phenol production from guaiacol, which has yet to be experimentally realized, but can help guide future materials research. Through contemplation of these case studies, we hope that the reader can take advantage of the generality of this model and use it to evaluate electrolytic technologies and guide future research in electrolytic reactions of interest. However, as individual projects garner deeper interest from a promising result using this model, additional economic and engineering complexities must be more deeply considered. First, a more detailed treatment of the necessary separations processes is important for strengthening physical connections between the reactor and separations performance. Second, the incorporation of more rigorous economic metrics that can be used to compare against alternative capital projects being considered by a manufacturer. Third and finally, more realistic descriptions of the physical processes within the electrochemical reactor would likely improve the accuracy of performance predictions albeit at the expense of model simplicity and generality. Indeed, chemistry-specific deviations to a typical cell architecture may impact many of the relationships developed here, which must be considered once initial economic feasibility metrics are met, but such specifics are beyond the intended scope of this initial assessment. Through this toolkit of electrochemical engineering economics, researchers will now have a straightforward way of identifying economically feasible chemistries and identifying the most pressing economic challenges at the highest level of analysis.

**Supporting Information**

Supporting Information is available in the attached PDF file.
| Name                           | Units          | Symbol       |
|-------------------------------|----------------|--------------|
| **Cost Variables**            |                |              |
| Average Electricity Price     | $ \text{kWh}^{-1}$ | $P_{\text{elec}}$ |
| Average Solvent Price         | $\text{kg}^{-1}$    | $P_{\text{solv}}$ |
| Average Electrolyte Price     | $\text{kg}^{-1}$    | $P_{\text{salt}}$ |
| Average Feed Price            | $\text{kg}^{-1}$    | $P_{R}$      |
| Average Catalyst Price        | $\text{kg}^{-1}$    | $F_{\text{cat}}$ |
| Areal Reactor Cost            | $\text{m}^{-2}$     | $F_{\text{cap}}$ |
| Areal BOP Cost                | $\text{m}^{-2}$     | $F_{\text{BOP}}$ |
| Additional Cost               | -               | $F_{\text{add}}$ |
| **Electrical Variables**      |                |              |
| Current Density               | $\text{A m}^{-2}$  | $j$          |
| Standard Cell Voltage         | $\text{V}$        | $V_{\text{θ}}$ |
| Product Faradaic Efficiency   | -               | $\varepsilon_{p}$ |
| HER Faradaic Efficiency       | -               | $\varepsilon_{\text{HER}}$ |
| Exchange Current Density      | $\text{A m}^{-2}$  | $j_0$        |
| Transfer Coefficient          | -               | $\alpha$     |
| Number of Electrons in Product| -               | $n_p$        |
| Number of Electrons in Waste  | -               | $n_w$        |
| **Reactor Variables**         |                |              |
| System Lifetime               | yr             | $\tau$      |
| Production Rate               | $\text{mol s}^{-1}$ | $\dot{N}_p$ |
| Temperature                   | K              | $T$         |
| Conversion                    | -              | $\chi$      |
| Cell Gap                      | cm             | $L_e$       |
| Feed Concentration            | $\text{mol m}^{-3}$ | $c_{R,f}$ |
| Feed Molality                 | $\text{mol kg}^{-1}$ | $m_{R}$ |
| Feed Molecular Mass           | $\text{g mol}^{-1}$ | $M_R$ |
| Salt Molecular Mass           | $\text{g mol}^{-1}$ | $M_{\text{salt}}$ |
| Product Molecular Mass        | $\text{g mol}^{-1}$ | $M_p$ |
| Electrolyte Ratio             | -              | $r$         |
| Diffusion Coefficient         | $\text{m}^{2} \text{s}^{-1}$ | $D$   |
| Boundary Layer Thickness      | m              | $\delta$    |
| Electrolyte Conductivity      | $\text{S cm}^{-1}$ | $\sigma$ |
| Electrode Catalyst Loading    | $\text{mg cm}^{-2}$ | $\rho_{\text{cat}}$ |
| **System Variables**          |                |              |
| Electrolyte Recycle Ratio     | -              | $\phi_E$    |
| Feed Recycle Ratio            | -              | $\phi_R$    |
| Solvent Recycle Ratio         | -              | $\phi_S$    |
| **Physical Constants**        |                |              |
| Faraday’s Constant            | $\text{C mol}^{-1}$ | $F$         |
| Universal Gas Constant        | $\text{J mol}^{-1} \text{K}^{-1}$ | $R$   |
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Conflicts of Interest

The authors declare no conflicts of interest.

