

## Techno-Commercial Evaluation of Chloride Based Production Routes for Technology Metals and Materials

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### Abstract

Chloride-based hydrometallurgy has historically been constrained to small and medium- scale niche-applications, as it puts high demands on the installed equipment, surrounding buildings and infrastructure in terms of corrosion resistance. Besides that, HCl and Cl<sub>2</sub> are not always available as cheap captive by-products from other chemical processes, and the handling of chlorine containing processing wastes poses various challenges. However, in certain industries, such as Aluminum and Nickel, as a consequence of either depletion of their traditional raw materials base, the demand for extremely low cash cost or for geostrategic reasons, producers are becoming increasingly open to consider novel flow sheets, based on leaching non-conventional primary or secondary feedstocks in hydrochloric acid before applying various separation and extraction steps. In the present paper, we will briefly introduce the building blocks of a Chloride-based flow sheet for a metallurgical process. We will then introduce a hypothetical flow sheet for a processing plant for the production of Smelter Grade Alumina (SGA) from Aluminous Clay (Kaolin), identify the technical challenges and economic success factors involved, present as well as compare several technical options and summarize our findings and recommendations.

**Keywords:** Chloride, Hydrometallurgy, Alumina, Economic Evaluation

### 1. Introduction

Despite metal chloride systems lending themselves to a variety of well-established and efficient metal separation techniques, such as selective volatilization and distillation, selective crystallization, oxidative precipitation, electrolysis, hydrolytic distillation and pyrohydrolysis, their use in industry remains fairly constrained to niche domains, such as the descaling of hot rolled steel in hydrochloric acid baths with closed-circuit regeneration of HCl by pyrohydrolysis, the production of titanium dioxide pigments or sponge titanium from (synthetic) rutile or chloride slag by means of fluidized bed chlorination (with carbonaceous reductants added) and production of synthetic carnallite or anhydrous MgCl<sub>2</sub> by similar processes with subsequent electrolysis of Mg metal.

The volatile and corrosive nature of hydrochloric acid, that frequently leads to a quick detrition of the surrounding buildings, requires the constant attention of experienced operating and maintenance staff, while Cl<sub>2</sub> leaks (or emergency blow-offs) pose a substantial health hazard.

Although reasonably sophisticated technologies for the recovery of Cl<sub>2</sub> and HCl from certain metal chloride solutions have been developed, only trivial cases, involving relatively pure or favorably composed solutions, yield attractive recovery rates beyond 98% in industrial setups.

This, in combination with the extreme requirements for corrosion and abrasion resistance put on the installed processing equipment (valves, pumps) by hot brines and slurries, as well as the sticky, scale-forming nature of certain chloride based liquid/solid mixtures as described in [1], has historically limited the application domain of chloride hydrometallurgy to cases, that are either technically straight-forward (pickling of carbon steel, where iron is the predominant metal and the

effects of the most common alloys/impurities on process performance are benign) or economically very attractive (pigment or Ti metal production, where chloride technology applied to traditional feedstocks yields lower cash cost, a reduced environmental impact and in many cases a superior product than the competing sulfate technology.

The prime motivators to – despite its challenging characteristics from an engineering perspective – still consider flow sheets based on chloride hydrometallurgy may differ from industry to industry.

Some industries (such as Nickel [3] where laterites in the long term will replace sulfates as the prime source) may face circumstances under which conventional feedstocks, which lend themselves to reasonably economic processing, are becoming scarce and have to be replaced by sources, that are more readily processable by chloride, than sulfate or smelter technology. Others may have difficulties getting expansions of their tailings dams (e.g. Bauxite Residue disposal areas at alumina refineries implementing the Bayer process) permitted and thus require processes with a minimized (in at least the dimension constrained by the permit) ecological footprint. Certain smelters may be under regulatory pressure to reduce their emissions or to re-utilize their slag instead of disposing it. Yet another corner of the industrial landscape may be motivated by changing their resource base from imported to domestically available materials to mitigate geostrategic risk or the burden of comparatively high transportation cost as a consequence of the relative remoteness of processing capacity.

Finally, some companies may eye winning their products by means of recycling out of waste (e.g. urban mining) instead of relying on conflicted primary sources potentially linked to slavery and child labor (as e.g. for at least some of world's Cobalt and Tantalum), where again they may benefit from the versatility and potentially high selectivity of a chloride based flowsheet.

## **2. Basic Operations in Chloride Hydrometallurgy**

### **2.1. HCl Leach**

As described in [2] most flow sheets based on chloride hydrometallurgy naturally “start” with dissolving some feed material in hydrochloric acid and/or another reactive chloride species (potentially improving reaction kinetics). This feed material may for example be an ore, a concentrate, a slag or other waste material, but also an ultra-pure source of a metal from which an oxide of particular chemical and physical properties is to be produced.

For many ores (such as in the focus of this paper: Kaolin), concentrates or slags, some thermal pre-treatment, pre-reduction or pre-oxidation may significantly improve their amenability towards leaching at moderate temperatures and pressures.

### **2.1. Oxidation and Oxidative Precipitation**

Oxidation of some metal chloride solutions yields chlorides of a higher oxidation number as well as metal oxides, which can be precipitated or – for combinations of certain species, such as iron and vanadium – co-precipitated.

### **2.3. Selective Crystallization**

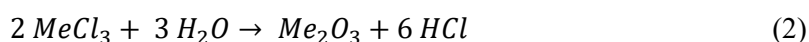
Metal chlorides have sufficiently different water-solubilities and can thus be selectively precipitated by oversaturating the containing solution with (e.g.) HCl.

## 2.4. Volatilization and Distillation

The boiling points of various (anhydrous) metal chlorides are usually low enough to allow separation by volatilization and distillation at technically viable temperatures.

## 2.5. Hydrolytic Distillation

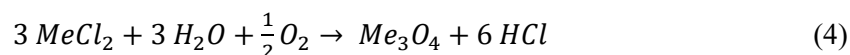
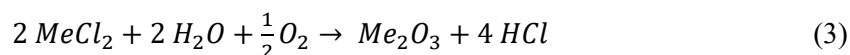
Hydrolysis/Hydrolytic Distillation describes the reaction of a metal chloride with water, typically either performed in a (directly or indirectly) fired furnace at elevated temperature or in a heated reaction vessel containing a highly concentrated metal chloride brine. The following general reaction formulae apply:



Hydrolysis of certain species, such as iron(III)chloride has been shown to be advantageous over pyrohydrolysis in terms of energy consumption. At the same time, various concepts to selectively hydrolyse a major background component of a brine (e.g. iron) in order to enrich it in certain target metals (e.g. Ni, Co) to allow their cost effective extraction have been proposed.

