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Citation: Stinn, Caspar and Antoine Allanore. "Estimating the Capital Costs of Electrowinning Processes." *The Electrochemical Society Interface* 29, 2 (June 2020): 44. © 2020 The Electrochemical Society

As Published: <http://dx.doi.org/10.1149/2.f06202if>

Publisher: The Electrochemical Society

Persistent URL: <https://hdl.handle.net/1721.1/131073>

Version: Author's final manuscript: final author's manuscript post peer review, without publisher's formatting or copy editing

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Estimating the Capital Costs of Electrowinning Processes

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Abstract

Novel electrolysis processes remain strong technological contenders for advances in sustainable materials processing, in particular metals, yet will need to compete economically with currently-deployed production facilities. To evaluate the technoeconomic efficacy of new electrolytic metal extraction processes, an understanding of the capital and operating costs of electrowinning is necessary. Estimation of electrochemical operating costs has been afforded due attention, yet capital cost (CAPEX) trends are far less understood. Herein, we attempt to show that estimating the capital costs of electrowinning processes via conventional chemical engineering scaling laws is not possible. Instead, we propose a capital cost model for electrochemical processes based on relevant operating parameters such as current density, temperature, and voltage. The new model for capital cost describes within ± 30 to 100% the capex for existing electrochemical processes, sufficient for order of magnitude and preliminary design capital cost estimation.

Keywords: electrowinning, electrolyzer, metal-extraction, capital cost, technoeconomics

Introduction

A drive towards industrial sustainability, coupled with rapidly evolving markets, motivates the recent reassessment of metallurgical extraction and metal recycling technologies. Challenges pertaining to the mitigation of greenhouse gas emissions from pyrometallurgical processes, along with a desire to minimize the use of toxic or expensive reagents from leaching and solvent extraction technologies, have led to significant research on more environmentally-benign electrochemical processing methods¹⁻³. Currently, electrochemical technologies are employed at an industrial scale for the production of metals including aluminum, beryllium, calcium, chromium, copper, lithium, magnesium, manganese, nickel, platinum group metals, rare earth metals, sodium, titanium, and zinc⁴. Other commodities produced via electrochemical methods include caustic sodium hydroxide (soda) and chlorine⁴. In recent years, innovative electrowinning processes have been proposed and reimaged for the primary extraction and production of metals such as copper⁵, iron⁶, titanium⁷, molybdenum⁸, rhenium⁸, rare earth metals⁹, and aluminum master alloys such as those of scandium¹⁰. Meanwhile, electrochemical approaches have been the focus of many new recycling and recovery technologies for strategic metals such as indium¹¹, cobalt¹², and lithium¹³.

While contemporary electrochemical research shows significant technological promise, any new, potentially disruptive processing technology will be in competition with existing and established processing facilities, some of them constructed less than a decade ago. Presently-employed metallurgical processes benefit from previously invested capital and ongoing projects, existing know-how, and have been deployed as state-of-the-art facilities in the last decade, for example in China. Furthermore, due to the inherent risk of new technology, current industrial best-practice embodies a significant degree of technological and economic inertia that new processing routes can only overcome through the promise of substantial technoeconomic improvements. For a new electrochemical process to replace an existing method, the reward must be worth the risk. In the current industrial climate, to ensure adoption the technological progress must result in a step change in relevant technoeconomic metrics such as cost, product purity, efficiency, sustainability, etc.

In order to ascertain the ability of a new electrochemical process to be technoeconomically feasible and competitive, an understanding of the capital and operating costs is necessary. Often, the relative effects of electrochemical metrics on cost are fairly apparent – for instance high current efficiency leads to lower electricity usage per unit of products, and therefore lower electricity costs. However, for comparison of electrochemical methods in very different operating conditions such as temperature or electrolytes, or comparison of electrowinning to solvent extraction or pyrometallurgical methods, a relative cost direction is

not enough to ascertain improvement or motivate a new investment – the economic benefits of new electrochemical processes must be determined and presented quantitatively, an exercise presently undertaken in few academic papers.

In contrast, when current industrial electrolytic processes such as the chlor-alkali process for production of chlorine and caustic soda were optimized in the early to mid-twentieth century, a dedicated research effort was put forth to quantify the effect on operating and capital costs of key process parameters such as cell size and geometry^{14,15}, as well as supporting electrolyzer infrastructure such as bus bars¹⁶ and rectifiers¹⁷. In the latter half of the twentieth century, researchers attempted to establish a framework for the economic analysis of electrochemical methods. For example, Hine studied electrowinning metrics such as the optimal number of cells to minimize cost¹⁸, eventually establishing a basis to study a broad range of technoeconomic aspects for electrochemical processes^{19,20}. Others such as Schmidt and Alkire et al. explored electrowinning optimization by attempting to maximize the profit of a bank of cells^{21,22}. The efforts of these and others³ show that operating costs for electrochemical processes are amenable to quantification and analysis. Meanwhile, the means to quantitatively estimate the capital cost (CAPEX) for an electrochemical facility are far less developed, and will be the emphasis of this article.