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Supporting Information

A general techno-economic model for evaluating emerging electrolytic processes

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This supplemental information contains:

11 pages, 3 figures, 4 tables

**Figure S1.** Block flow diagram of electrolysis model
**Figure S2.** Economic Monte Carlo results manipulating one variable
**Figure S3.** Economic Monte Carlo results manipulating all variables
**Table S1.** Scaling of capital and balance of plant cost factors
**Table S2.** Output set assignment of process model
**Table S3-S4.** List of symbols and their base values for the different models
Model Derivation

The model presented in the manuscript relies on several fundamental mass balance relationships, which can all be derived from different control volumes around the black-box processes shown in Figure S1 below.

Figure S1: Block flow diagram with the four most important control volumes considered in the economic model. The red box contains the entire process, including all recycle loops, while the blue box only contains the reactor components, the purple box only contains the separations components, and the orange box contains the point where the recycle and fresh streams mix.

Based on a mass balance on the product exiting the reactor, and the assumption that the product separation is perfect – so there is no product in the recycle loop – we can calculate the total current. There is no product fed to the reactor, the amount of product generated can be calculated based on the current and Faraday’s law, and there is a specified amount of product leaving the reactor ($\dot{N}_p$). Assuming steady state operation (e.g., no change in product concentrations as a function of time), the mass balance simplifies to the following equation, which can be rearranged to arrive at an equation for the current ($I$) in terms of the specified production rate ($\dot{N}_p$), the Faradaic efficiency ($\varepsilon_p$), the Faraday constant ($F$) and the number of electrons consumed by the reaction to product ($n_p$).

$$0 = \dot{N}_p + \frac{\varepsilon_p I}{n_p F}$$  \hspace{1cm} (S1)

Equivalently, the component mass balance considering the waste material can be used to identify the waste flow rate ($\dot{N}_w$), as the total current ($I$), the Faradaic efficiency to waste ($\varepsilon_w$), and the number of electrons consumed by the waste ($n_w$) are all known values.

$$0 = \dot{N}_w + \frac{\varepsilon_w I}{n_w F}$$  \hspace{1cm} (S2)
Next, based on the mass balance of the reactant material around the reactor, and that \( \dot{N}_W + \dot{N}_P + \dot{N}_R = \dot{N}_{R,0} \), we can calculate the amount of reactant that must be fed (\( \dot{N}_{R,0} \)).

\[
\dot{N}_{R,0} = \dot{N}_R + \frac{-1}{F} \left( \frac{e_P}{n_p} + \frac{e_W}{n_W} \right)
\]  

(S3)

However, we do not explicitly specify the exit reactant flow rate (\( \dot{N}_R \)), except through the conversion (\( \chi \)), which is defined by Equation S4 below.

\[
\chi = 1 - \frac{\dot{N}_R}{\dot{N}_{R,0}}
\]  

(S4)

Through this definition of the conversion (which is explicitly specified), and rearrangement of Equation S3, we arrive at Equation 6 presented in the main article.