## 2.6. Pyrohydrolysis

As described in [5] pyrohydrolysis describes the reaction of a metal chloride with water and oxygen, typically performed in a directly fired furnace (with burners operated deliberately on a surplus of air or oxygen) at elevated temperatures. The following general reaction formulae apply:



One known limitation of commercially available pyrohydrolysis technology is, that their hydrochloric acid product is limited in concentration to azeotropic HCl to H<sub>2</sub>O ratios – typically around 18% (wt/wt).

## 2.7. Rectification

By means of rectification, hydrochloric acid can be increased in concentration beyond the azeotropic point. Due to nature of hydrochloric acid forming a boiling point maximum azeotrope, HCl gas and acidic water are obtained at the heads of two distillation columns with a circulating sump liquor. The azeotrope is then split either by pressure swing absorption or by addition of an azeotrope breaker as sulfuric acid or CaCl<sub>2</sub>.

## 3. An exemplary Flow Sheet for Smelter Grade Alumina from Kaolin

### 3.1. Motivation

Imagine a – purely fictional - operator of Aluminum smelters, comparatively large by global industry standards, having (e.g. by coincidence) his smelters located reasonably close to some large-scale hydroelectric dams along some mighty northern-hemisphere rivers, not only providing electricity at (almost) unmatched cost and reliability, but also from a 100% renewable

source. Now, imagine the same operator sitting in a relatively inland location, thousands of kilometers away from the next alumina refinery or world class Bauxite deposit, thus incurring close-to-prohibitive transportation cost for his key raw material(s).

While this operator could (in an extension of our fictional scenario) be in a position to turn towards the processing of locally abundant Nepheline ore instead of Bauxite, the associated – presumably large - capex and payback period may be hard to justify in the absence of massive government subsidies as well as strong local markets for the by-products of Nepheline processing.

Coincidentally, however, the geological make-up of our planet has placed at least some cheap-to-mine Kaolin deposits in attractively close proximity to the large streams, that attract the construction of large hydroelectric facilities and aluminum smelters, hungry for low-cost (clean) electricity. Under those circumstances it's not exactly far-fetched to imagine this Aluminum producer investigating possibilities to turn Kaolin into Smelter Grade Alumina.

Those among the esteemed readers of the present paper, who ever had the sobering experience of counting dead animals around a red mud dam, or live close enough to Hungary's unfortunate Ajka plant to get a first-hand experience of the aftermath of its 2010 dam break, sense, that the Bayer process may not be here to stay – particularly not in a world in which more and more people compete for smaller and smaller permissible ecological footprints and our industrial forefather's favorite profit-maximizing-practice of simply passing on the hassle of dealing with long-term environmental liabilities to the next generation is clearly understood to be a non-sustainable idea. As of today, despite numerous bold efforts by highly respectable scientists and companies which have produced significant improvements, red mud remains a lasting ecological liability, with – in the personal view of the author - no reasonably priced cure in sight. It is thus – in the long run and as a consequence of (rarely observed, but not impossible) regulatory long-term thinking – not entirely inconceivable, that Bauxite will be replaced as the primary source of alumina by a material that can be processed into  $Al_2O_3$  at significantly lower ecological expense.

On the other hand, it is apparent that world's smelters (many of which as a consequence of the comparatively large amounts of capital invested are not going to be phased out anytime soon) have been designed to consume Bayer alumina, not only in regard to its chemical and crystalline properties, but also with certain narrowly limited physical properties, such as particle size distribution, floatability, bulk density and free-flowing properties. Producing alumina from sources other than Bauxite by methods other than Bayer may – while chemically and thermodynamically potentially viable – yield alumina which cannot be immediately received by existing smelters.

### 3.2. Process

A rather typical Kaolin could consist of 33%  $Al_2O_3$ , 53%  $SiO_2$ , 0.5%  $Fe_2O_3$  and 0.5% of other metal oxides, and have a loss on ignition of about 13%.

A matching general, chloride based process for producing smelter grade (or other) alumina from Kaolin would be based on the mechanical preparation (screening, sizing), an acid leach (under ambient pressure after thermal activation, or in an autoclave), silica filtration, crystallization of aluminum chloride hexahydrate (ACH) by means of HCl sparging or concentration by evaporation, ACH calcination and final conversion (deliberately not discussed in this paper) of the resulting "crude" alumina into a product that can be readily received and processed by existing smelters.

A sophisticated chloride and water management (for the purpose of this paper summarized as “acid regeneration”) is required to minimize the fresh acid and energy consumption of the system.

Figure 1 gives an exemplary – very high level – illustration of the process flow through such a process, together with some rough quantitative estimates for the mass flows, that allow the interested reader a first assessment of the expected scale of a processing operation which implements the flow sheet.