This work focuses on estimation of the direct capital cost component of electrochemical processes. For our purposes, we define the direct capital cost to be the installed cost of a facility, ignoring site or geographic specific costs^{23,24}. Our cost analysis does not include operating costs such as electricity, labor, chemical feedstocks, water, or maintenance – methodologies for these costs being well-documented elsewhere¹⁹. We also do not attempt to amortize the capital cost to determine a yearly payback, depreciation, or capital contribution to the overall product cost. Indeed, the manner in which the capital investment of a chemical facility is paid off is largely geography and market dependent, as well as affected by the financial structure and practices of the company or country undertaking the project. Such analysis is beyond the scope of this work. In this paper we present frameworks and scaling relations for quantitative order of magnitude estimates for direct capital cost, and demonstrate that this framework accurately describes the capital costs of existing electrochemical processes. Such order of magnitude estimates cannot replace detailed design analysis however, yet they remain useful for comparison of technologies.

Methodology of Capital Cost Estimation

In determining the economic feasibility of a process, different levels of detail are explored throughout the design period, with varying levels of stated accuracy. The American Association of Chemical Engineering (AIChE) International define five levels of cost detail: Class 5 – Order of Magnitude (within ± 50 -100 %), Class 4 – Preliminary (within ± 30 -50 %), Class 3 – Definitive (within ± 10 -15 %), Class 2 – Detailed (within ± 5 -10 %), and Class 1 – As Bid (within ± 5 -10 %) ²³. Early Class 5 and Class 4 cost estimates focus on determining whether or not a product is economically tenable to produce, and by which method. This would be used to compare various processing routes, such direct electrochemical versus hydrometallurgical versus pyrometallurgical. Later Class 3-1 cost estimates utilize detailed process flow diagrams, piping and instrument diagrams, and interactions with contractors to further define the construction costs of a process. The correlations and scaling relations described and fitted in this work contribute to Class 5 and 4 estimates.

To consider the effects of inflation and evolving manufacturing techniques, a yearly index called the Chemical Engineering Index (CEI) is tabulated by *Chemical Engineering Magazine*²⁵. While direct comparison of capital costs across different years and construction locations is often dubious, the CEI and similar indices are often the only available tools to assess the magnitudes of previous investments for the purposes of estimating new facility costs^{23,24}. For reported scaling laws in this paper, the CEI has been applied to convert all values to 2018 US dollars.

Application of Conventional Chemical Engineering Capital Cost Models to Electrochemical Processes

Class 5 and Class 4 cost estimates utilize scaling relations to estimate the cost of a new facility from that of similar facilities constructed in the past. Such relations typically take the form of an exponential expression, as shown in Eq. 1, where C

is the direct capital cost (CAPEX) (excluding off-site costs) α is a pre-exponential factor, x is a relevant process parameter (often the production capacity or the power utilized), and β is an exponent.

$$C = \alpha x^\beta \quad (1)$$

Exponent values less than one suggest that the process is more economical at larger scales, whereas exponent values of greater than 1 suggest that the process is more economical at smaller scales^{23,24}. For a given process, α and β are fitted using data for installed capacity and total direct capital investment. For most chemical processes, an exponent of 0.6 is observed, which is called the conventional chemical scaling law^{23,24}. Utilization of an overall process scaling relation like Eq. 1 is characteristic of a Class 5 cost estimate. Class 4 cost estimates require scaling laws for each of the major unit operations within a process, with each law often following the form of Eq. 1.

Despite the use of electrolysis to produce numerous products at very large but different scales, limited literature exists for estimating the capital costs associated with these electrowinning processes. To the author's knowledge, a comprehensive study of existing industrial scale electrowinning processes with the goal of establishing predictive capability for capital cost has not been undertaken. Nevertheless, total capital costs can be found for individual electrowinning facilities for aluminum²⁶, copper²⁶⁻³⁵, magnesium^{26,36}, sodium³⁶, zinc^{36-38,38,39} and chlorine^{40,41}, as presented in Table 1.

Eq. 1 can be fitted to the capital cost of electrochemical processes, where installed capacity (P) is taken to be the relevant operating parameter, x . Other logical choices for the relevant operating parameter could include current or power, however these metrics are often proprietary or difficult to determine. Fittings for the capital cost are reported in Table 2 for aluminum molten salt electrolysis, copper and zinc aqueous electrowinning, and the chlor-alkali process, along with upper and lower bounds for fit validity. These fittings, normalized by tonne of installed capacity, are also presented in Figure 1, extrapolated beyond their applicable capacity bounds, to evaluate the broad range of operating scales for electrochemical processes.

Table 1: Reported total capital costs (2018 USD, using CEI).