To determine the amount of fresh feed that is needed (\( \dot{N}_{R,f} \)), we can use the control volume that is shown by the orange rectangle in Figure S1. As we have already calculated \( \dot{N}_{R,0} \), and would like to calculate \( \dot{N}_{R,f} \) (see Figure S1), we must only calculate the amount of reactant that is recycled (\( \dot{N}_{R,r} \)). For the sake of simplicity surrounding the separations process, we assume that a certain fraction of the entering stream can be recycled. That is, by examining the purple block, \( \dot{N}_{R,r} = \phi_R \dot{N}_R \). Returning to the orange control volume, where \( \dot{N}_{R,f} + \dot{N}_{R,r} = \dot{N}_{R,0} \), we can now directly show the following relationship between the fresh feed and necessary reactant feed, conversion, and recycle ratio.

\[
\dot{N}_{R,f} = (1 - \phi_R (1 - \chi)) \dot{N}_{R,0}
\]  

(S5)

This relationship is shown in Equation 7 of the main article. The remaining mass balances that are shown in Equation 4 in the main article are for any solvent requirements and supporting electrolyte solvents respectively. To determine the solvent requirement, we use the definition of molality to compute the solvent flow rate. As we already have the reactant flow rate in mol/s, by dividing by the reactant molality (mol solute/kg solvent), we obtain the solvent flow rate in kg/s. As the separations train has already been discussed, this flow rate can then be weighted by the necessary recycle ratio to obtain the fresh solvent mass flow rate that is necessary, as shown by the central term of Equation 4. Finally, the electrolyte is assumed to operate at a fixed ratio of number of ions needed to number of electrons passed. Based on the fact that the current has already been determined, the number of ions can directly be calculated as shown by the last term in Equation 4.

For both the capital and balance of plant costs shown in Equations 1 and 2 in the main article, there is a factor that weights the reactor area - \( F_{cap} \) and \( F_{BOP} \) respectively. We use the data from the US DOE H2A model to calculate what these factors will be and assume their general applicability per unit electrolyzer stack area. Although likely imperfect, this assumption is anticipated to be relatively accurate for different hypothetical chemistries, as, at present, most continuous electrolysis devices consist of similar architecture to the polymer exchange membrane water electrolyzers.
The cost factors that can be identified from these systems pertain to hardware for flow distributors, current collectors, membranes, gaskets, and heating elements and should be comparable across devices with similar architecture. Chemistry-specific deviations in cell design and engineering would impact this relationship, but such analyses are beyond the intended scope of this initial modeling tool. These factors should change with the reactor throughput, as suggested by the data presented in the H2A model. From the two scales that we examined in this manuscript, at $5 \times 10^4$ kg/day and $1.5 \times 10^3$ kg/day, we extracted the capital and balance of plant cost factors, as shown in Table S1 below. A power density of 27.6 kW/m$^2$ was estimated from the data provided to convert from costs per power requirement to costs per area. This assumption of power density allowed us to convert from hydrogen systems that operate at high power density to systems that may operate at much lower power density but have similar architecture requirements (i.e. other electrolysis systems).

Table S1: Capital and balance of plant costs at the different scales considered by the H2A model.

| Scale  | Capital Costs | BOP Costs |
|--------|---------------|-----------|
| kg/day | $/kW$ | k$/m^2$ | $/kW$ | k$/m^2$ |
| $5 \times 10^4$ | 423 | 11.7 | 477 | 13.2 |
| $1.5 \times 10^3$ | 385 | 10.6 | 555 | 15.3 |

Based on these data, we assumed that a power law description could be used to describe throughputs near those considered, of the form shown in Equation S6.

$$\ln F_i = \ln F_{i,0} + \frac{\ln \left( \frac{F_{i,1}}{F_{i,0}} \right)}{\ln \left( \frac{5 \times 10^4}{1.5 \times 10^3} \right)} \ln \left( \frac{\dot{N}_P}{\dot{N}_{P,0}} \right) \quad (S6)$$

Here, $F_i$ is the capital or balance of plant factor that is desired at the molar throughput $\dot{N}_P$, where the base mass throughput is 1500 kg/day. The terms $F_{i,1}$ and $F_{i,0}$ are the cost factors at 50000 and 1500 kg/day production rates respectively.