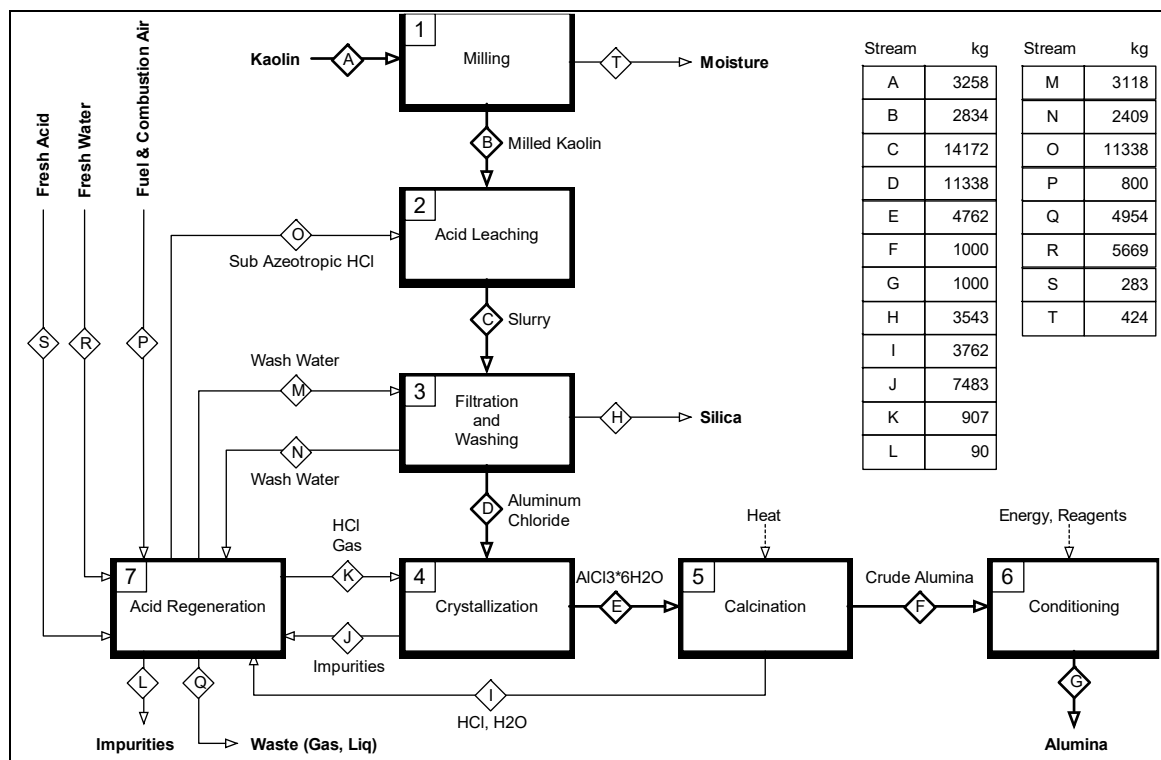


Figure 1. An exemplary flow sheet for SGA production via the chloride route

### 3.2.1. Milling

In order to prepare the Kaolin for leaching at acceptable rates, it needs to be dried and milled to a particle size of typically less than 300µm. Additional mechanical feed preparation steps, such as wet screening of oversize silica and quartz may prove effective in reducing milling cost, the downstream SiO<sub>2</sub> solid load as well as the associated processing cost.

### 3.2.2. Acid Leaching

The milled Kaolin is leached in sub-azeotropic hydrochloric acid at a concentration of approx. 18%. The leach may be performed under atmospheric pressure after thermal activation (higher energy consumption as a consequence of said activation step) or in an autoclave (reduced energy consumption, but higher capital cost and increased complexity of safe and reliable operation). All components of the Kaolin, with the exception of SiO<sub>2</sub> - Fe<sub>2</sub>O<sub>3</sub>, CaO, MgO, Na<sub>2</sub>O, K<sub>2</sub>O) - react with HCl to form their respective chlorides. 1kg of Al<sub>2</sub>O<sub>3</sub> requires 2.14 kg pure HCl to form 2.61 kg AlCl<sub>3</sub> and 0.53 kg H<sub>2</sub>O. The leaching operation can be batch or continuous and should be provided with an appropriate over-stoichiometric amount of acid in order to keep the required leach retention times low.

### 3.2.3. Filtration and Washing

The insoluble SiO<sub>2</sub> is filtered from the slurry. Chlorides adhering to the surface of the filtered silica are washed out and recovered by a – usually countercurrent – washing process. Typically a final filter cake with a moisture of less than 40% and a chloride content of less than 1% would be produced.

### 3.2.4. Crystallization

After the removal of solids, the metal chloride solution (mainly AlCl<sub>3</sub>) is oversaturated by bubbling dry hydrogen chloride or evaporation to increase the HCl concentration in liquor. As a consequence of its low solubility (compared to other species in the solution), aluminum chloride precipitates as hexahydrate crystals, which can be separated via centrifuges or belt filters, and further purified by a concentrated HCl wash. Depending on the required purity of the (crude) Alumina product, the crystals can be re-dissolved in hydrochloric acid and the crystallization step can be repeated for as many times as necessary. The number of required HCl sparging/crystallization steps depends on the purity requirements for the produced AlCl<sub>3</sub> crystals.

### 3.2.5. Calcination

The Aluminum Chloride Hexahydrate is calcined into alumina via hydrolysis with its crystal water in an indirectly heated kiln, to avoid the introduction of additional water in the calcination off gas. The off gas from this kiln (mainly HCl and H<sub>2</sub>O) is fed to the acid regeneration unit.

### 3.2.6. Conditioning

The crude alumina produced in the calcination step is not expected to have physical properties similar to conventional Bayer Alumina. A product with a significantly lower bulk density, not readily usable in a conventional Aluminum smelter is expected. Thus, an additional – typically proprietary – process step, which requires the physical conditioning or dissolution and re-precipitation of the alumina (not necessarily in a Chloride environment, though), is required.

### 3.2.7. Acid Regeneration

At the core of the proposed chloride based flow sheet for SGA production is a sophisticated acid regeneration system, which – by means of rectification, pyrohydrolysis and potentially hydrolysis – closes the acid balance of the plant at the required concentrations, and effectively separates the main impurities from the system as their oxides or chlorides.

## 3.3. Technical Viability / Scalability

### 3.3.1. Robustness of Flow Sheet

The flow sheet itself is robust in the sense, that it can deal with fluctuations in feed composition and throughput over a relatively wide band width. A major challenge for any practical implementation may however be presented by the fact, that the management and closed-loop recirculation of HCl gas, hydrochloric acid and water at different levels of concentration and/or purity implies a tightly coupled setup, where any functional interruption of a particular process section will impact or impair the operation of others. At the same time after the start-up of the process after an interruption it may take considerable time until stable production at thermal and chemical equilibrium is obtained.

From a mass- and heat balance point of view, the flow sheet is highly sensitive towards deviations of the water balance from design conditions. Adequate preconditions in process detail design to be able to deal with operational disturbances of the water balance should be foreseen.

### **3.3.2. Materials of Construction**

The required materials of construction for the expected temperature and pressure ranges are commercially available (polymers such as PP, PVDF, PTFE and PFA, glass-, rubber-, ebonite- or brick lined steel) and have demonstrated their viability under similar operating conditions in other industries. Titanium (in the presence of Fe<sup>3+</sup> as a corrosion inhibitor), preferably alloyed with palladium, as well as Niobium and Tantalum – when welded under quality controlled conditions – may also prove adequate.

### **3.3.3. Utilized Equipment/Scale Up**

The required equipment for any (in a commercial sense) meaningfully large crude alumina capacity (e.g. 1 million tons per annum) is comparatively large and will in many instances (especially valves and pumps, where they cannot be avoided or replaced by gravity flow, syphons or other technical substituted) require the installation of parallel equipment.

### **3.3.4. Operability and Maintainability**

Mixed metal chlorides (especially when laden with solids or suspended particles) may at certain temperatures and pressures (which may not only occur during regular operation, but e.g. during start-up and shut-down of the plant) form a hard-to-remove scale layer inside vessels, pipes, heat exchangers, pumps and valves. This scale layer can render some of the affected equipment dysfunctional, compromise its performance and/or increase the plant maintenance requirements while reducing available net operating time.