	Capacity (kta)	Investment (USD)	USD / tonne	Year	Year CEI	2018 CEI	2018 Investment	USD / tonne	Cell Count	Notes
Al	415	\$ 2,937,550,460	\$ 7,070	2008	576	607	\$ 3,095,647,794	\$ 7,451	517	1 line Rectifier
	415	\$ 2,920,930,656	\$ 7,030	2008	576	607	\$ 3,078,133,521	\$ 7,408	517	1 line Rectifier
	588	\$ 3,691,655,708	\$ 6,275	2008	576	607	\$ 3,890,338,568	\$ 6,613	732	1 line Rectifier
	588	\$ 3,753,428,433	\$ 6,380	2008	576	607	\$ 3,955,435,866	\$ 6,723	732	1 line Rectifier
	735	\$ 4,894,017,677	\$ 6,655	2008	576	607	\$ 5,157,410,990	\$ 7,013	915	1 line Rectifier
	735	\$ 4,360,860,229	\$ 5,930	2008	576	607	\$ 4,595,559,304	\$ 6,249	915	2 line Rectifier
	882	\$ 4,994,766,050	\$ 5,660	2008	576	607	\$ 5,263,581,584	\$ 5,965	1098	2 line Rectifier
	882	\$ 5,599,256,288	\$ 6,345	2008	576	607	\$ 5,900,605,150	\$ 6,686	1098	1 line Rectifier
	1177	\$ 6,712,636,117	\$ 5,705	2008	576	607	\$ 7,073,906,463	\$ 6,012	1465	2 line Rectifier
	1177	\$ 6,518,493,267	\$ 5,540	2008	576	607	\$ 6,869,314,953	\$ 5,838	1465	2 line Rectifier
	1691	\$ 8,676,861,694	\$ 5,130	2008	576	607	\$ 9,143,845,570	\$ 5,406	2105	2 line Rectifier
	500	\$ 3,250,000,000	\$ 6,500	2001	394	607	\$ 5,006,979,695	\$ 10,014	622	2 line Rectifier
	353	\$ 1,899,846,000	\$ 5,382	2008	576	608	\$ 2,005,393,000	\$ 5,681	439	1 line Rectifier
Mg	50	\$ 400,000,000	\$ 8,000	2001	394	607	\$ 616,243,655	\$ 12,325	41	IG Cell
	109	\$ 359,700,000	\$ 3,300	1979	239	607	\$ 913,547,699	\$ 8,381	88	IG Cell
	22	\$ 72,600,000	\$ 3,300	1979	239	607	\$ 184,385,774	\$ 8,381	18	IG Cell
	4.5	\$ 14,850,000	\$ 3,300	1979	239	607	\$ 37,715,272	\$ 8,381	4	IG Cell
	22.5	\$ 74,250,000	\$ 3,300	1979	239	607	\$ 188,576,360	\$ 8,381	18	IG Cell
Na	50	\$ 82,500,000	\$ 1,650	1979	239	607	\$ 209,529,289	\$ 4,191	14	Downs Cell
Zn	100	\$ 250,000,000	\$ 2,500	2006	500	607	\$ 303,500,000	\$ 3,035	766	
	100	\$ 60,000,000	\$ 600	1974	164. 4	607	\$ 221,532,847	\$ 2,215	766	
	120	\$ 508,500,000	\$ 4,238	2011	590	607	\$ 523,151,695	\$ 4,360	919	
Cu	1.1	\$ 2,600,000	\$ 2,398	1980	261	607	\$ 6,046,743	\$ 5,577	15	SX-EW
	1.7	\$ 3,700,000	\$ 2,239	1980	261	607	\$ 8,604,981	\$ 5,207	23	SX-EW
	5.0	\$ 12,400,000	\$ 2,481	1980	261	607	\$ 28,838,314	\$ 5,770	70	SX-EW
	4.6	\$ 9,800,000	\$ 2,140	1980	261	607	\$ 22,791,571	\$ 4,977	64	SX-EW
	15.1	\$ 26,000,000	\$ 1,727	1980	261	607	\$ 60,467,433	\$ 4,016	210	SX-EW
	22.7	\$ 32,700,000	\$ 1,441	1980	261	607	\$ 76,049,425	\$ 3,351	316	SX-EW
	21.2	\$ 33,500,000	\$ 1,579	1980	261	607	\$ 77,909,962	\$ 3,672	296	SX-EW
	27.0	\$ 100,000,000	\$ 3,703	1998	390	607	\$ 155,641,026	\$ 5,764	377	SX-EW
	1.3	\$ 3,000,000	\$ 2,253	1980	390	607	\$ 4,669,231	\$ 3,507	19	SX-EW
	14.6	\$ 25,300,000	\$ 1,728	1980	390	607	\$ 39,377,179	\$ 2,689	204	SX-EW
	50	\$ 268,500,000	\$ 5,370	2014	580	607	\$ 280,999,138	\$ 5,620	697	SX-EW
60	\$ 298,000,000	\$ 4,966	2007	530	607	\$ 341,294,340	\$ 5,688	837	SX-EW	
Cl ₂	200	\$ 106,000,000	\$ 530	1980	261	607	\$ 246,521,073	\$ 1,233	126	diaphragm
	200	\$ 111,500,000	\$ 558	1980	261	607	\$ 259,312,261	\$ 1,297	126	membrane
	200	\$ 112,800,000	\$ 564	1980	261	607	\$ 262,335,632	\$ 1,312	126	membrane + inert anodes
	166	\$ 111,000,000	\$ 669	1990	358	607	\$ 188,203,911	\$ 1,134	105	

Table 2: Scaling law (Eq. 1) parameters for CAPEX estimation of aluminum, copper, zinc, and chloralkali electrowinning.