The Nernst equation is used to describe the thermodynamic cell voltage ($V_0$) at conditions differing from unit activity. Assuming ideal solutions with unit activity coefficient, the molar concentration of the product and reactant can be used directly in the Nernst equation as shown in Equation S7 below.

$$V_0 = V^\theta + \frac{RT}{n_F} \ln \left( \frac{c_R}{c_P} \right) \quad (S7)$$

Here, the open circuit voltage ($V_0$) is calculated as a difference from the standard reduction voltage ($V^\theta$) for the two half reactions where the reactant species has a bulk concentration of $c_R$ and the product species has a bulk concentration of $c_P$. Accordingly, the bulk reactant and product concentrations can be determined through the definition of the conversion as shown in Equation S4. Due to the fact that both concentrations are referenced to the same volumetric flow rate, we can directly use their molar flow rates as proxies for the species concentration. However, these flow rates
can be simplified back to the reactor conversion and Faradaic efficiency as \( \dot{N}_R = (1 - \chi)\dot{N}_{R,0} \) and \( \dot{N}_P = \chi \varepsilon_p \dot{N}_{R,0} \). Substitution of these relationships into Equation S7 allows us to arrive at Equation 9 in the main text.

The order that these equations must be solved, in addition to the specifications that are necessary, can be identified rigorously by an output set assignment, as shown by Table S2 below.

**Monte Carlo Sensitivity Analysis**

To better understand the sensitivity of the model to the parameters, we performed a Monte Carlo analysis of the total cost to the consumer while varying each of the parameters that did not reflect a property of the model chemistry such as the number of electrons transferred or the standard reduction potential. To perform this analysis, the price of any supporting electrolyte, solvent, electricity, and catalyst were sampled from normal distributions with means of 0.10 $ kg\(^{-1}\), 0.001 $ kg\(^{-1}\), 0.0612 $ kg\(^{-1}\), and 32000 $ kg\(^{-1}\), respectively. The additional factor was sampled from a normal distribution with mean 0.14. The current density, exchange current density, diffusion coefficient, boundary layer thickness, cell gap, conductivity, temperature, catalyst loading, areal reactor cost, and lifetime were all sampled from a normal distribution with means at the baseline values used in the hydrogen production case. For all normal distributions sampled, we used a relative standard deviation of 10%. As the transfer coefficient, conversion, and faradaic efficiency are all bounded between 0 and 1, we used a sigmoid function to transform an unbounded normal distribution to this bounded domain, which was then sampled. Figure S2 shows the results of sampling each of these variables independently and in isolation, such that only one parameter value is modified from the base case at a time. This analysis yields a more complete depiction of the data shown in Figure 2(b) but leads to the same overall conclusions. Alternatively, each parameter could be sampled independently but simultaneously, leading to the cost distribution shown in Figure S3, which appears normal with a sample mean of 5.26 $ kg\(^{-1}\) and a relative standard deviation of 8.5%.