Glass lined pipes and vessels show excellent resistance against chloride corrosion and abrasion, but are notoriously difficult to repair in-situ. They are prone to suffer from mechanical cracking at flanges and joints as a consequence of mechanical tension (e.g. induced by heat expansion) and require highly qualified staff for installation and maintenance. Naturally the accidental introduction of sufficiently large, hard solids into glass lined pipes during operation or (sloppy) maintenance has to be avoided/prevented, as those could cause a damage of the glass lining, too.

At the same time at any plant operating with hydrochloric acid, high-quality maintenance and the strict avoidance of the escape of HCl vapor is a prerequisite for operational success. The presence of a well experienced, well trained, well managed and well equipped operation and maintenance crew is thus expected to be of crucial importance.

### **3.3.5. Environmental impact**

A hydrometallurgical process that relies on the recycling of a main reactant is more often than not a large consumer of heat, which requires fuels, that can within the boundaries of economic feasibility only be drawn from fossil fuels. While such fuels (lignite, natural gas) may be available in abundant quantities near potential locations of future chloride based SGA refineries, their combustion will burden the process with a significant CO<sub>2</sub> footprint. Assuming roughly 50 GJ in thermal energy consumption per ton of crude alumina produced, this yields approx. 2.8 tons of CO<sub>2</sub> per ton in case of firing natural gas and approx. 5.5 tons of CO<sub>2</sub> per ton in case of firing lignite. This is more than three times the CO<sub>2</sub> footprint of the Bayer process. However some of the excess CO<sub>2</sub> emitted by the proposed process may be set off by the avoidance of – potentially long distance – Bauxite and alumina transportation cost.

### 3.4. A Framework for assessing Economic Feasibility: BPA

When considering an investment into a new production facility for smelter grade alumina (like for any other capital asset), a rational investor with limited resources would most likely resort to various (to some extent arbitrarily selected and weighted) constraints and metrics, such as expected Net Present Value (NPV), Internal Rate of Return (IRR) or Payback Period (PP).

Over the recent years, the author of the present paper has developed the BPA (Basic Project Analysis) framework for the quick analysis of investment projects. At its core is the BPA diagram, which is a comfortable tool for quickly devising the expected IRR of a project without going through too much mathematical hassle and within a ball park small enough to aid making the decision on whether or not it is worthwhile to pursue a more thorough investment analysis. In the experience of the author (caution: anecdotal evidence!) investors frequently use the double of the S&P global WACC (Weighted Average Cost of Capital) as the pre-tax hurdle rate for an investment. The SPGI WACC presently is at 7.5%. Therefore the hurdle rate (green line in the BPA diagram) for the purpose of this paper was set at 15%.

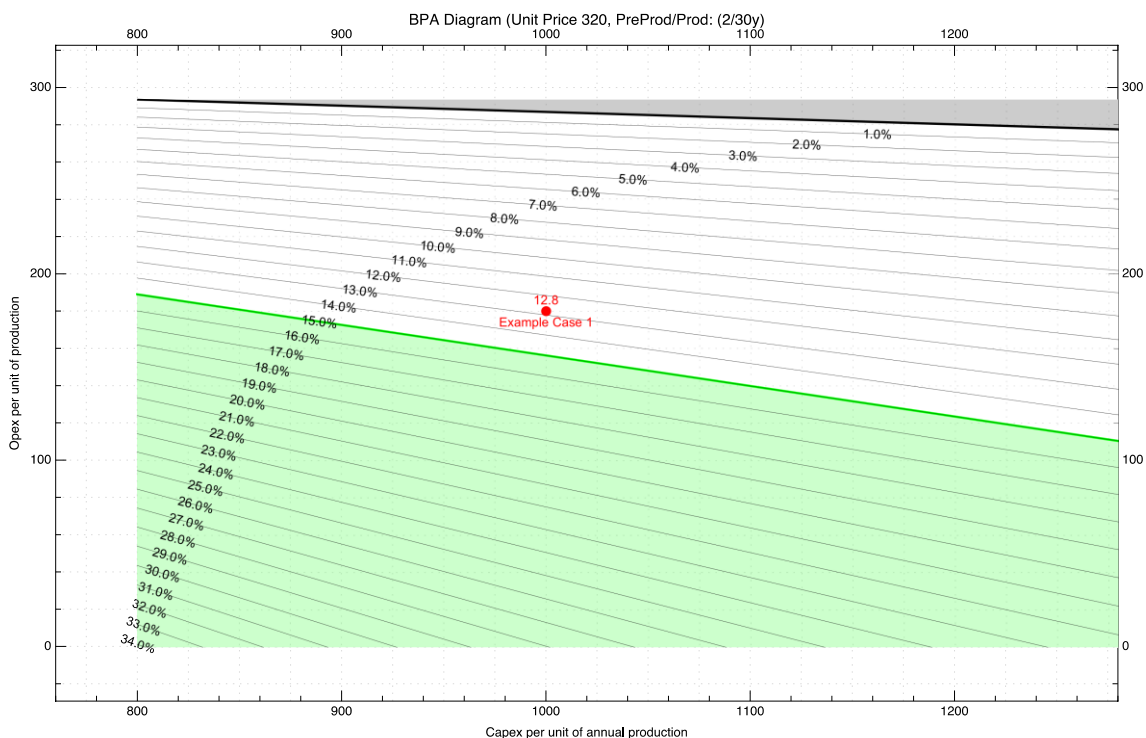


Figure 2. Exemplary Assessment of the IRR of a project.

The BPA diagram's horizontal axis shows the capex per unit of annual production, while its vertical axis shows the operating expense incurred when producing one unit. The lines with “%” labels show the resulting IRR at particular combinations of relative capex and opex per unit produced. (N.b.: Converting all operational cash flows into their net discounted amounts, BPA can also be used to deal with time variable production revenue and cost.)

BPA assumes a fixed sales price per unit produced (or the net discounted and volume weighted value of a time variable sales price), a pre-production period during which the capex will be spent, and a production period during which the plant will be operated. For the purpose of this paper, an investment case with 100% equity (no debt) and zero tax has been assumed.