	Al	Zn	Cu	Cl ₂
α	110,100	340,100	5980	374,100
β	0.787	0.597	0.971	0.521
Capacity Lower Bound, metric tonnes	340,000	80,000	1,000	150,000
Capacity Upper Bound, metric tonnes	2,000,000	130,000	70,000	1,050,000

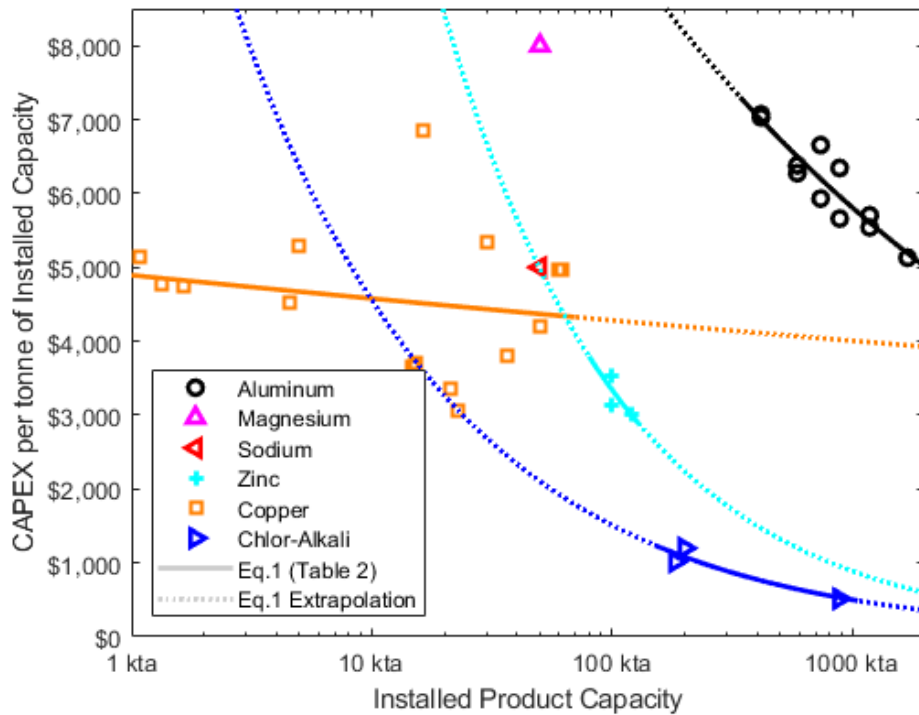


Figure 1: Extrapolation of the Conventional Chemical Engineering Scaling Law to the Electrochemical Processes from Table 1. Eq. 1 describes well the individual process it is fitted to. However fittings of Eq. 1 for one product fails to describe the capital cost for another product, demonstrating that a single model for electrolytic process CAPEX estimates does not exist.

Comparison of the extrapolations for the capital cost show that there is no single model to compare and extrapolate electrowinning economics. The cost curve for aluminum production via molten salt electrolysis drastically over-predicts the capital cost for sodium and magnesium electrolyzer facilities. Proposing aluminum electrolysis as a model system for the capital cost of molten salt processes using a fit to Eq. 1 would suggest that lower tonnage metals could never be produced economically by such a method, a claim that is disputed by the industrial reality of sodium and magnesium production. Meanwhile for aqueous electrowinning processes, the scaling behavior of copper is very different from both zinc and chlor-alkali processing, again highlighting that any one metal aqueous electrowinning process cannot be used to predict the economics of another.

Two shortcomings of the conventional scaling law are therefore apparent: 1) the conventional chemical engineering scaling relation is misleading for estimating the economics of one electrochemical process from another, and 2) a single model electrolytic metal extraction or recovery process does not exist for estimating the capital cost of other electrolytic processes. This finding is not a surprise; indeed, electrochemical reactor productivity scales with the electrode areas (a horizontal surface for aluminum electrolysis in molten salts and a vertical surface for copper electrowinning), as opposed to volume for reactors, furnaces, autoclaves, or other more conventional chemical processing equipment. Furthermore, the operating conditions of the electrochemical process have a significant effect on the upstream and downstream processing that is required in addition to the electrolysis step, as well as the construction of individual electrolysis cells. Relevant conditions such as current density, cell voltage, and temperature have a significant effect on electrochemical process economics that are not captured via conventional chemical engineering scaling laws. We therefore present in the next section an attempt to include those parameters in a model scaling law.

Development of an Electrochemical Engineering Capital Cost Model

We herein develop a model to describe the capital cost of electrolysis processes by considering the relevant operating parameters related to the cathode surface (e.g. current density), as well as available capital cost breakdowns for aluminum, magnesium, copper, and chlor-alkali production. As presented in Table 3, each process has its own operating features that cover a wide range of temperature, electrolytes and electrolysis parameters. This survey shows three main categories of capital investments: front end processing (F), electrolysis and product handling (E), and rectifier (R)^{26,40,41}. Their variation with temperature across chlor-alkali, magnesium, and aluminum production is reported in Figure 2. The total capital investment for an electrochemical process (C) is then described following:

$$C = F + E + R \quad (2)$$

Table 3: Reported Operating Conditions for Electrolysis Processes.