**Suitability of Sherwood Approximations**

To attempt to validate the separations cost estimates provided by the Sherwood analysis, we completed several ASPEN simulations for the distillation of liquid products generated during the CO\(_2\) reduction reaction. These simulations are now included in the supporting information. For the case of formic acid, we simulated a process that consisted of a stoichiometric reactor and a pressure swing distillation with a recycle loop. The first distillation column was operated at 2.07 barg to produce water at 99.9% purity and the azeotrope of formic acid and water at 73% formic acid. The azeotrope was then broken in the second column operated at ambient pressure, where the azeotropic composition exists at 58% formic acid. From this separation, formic acid could be produced at 99% purity, at a total throughput of 0.4 mole s\(^{-1}\) (~66 kg h\(^{-1}\)). Using the ASPEN economic evaluator, the utilities cost for the separation is $580000 y^{-1} with a total installed cost of $1.37 million. Using the effective annual cost (EAC) for this separation with a 7 year lifetime and 0% cost of capital, equivalent to that modeled in our manuscript, is $776000 y^{-1}. For the same process estimated
using the Sherwood plots with a $k_p$ of $0.4 \text{ kg}^{-1} \text{ mol}^{-1}$ of the mixture, the EAC was $576000 \text{ y}^{-1}$. As we clearly state in the manuscript, the Sherwood estimates will not provide all the details that might be necessary to design a full separations train, but they do appear to provide separations costs that arrive at the correct order of magnitude for these separations processes. For the ethanol separation process, another pressure swing setup was used to break the azeotrope, where the ambient pressure column comes first followed by a 2.07 barg column. For these separations, ethanol could be recovered at 99% purity, again at a throughput of 0.4 mol s$^{-1}$, with a predicted installed cost of $1.28$ million and $358000 \text{ y}^{-1}$ of utilities, resulting in a predicted EAC of $542000 \text{ y}^{-1}$ over a 7 year lifetime. When compared to the predictions from the Sherwood plot for this separation, again with a $k_p$ of $0.4 \text{ kg}^{-1} \text{ mol}^{-1}$ of mixture, these agree well with the predicted cost of $577000 \text{ y}^{-1}$. For each of these ASPEN simulations, the NRTL-RK property set was used after fitting to binary vapor liquid equilibrium data that was available from the NIST database. From both scenarios, we observe that although the Sherwood plots are not exactly accurate, they do match the order of magnitude separations EAC as predicted by ASPEN simulations.
Table S2 Output set assignment for the 49 parameters and variables involved in the techno-economic model of electrolytic processes. The numbers correspond to equation numbers within the document, while no number with a diagonal matrix element indicates a specification, which can be found above the gray line.
Figure S2 Simulated cost to the consumer of hydrogen after varying the respective parameter as described in the Monte Carlo economics section.
Figure S3 Simulated cost to the consumer of hydrogen upon 10000 repeated calculations while varying the inputs as described in the Monte Carlo economics section.

Symbols

Table S3 Calculated output variables from techno-economic model and their symbolic representations.

| Name                          | Units    | Symbol |
|-------------------------------|----------|--------|
| **Cost Variables**            |          |        |
| Cost of Production            | $ mol^{-1}| C      |
| Total Capital Cost            | $        | C_{cap}|
| Balance of Plant Cost         | $        | C_{BOP}|
| Separations Costs             | $        | C_{sep}|
| Electrolyte Cost Component    | $ s^{-1} | C_{\pm}|
| Electricity Cost Component    | $ s^{-1} | C_{elec}|
| **Electrical Variables**      |          |        |
| Limiting Current Density      | A m\(^2\) | j_{lim}|
| Mass Transfer Overpotential   | V        | \eta_{mt}|
| Kinetic Overpotential         | V        | \eta_{kin}|
| Ohmic Overpotential           | V        | \eta_{ohm}|
| Open Circuit Voltage          | V        | V_0    |
| Cell Voltage                  | V        | V      |
| Total Current                 | A        | I      |
| **Reactor Variables**         |          |        |
| Reactor Cross Sectional Area  | m\(^2\)  | A_c    |
| Feed Rate of R to Reactor     | mol s\(^{-1}\) | N_{R,0}|
| Feed Rate of R to System      | mol s\(^{-1}\) | N_{R,f}|

Table S4 Input values to the techno-economic model and their symbolic representations. Any empty locations in the table indicate values used from the case of H\(_2\) production. *Value displayed for conversion to CO, for conversion to formic acid, methane, ethylene, and ethanol, values of -1.48 V, -1.06 V, -1.17 V, and -1.15 V were used respectively.
Values for production of CO₂, for production of formic acid, methane, ethylene, and ethanol, values of 0.95, 0.5, 0.5, and 0.3 were used respectively, while the balance for each case produced hydrogen.