Example Case 1, plotted in Figure 2, shows a hypothetical project with the following characteristics:

Capex per unit of annual production	...	1000 USD
Opex per unit (all in)	...	180 USD
Product sales price per unit	...	320 USD
Construction time	...	2 years
Production time	...	30 years
The resulting IRR	=	12.8%

Please note, that the layout of the diagram will change when basic project assumptions, such as the project duration or the sales price of the product change. See the following, alternative example for comparison with the first one:

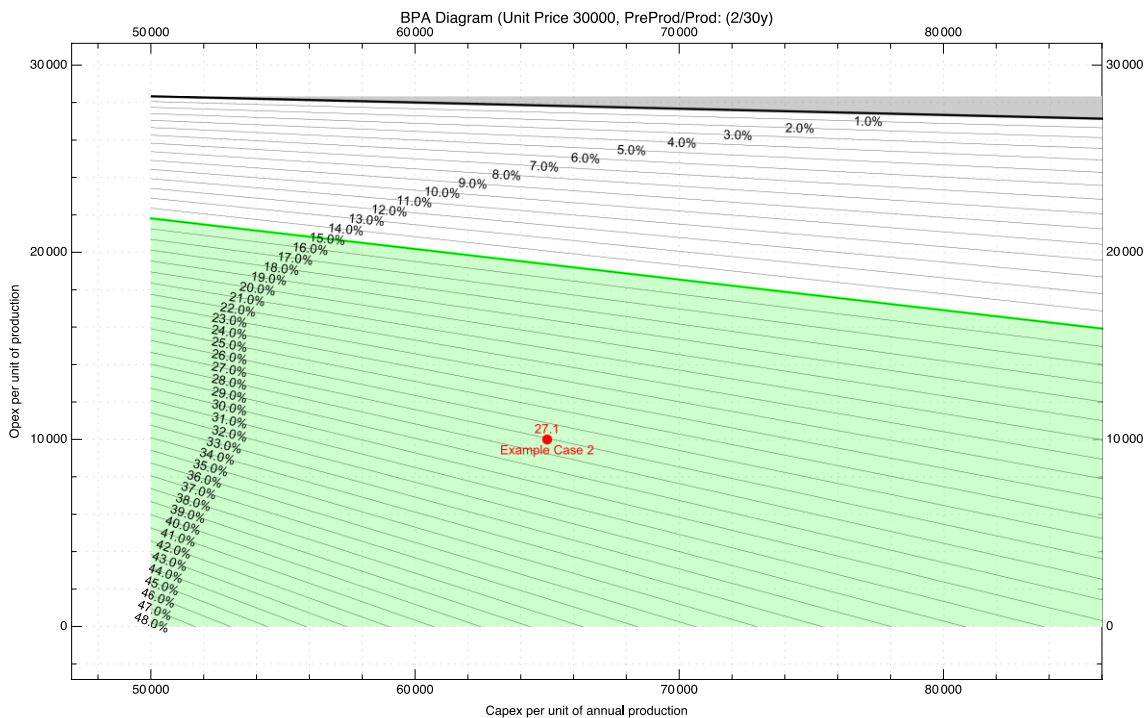


Figure 3. Exemplary Assessment of the IRR of a more profitable Project.

Example Case 2, plotted in Figure 3, shows another hypothetical project with the following characteristics:

Capex per unit of annual production	...	80000 USD
Opex per unit (all in)	...	10000 USD
Product sales price per unit	...	30000 USD
Construction time	...	2 years
Production time	...	30 years
The resulting IRR	=	27.1%

### Choosing an appropriate Formula to estimate Capex

The annual capital expenditure (capex) for any non-trivial investment project consists of various elements, which rely on certain project specific variables. For the case in point, we'll constrain most of those variables (e.g. location specific constants) to fixed assumptions and focus on the capex as a function of plant capacity (and time).

Quite naturally, different elements of the capital expenditure will occur at different times and relate to the plant capacity in different ways. For the sake of simplicity, we will assume that all Capex is spent in even installments by each end of year during the construction period of the project, and that every year thereafter a fixed percentage of the total capex is spent as sustaining capex.

$$\widehat{Capex}(c, t) = \begin{cases} \frac{\widehat{Capex}(c)}{t_{Build}}, & t \leq t_{Build} \\ \widehat{Capex}(c) \times \gamma, & t > t_{Build} \end{cases} \quad (5)$$

### Capacity Factored Capex Estimation

Capacity factored estimates are used to provide a relatively quick and - more often than not - sufficiently accurate first indicator, determining (based on capex) whether the detailed commercial evaluation of a proposed project should be continued. It can also be used to decide between alternative designs or plant sizes. This early screening method is often used to estimate the cost of complete plants within their defined battery limits and scopes, but can also be applied to individual equipment items. The estimated cost for a new plant/equipment is derived from the cost of a similar plant/equipment of a known capacity with a similarly structured bill of materials. Its most common implementation relies on the nonlinear relationship between capacity and cost as per the following equation:

$$Cost_A / Cost_B = \left( Capacity_A / Capacity_B \right)^r \quad (6)$$

The exponent  $r$  (or the capacity factor) is actually the slope of the log-curve that is drawn to reflect the relationship between actual costs and capacities of two or more completed projects. Values for  $r$  can be assigned individually in the range from 0 to 1 for each capex item, whereas 0 indicates a constant cost, 1 represents a cost proportional to the plant capacity, and a value in between represents a cost that benefits from economies of scale, that we expect from an increase in the capacity of a project. Common values for  $r$  when estimating the cost of processing plants and equipment range from 0.5 to 0.85.

For the a more generalized comparability of investment options, the relative estimated capex (the capex per ton of production capacity) is often a more convenient measure. It can be trivially calculated by dividing the estimated capex by the capacity:

$$\widehat{Capex}_{rel}(c) = \frac{\widehat{Capex}(c)}{c} \quad (7)$$

The corresponding relationship between relative estimated capex and plant capacity is shown in Figure 4.

### Limits of the Method

There are two common practical constraints, which further affect capex estimations for values of  $r$  that are neither 0 nor 1, i.e. describe economies of scale:

- (1) Some capex items (including practically all catalogue items or standardized components) may only be available in a limited number of discrete sizes.
- (2) Some capex items may have a minimum size, a maximum size or both.

Quite naturally, wherever there are economies of scale, there are limits of scalability. Thus for very large scale-ups (the definition of “very large” depending on the domain of applications), typically an increase in the capacity factors is advisable. This can be modeled into the capacity factoring formula by a correction coefficient, however for the sake of analytical approachability the present paper will resort to the basic formulae introduced above.

While there’s certainly some reward for the mathematically inclined to be gained from modeling those constraints and solving the related non-linear Knapsack problems when performing a detailed valuation exercise, we shall for the purpose of the present paper approximate the resulting consequences by sensibly adjusting the value of  $r$  for the concerned items.