	Al	Mg	Na	Zn	Cu	Cl ₂
Temperature, °C	1000	750	600	50	40	90
Current density, A/m ²	10000	6000	10000	300	300	2700
Current efficiency	0.95	0.9	0.85	0.85	0.8	0.96
Operating Potential (V)	4.18	6.00	5.70	3.50	3.50	3.79
Electrode Area / Cell (m ²)	30	60	60	50	30	55
Current / Cell (A) (kA)	300	360	600	15	9	149
Power / Cell (MW)	1.25	2.16	3.42	0.0525	0.0315	0.563
Electrons per product	3	2	1	2	2	2
Product molar mass (kg)	0.03	0.02	0.02	0.07	0.06	0.07
Yearly productivity / cell (kta)	0.80	1.23	3.67	0.13	0.07	1.59

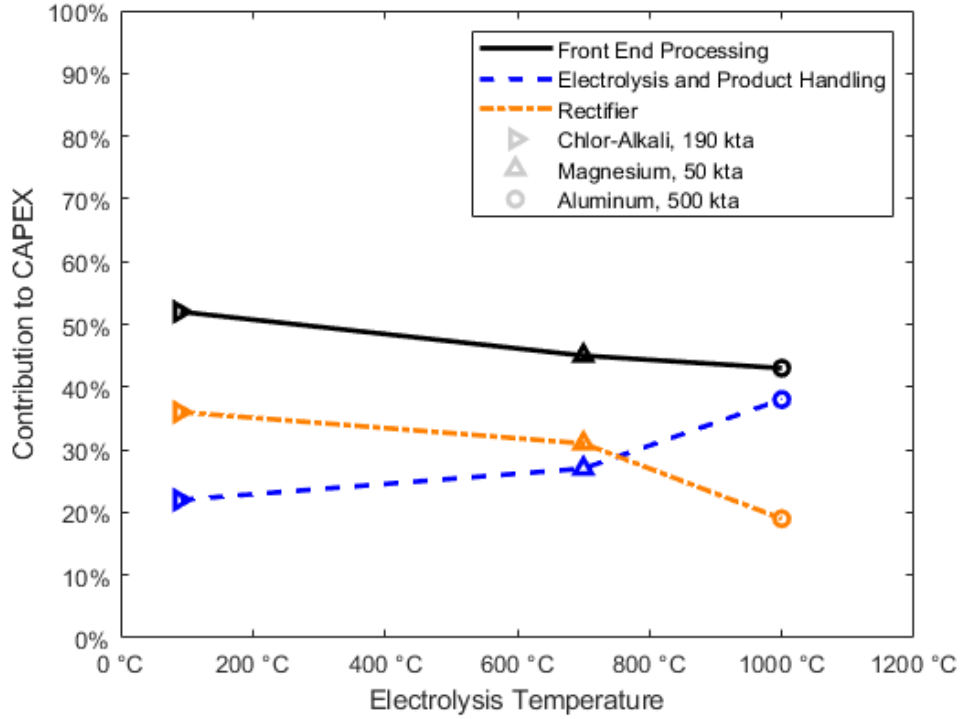


Figure 2: Capital Cost breakdown for current industrial electrolytic production of chlor-alkali, magnesium and aluminum. Front end processing and rectifier cost percentages are shown to decrease with temperature, while electrolyzer and product handling capital cost percentages increase with temperature.

The front-end processing capital costs scale in a similar way as chemical processes and Eq. 1, based on installed capacity (P) and electrolysis temperature (T), where $\alpha_1(T)$ and x_1 are fitted parameters, with x_1 taken to be 0.8 in accordance with correlations for crushers, driers, heaters, and mixers^{23,24}:

$$F = \alpha_1(T)P^{x_1} \quad (3)$$

Capital costs associated with the electrolytic and metal recovery (e.g. stripping of metal or casting) process (E) scales with production capacity via the number of electrolyzers required³¹. The number of electrolyzers required is a function of the total production rate (p), product molar mass (M), current density (j), number of electrons per mole of product (z), current efficiency (ϵ), and electrode (for metals, cathode) area (A)¹⁹. The number of cells is multiplied by a temperature-dependent proportionality constant $\alpha_2(T)$ and raised to some power x_2 expected to fall between x_1 and 1 since reactor production capacity scales by area instead of volume, with less economy of scale due to having many smaller reactors^{23,24,31}. A value of 0.9 for x_2 is proposed, leading to Eq. 4:

$$E = \alpha_2(T) \left(\frac{pz\mathcal{F}}{jA\epsilon M} \right)^{x_2} \quad (4)$$