| Name                        | Units            | Symbol  | H₂ Values | CO₂ Values | Phenol Values |
|-----------------------------|------------------|---------|-----------|------------|---------------|
| **Cost Variables**          |                  |         |           |            |               |
| Average Electricity Price   | $ kWh⁻¹          | $ₚₑlec | 0.0612⁷  | 0          |               |
| Average Solvent Price       | $ kg⁻¹           | $ₚₛ𝚘ˡᵥ | 0.001²   | 0          |               |
| Average Electrolyte Price   | $ kg⁻¹           | $ₚₛ𝗮ˡ𝐭 | 0.1       | 0          |               |
| Average Feed Price          | $ kg⁻¹           | $ₚ₆    | 0         | 0.0165     | 0.01          |
| Average Catalyst Price      | $ kg⁻¹           | $ₚ₇ₖ  | 32000³⁵   | 552        |               |
| Areal Reactor Cost          | $ m⁻²            | $ₚ₉₆  | 11000⁴⁵   |            |               |
| Areal BOP Cost              | $ m⁻²            | $ₚ₉₇  | 15900⁴⁵   |            |               |
| Additional Cost             |                  | $ₚ₉₈  | 0.14³⁵    |            |               |
| **Electrical Variables**    |                  |         |           |            |               |
| Current Density             | A m⁻²            | j       | -15000⁴⁵   | -1000⁶     | -500⁷⁸       |
| Standard Cell Voltage       | V                | Vₜ      | -1.233⁴⁵   | -1.33³⁹ᵃ   | -1.5         |
| Product Faradaic Efficiency | -                | εₚ      | 0.9⁴⁵     | 0.95⁶ᵇ     | 0.03⁷⁸       |
| HER Faradaic Efficiency     | -                | εₜₜ     | 0         | 0.05⁶ᵇ     | 0.75⁷⁸       |
| Exchange Current Density    | A m⁻²            | j₀      | -0.1      |            |               |
| Transfer Coefficient        | -                | 𝛼       | 0.5       | 0.1        | 0.1          |
| Number of Electrons in Product | -             | nₚ      | 2         |            |               |
| Number of Electrons in Waste | -              | nₜₜ     | 2         |            |               |
| **Reactor Variables**       |                  |         |           |            |               |
| System Lifetime             | y                | τ       | 7         | 20         |               |
| Production Rate             | mol s⁻¹          | ḇₚ₆    | 8.6       | 0.4        | 0.2          |
| Temperature                 | K                | τ       | 298.15    |            |               |
| Conversion                  | -                | 𝛼       | 0.5       | 0.2        |               |
| Cell Gap                    | cm               | Lₑ      | 0.02      |            |               |
| Feed Concentration          | mol m⁻³          | cₕₜ    | 1000      | 44.64      |               |
| Feed Molality               | mol kg⁻¹         | mₚ₆    | 1         | ∞          |               |
| Feed Molecular Mass         | g mol⁻¹          | Mₚ₆    | 1.01      | 44         | 124.14       |
| Salt Molecular Mass         | g mol⁻¹          | Mₛᵃˡᵗ  | 98        |            |               |
| Product Molecular Mass      | g mol⁻¹          | Mₚ₇    | 2.02      | 28         | 94.11        |
| Electrolyte Ratio           | -                | r       | 1         | 0          |               |
| Diffusion Coefficient       | m² s⁻¹           | D       | 5×10⁹      | 2.2×10⁹     |               |
| Boundary Layer Thickness    | m                | δ       | 1×10⁵     | 3×10⁶      |               |
| Electrolyte Conductivity    | S cm⁻¹           | σ       | 0.1       |            |               |
| Electrode Catalyst Loading  | mg cm⁻²          | ρₖᵣₑₗ | 0.01      | 10         |               |
| **System Variables**        |                  |         |           |            |               |
| Electrolyte Recycle Ratio   | -                | φₑ      | 0.99      |            |               |
| Feed Recycle Ratio          | -                | φₖ      | 0.99      |            |               |
| Solvent Recycle Ratio       | -                | φₛ      | 0.99      |            |               |
| **Physical Constants**       |                  |         |           |            |               |
| Faraday’s Constant          | C mol⁻¹          | F       | 96485     |            |               |
| Universal Gas Constant      | J mol⁻¹ K⁻¹      | R       | 8.314     |            |               |
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