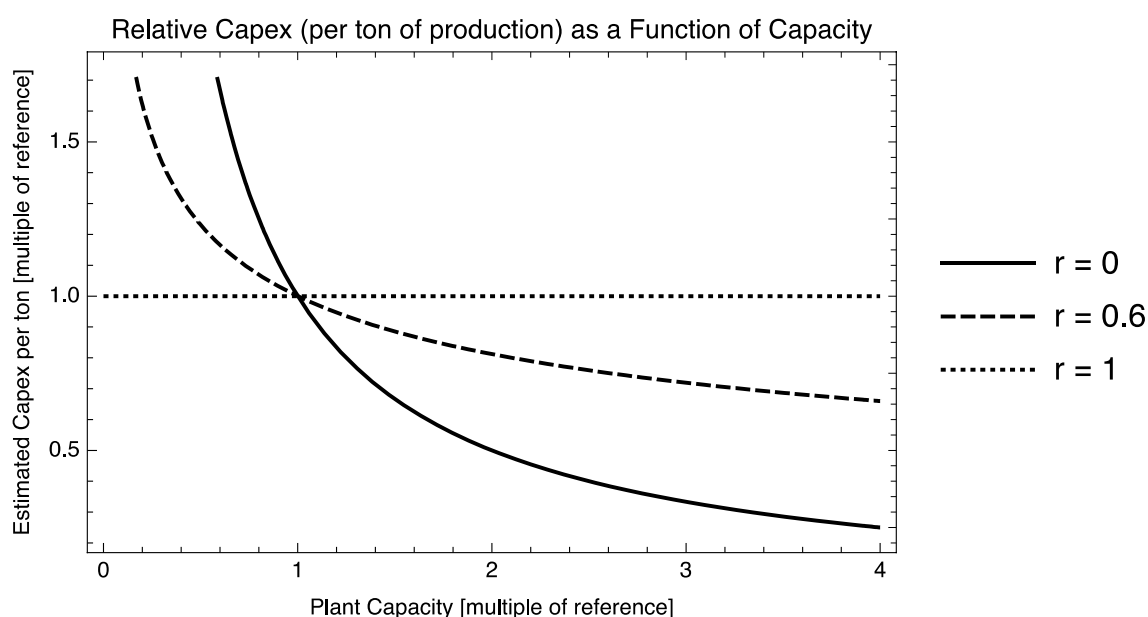


Figure 4. Relative Capex Estimation based on Capacity Factoring

### Choosing an appropriate Formula to estimate Opex

The annual operating expenditure (Opex) of an industrial facility is the sum of **fixed** and **variable** operating cost. As for the Capex (see above) economies of scale apply and larger plants tend to have lower unit production cost than smaller installations as a consequence of typically higher overall efficiency and lower unit cost following from bulk discounts on consumable and utility procurement.

As an SGA refinery typically is a large installation, operated by an entity that is expected to have explored its bulk discount potential to an exhaustive degree, we will however assume fixed numbers per unit produced. The development of adequate price elasticity models for the operative business is deliberately left to the interested reader.

N.b.: For the sake of the present exercise, all cash flows shall be given in real terms. Therefore no inflation or price escalation effects are included in above formulae.

### 3.5. The Economic Feasibility of Chloride-based SGA Production

In the following we shall use the Framework introduced in 3.4. to assess the economic feasibility of chloride based SGA production. We will do this as follows:

- (1) Develop a model for the capex per ton of annual production capacity as a function of annual production capacity
- (2) Develop a model for the opex per ton of product as a function of annual production capacity (implicitly assuming 100% plant utilization)
- (3) Plot the model results into the BPA diagram and study the resulting IRRs for different annual production capacities

### 3.5.1. A Capex Model for Chloride-based SGA Production

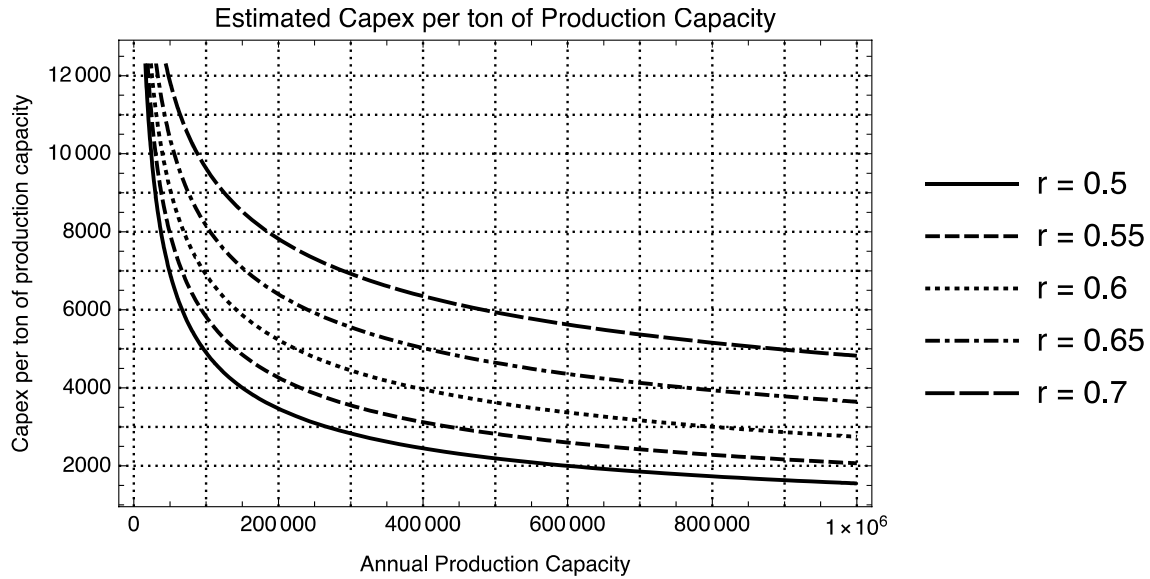
Devising a reasonably representative estimate for the capital expenditure required to build a smelter grade alumina plant based on a variation of the concept introduced above is challenging, as no such plant has ever been built, and thus no data point exists, from which such an estimate could be produced by means of extrapolation.

The only two related reference projects, known to the author of the present paper, producing Alumina from Kaolin are the high purity alumina (HPA) projects pursued by Orbite Aluminae (Canada) and Altech Chemicals (Australia). Those projects reported the following capital expenditure:

Orbite Aluminae	1000 t/a	80 Mio USD
Altech Chemicals	6000 t/a	280 Mio USD

The complexity of such a HPA plant is substantially higher than for an SGA plant, as a product purity of 4N (99.99%) as well as the maintenance of very narrow tolerances on physical product properties are required. Stripping the Altech Chemicals Plant (which is being built by SMS group, the company the author of the present paper is working for) down to produce 4.500 tons per annum of SGA instead of HPA, would yield a cost reduction by roughly 60-70 percent. Sensibly assuming cost reductions for a simplified Orbite plant (based on what is known about this plant from public announcements) would yield a price basis of 32 Mio for the Orbite plant and 92 Mio for the Altech Plant.