Previously, Hine described¹⁹ the capital cost of a single rectifier, which is expanded here to account for multiple rectifier lines at a given facility. The capital cost of the rectifier (R) is considered a function of installed power capacity in MW (Q) and cell operating voltage (V), where x_3 is approximately equal to 0.15 and α_3 is a proportionality constant. N is the number of rectifier lines, and x_4 is equal to 0.5:

$$R = \alpha_3(T)QV^{x_3}N^{x_4} \quad (5)$$

The following scaling law is then proposed for the overall capital cost estimates for electrochemical processes:

$$C = \alpha_1(T)P^{x_1} + \alpha_2(T)\left(\frac{pZ\mathcal{F}}{jA\epsilon M}\right)^{x_2} + \alpha_3(T)QV^{x_3}N^{x_4} \quad (6)$$

Using operating conditions for chlor-alkali, magnesium, and aluminum production such as those presented in Table 3, the proportionality constants for Eq. 3-6 α_1 , α_2 , and α_3 can be fitted as a function of electrolysis operating temperature (Figure 3). Electrolysis operating temperature is chosen as the key parameter for the pre-exponential for two reasons; pre-electrolysis processing for aqueous electrolysis predominantly involves fluids whereas high temperature electrolysis requires upstream solids handling, and low temperature electrolysis uses standard engineering materials (polymers, metals, etc.) whereas high temperature electrolysis utilizes refractory materials. Therefore, for front end processing (α_1) and electrolysis / product handling (α_2), distinct low temperature and high temperature regimes exist, with the magnitudes of the proportionality constants increasing with temperature. However, the nature of the transition region from low to high temperature remains unclear. A logistic fit is hypothesized to describe the trends of α_1 and α_2 with increasing temperature, as low temperature processes are more similar to each other than high temperature processes, and vice versa. However, to verify the functional nature of this relation (logistic, stepwise, linear, etc.), technoeconomic analysis of additional intermediate temperature processes will be necessary; however, such information is often proprietary, hindering more in-depth analysis at this time. For rectifiers α_3 is not expected to show temperature dependence, as the nature of rectifiers is dominated by operating power and voltage, not the temperature at which the energy is utilized. Indeed no clear temperature dependence in α_3 is observed for chlor-alkali, magnesium, and aluminum production. Utilizing the fitted values for the proportionality constants α_1 , α_2 , and α_3 , altogether the proposed equation for capital costs reads:

$$C = \frac{51010}{1 + e^{-3.823 \cdot 10^{-3} \cdot (T-631)}} P^{0.8} + \frac{5634000}{1 + e^{-7.813 \cdot 10^{-3} \cdot (T-349)}} \left(\frac{pZ\mathcal{F}}{jA\epsilon M}\right)^{0.9} + 750000QV^{0.15}N^{0.5} \quad (7)$$

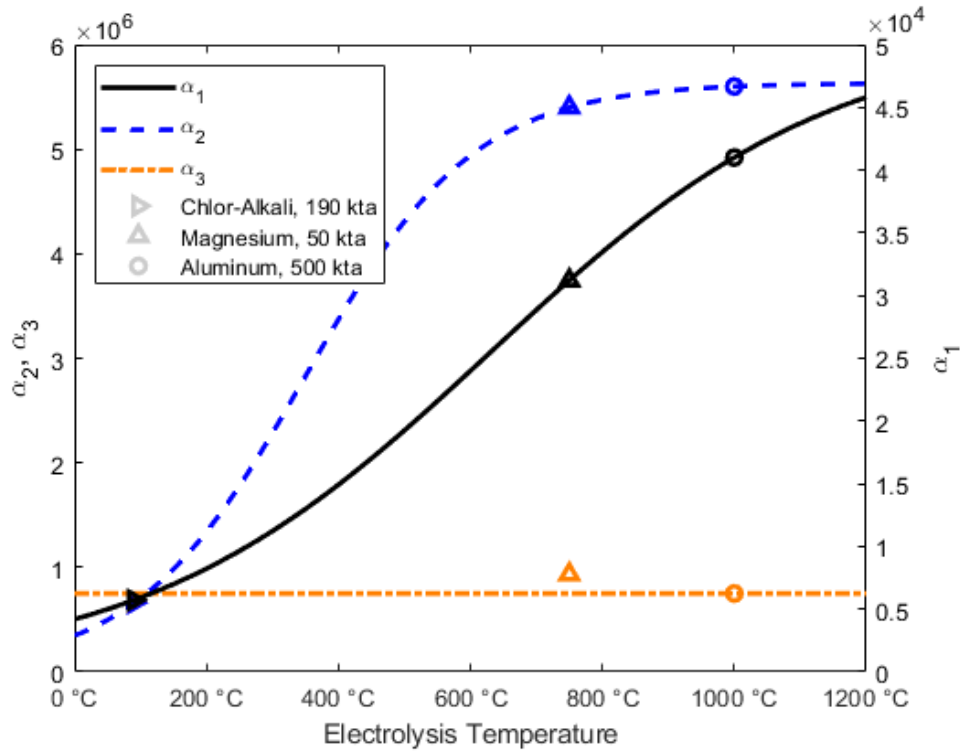


Figure 3: Temperature dependence of the three coefficients for capital cost estimates of electrolysis processes. For front end processing and electrolytic / product handling, the coefficients show a low and high temperature regime. The coefficient for rectifier capital cost is temperature independent.

As shown in Figure 4, Eq. 7 is able to predict the production-capacity-normalized capital cost of electrolytic processes from their relevant electrochemical operating parameters. The results of Eq. 7 fall within the error bars for Class 5 (50-100%) and Class 4 (30-50%) capital cost for all current electrolytic processes for which data is available, suggesting that Eq. 7 is a valuable tool for estimating the capital cost of new electrochemical processes from their operating conditions. This is especially important for estimating the economic tenability of new lab-scale processes where target operating parameters are relatively-well known from basic operating costs estimates. However, those new electrolytic processes are likely to exist outside the framework of existing industrial reality. This means they may have a very different front-end processing compared to existing electrochemical processes, necessitating an understanding of the key unit operation for electrolytic processes: the electrolyzer.

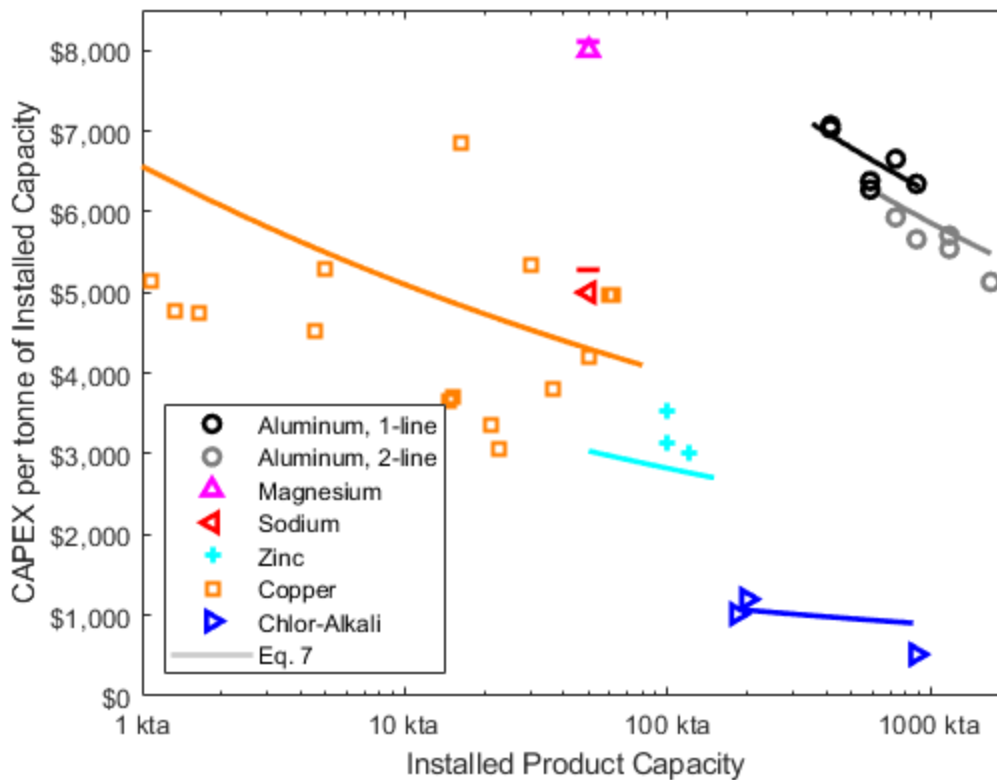


Figure 4: Electrochemical engineering capital cost model applied to current industrial processes. Using relevant electrochemical operating parameters, a single capital cost model, presented in Eq. 7, well-describes the capital cost of electrochemical processes.

Temperature is chosen as the relevant parameter to estimate the cost of electrolyzers for two reasons; both the material cost to build the cell (refractory versus room temperature materials and cell geometry) and the supported current density (moving from aqueous to molten salt electrolytes) are related to the operating temperature. From the data in Table 3, hypothetical installed amperage cost can be derived as a function of temperature (Figure 5), and is predicted to peak between aqueous and molten salt operating temperatures, with aqueous and molten salt electrochemical regimes demonstrating cost competitiveness. This is not a surprise since both aqueous and molten salt process regimes are utilized in industry, with utilization of one regime over the other chosen based on chemistry and thermodynamic considerations. Novel molten sulfide^{5,8,42} and oxide² systems operating above 1000°C have the potential to show even lower net electrolyzer costs for a given productivity than existing molten salt methods. This assumes that the materials cost for the relevant temperature does not increase linearly with temperature. If such conditions (and materials choices) are found, the continual increase in current densities anticipated with temperature suggests high temperature electrolysis as a promising candidate for metallurgical processes from the perspective of capital cost.