Fitting a capacity factor model with factors between 0.5 (very optimistically assuming substantial economies of scale) and 0.7 (rather conservative for a processing plant) to the Orbite and Altech data points via least-squares method yields a (capacity factored) capex estimate between 1.5 billion and 4.5 billion USD for a plant with an output of 1 Million tons per annum of SGA. The respective results are shown in Figure 5.



**Figure 5. Capacity Factoring for a hypothetical, chloride based SGA Plant**

Please note, that in this specific case no country/location specific construction cost factor has been applied, and that such factors may differ vastly between regions with easily available, cheap qualified construction labor, existing infrastructure and brown-field advantages vs. locations that suffer from less favorable conditions. As construction cost related factors relate to about 30%-50% of the capex, their impact on the real capital cost is not to be underestimated.

It also has to be duly noted, that scaling up by a factor in the magnitude of one hundred by capacity factoring, based on using only two available data points from the lowest 2% of the concerned range as a reference is introducing significant methodological uncertainty and potential error. For this reasons it shall remain reserved for only the most overconfident of estimators as well as for the writers of conference papers. With real money at risk, one would require substantially more data, detail and reference to get sufficiently confident in the appropriate choice of factors.

### Accounting for Uncertainty

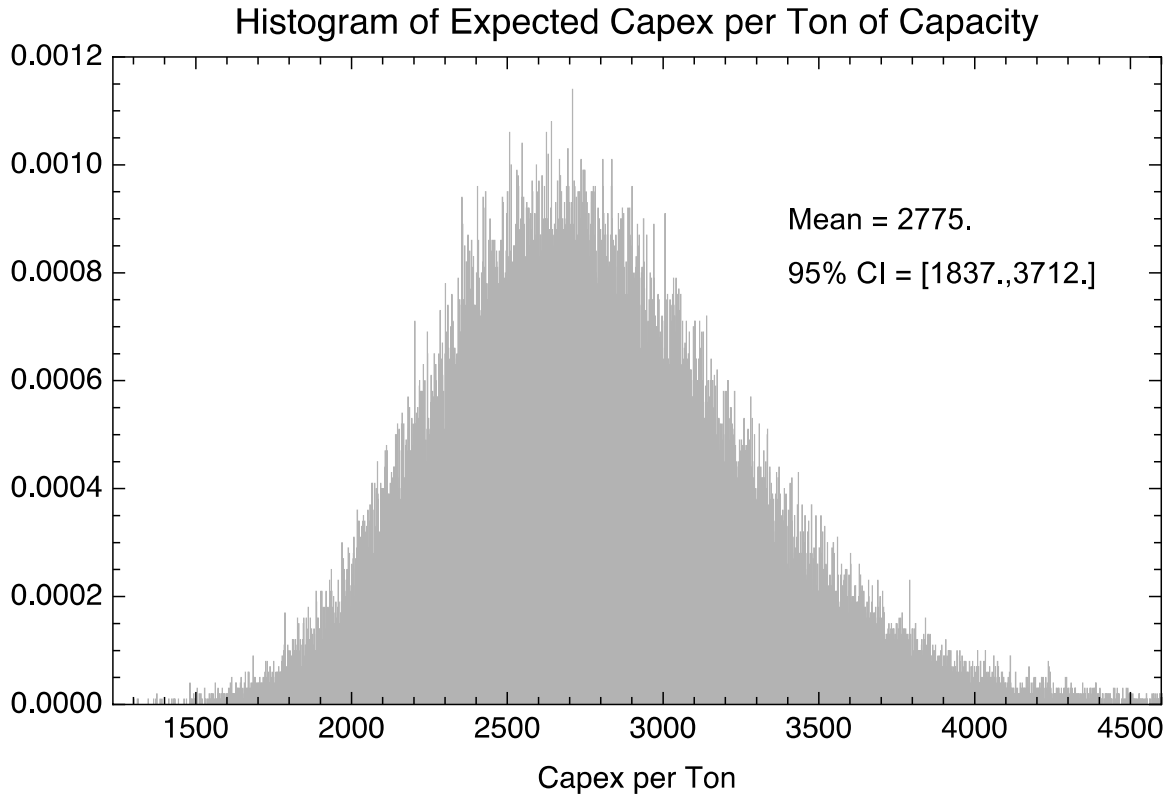
In order to render the aforementioned estimation error at least a little more tangible, we shall set up a simple stochastic model. There are three main uncertainties in above model: The two reference capex figures estimated for the Orbite and Altech case, as well as the appropriateness of the applied capacity factors.

We will in a next step assess the effects of those uncertainties on capex by assuming reasonable probability distributions for the values of the respective parameters. For simplicity's sake we'll use normally distributed probabilities and the following 95% confidence intervals:

Capex for 1000 tpa (derived from Orbite)	95% CI = 32 +/- 20%
Capex for 4500 tpa (derived from Altech)	95% CI = 112 +/- 20%
Appropriate Capacity Factor	95% CI = [0.55,0.65]

Using those (strictly subjective a priori estimates for the confidence intervals as described in [4], correlations not accounted for) parameters in a Monte Carlo simulation (results shown in Figure 6) with 1 Million samples, the estimated value for the capex for the 1Mtpa case is 2.775 Billion, at the 95% confidence interval is [1837,3712]. The relatively wide confidence interval indicates that

further research in the detailed composition of the capital structure of the respective plant is highly advisable.



**Figure 6. Resulting estimate for the Capex for a 1Mtpa chloride based SGA plant**

### Sustaining Capex

For the sustaining capex (repairs and maintenance) 2% of pre-production capex p.a. shall be assumed (based on operating data from other HCl-based metallurgical processes).

### 3.5.2. An Opex Model for Chloride-based SGA Production

The main opex components for the chloride based alumina production process are kaolin, energy (fuel and electricity), hydrochloric acid, manpower and maintenance (sustaining capex).