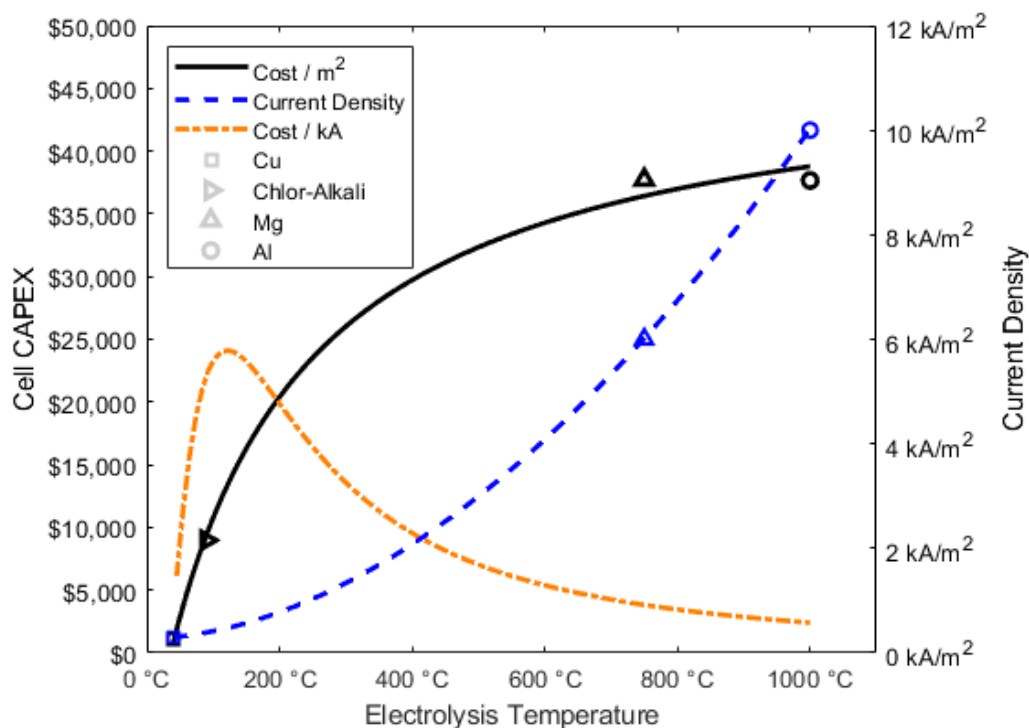


Figure 5: Capital Cost per amperage of electrolyzers. Both aqueous and molten salt cells are shown to be cost competitive when trends in current density are accounted for. Therefore, the choice between aqueous and molten salt methods is first and foremost dependent on system chemistry.

Concluding Remarks

We present a model for estimating the direct capital cost of electrolysis processes. Herein, capital cost refers to the direct cost to build the electrolysis facility, ignoring offsite or location-specific costs. We do not consider operating costs in this work such as the cost of water, electricity, or maintenance, as these are previously described elsewhere. In practice, a balance must be struck between capital and operating costs. For example, a process that has a higher capital cost may be economically superior to one with a lower capital cost once operating costs and amortization are taken into effect. The opposite can also be true – just because a process has a low operating cost does not guarantee it will be economically feasible. A process with a low operating cost can prove untenable if the amortized capital cost is prohibitively expensive. We this tradeoff in mind, we address the capital cost side of this equation.

We fit conventional chemical engineering capital cost scaling laws to aluminum, copper, zinc, and chlor-alkali electrowinning processes and demonstrate that no single, representative electrolytic process exists for predicting electrolytic capital costs of novel processes. Therefore, we develop a new, electrochemical engineering scaling law to predict the capital cost of electrochemical facilities based on relevant operating parameters such as current density, voltage, and electrolysis temperature. We also derive a cost per amperage for estimation of the cost of an individual electrolyzer. This understanding of capital cost allows for the comparison of different electrowinning technologies, as well as the comparison of electrochemical to hydrometallurgical and pyrometallurgical processes.

Symbol List

- A Electrode (for metals, cathode) area, m²
- C Total direct capital cost, 2018 US dollars

E	Electrolysis and product handling contribution to total direct capital cost, 2018 US dollars
F	Front-end processing contribution to total direct capital cost, 2018 US dollars
\mathcal{F}	Magnitude of electric charge per mole of electrons
j	Current density, A/m ²
M	Electrolysis product molar mass, kg/mol
N	Number of rectifier lines
P	Installed yearly production capacity, metric tonnes
p	Total installed production rate, kg/s
Q	Installed power capacity, MW
R	Power rectifying contribution to total direct capital cost, 2018 US dollars
T	Electrolysis temperature, °C
V	Cell operating voltage, V
x	Relevant process parameter for the conventional chemical engineering scaling law
x ₁	Exponent for front-end processing
x ₂	Exponent for number of electrolysis cells
x ₃	Exponent for cell voltage
x ₄	Exponent for number of rectifier lines
z	Moles of electrons reacting to produce a mole of product
α	Proportionality constant for the conventional chemical engineering scaling law
α_1	Temperature-dependent proportionality constant for front-end processing
α_2	Temperature-dependent proportionality constant for electrolysis and product processing
α_3	Proportionality constant for power rectifying
β	Exponent for the conventional chemical engineering scaling law
ε	Current efficiency

About the Authors

Caspar Stinn received his bachelor's degree in chemical engineering from the Massachusetts Institute of Technology (USA) in 2018, where he was awarded the 2018 Haslam Cup for outstanding professional promise in chemical engineering. He is currently a graduate student in the Department of Materials Science and Engineering at the same institution and a member of Prof. Allanore's research group since 2015. Stinn's research focuses on sustainable metallurgical processing, and can be reached at stinn@mit.edu.

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