We (arbitrarily, but probably reasonably close to at least some real world cases) assume the following cost:

#### **Kaolin**

3.25t per t of crude alumina mining cost, assuming processing at mine site 5 USD/t

#### **Manpower**

500 base staff + 0.5 employees per 1000 ton of annual production capacity  
Average salary: USD 25.000 / year

#### **Fuel**

50 GJ per t of crude alumina 2.25 USD/GJ

**Electricity**

0.5 MWh per ton of crude alumina

4 USD / MWh

**Hydrochloric Acid**

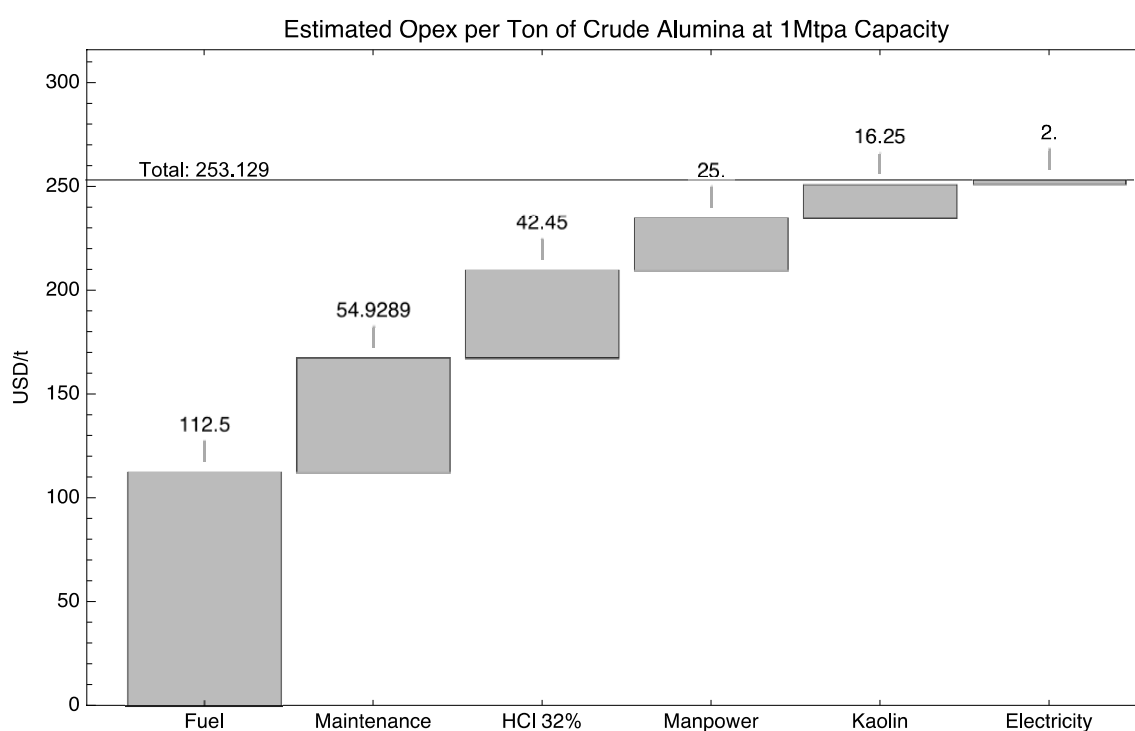
283 kg of 32% HCl per ton of crude alumina

150 USD/t

**Maintenance/Sustaining Capex**

2% of capex

variable



**Figure 7. Estimated Opex for the chloride based SGA process**

The opex estimate was done on a deliberately conservative basis, making worst case assumptions on the efficiency of certain process steps. The energy consumption of a real, large scale implementation of such a process may theoretically be about as low as 30 GJ/t (yielding opex savings of 45 USD/t), while the implementation of sophisticated acid recovery technology could reduce HCl consumption to about 150 kg/t (resulting in savings of 20 USD/t). The corresponding best case opex would ceteris paribus be around 190 USD/t.

**3.5.3. Expected IRR : Not satisfactory**

As shown in Figure 8 the expected IRR for the 1 Mtpa case is negative (negative IRRs represented by the grayed out area in the BPA diagram), and even the best case scenario (lower bound of our 95% confidence interval for the capex), accounting for a reduction in annual expenses as a matter of proportionally reduced sustaining capex shows only an internal rate of return of 2.3%.

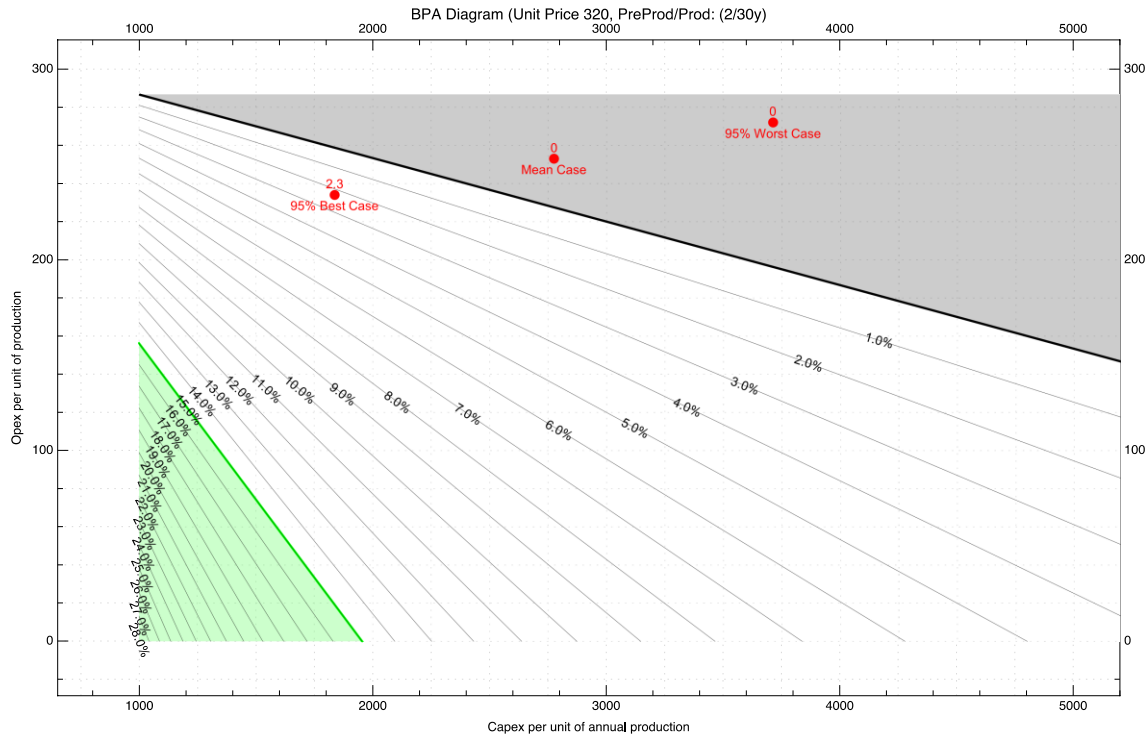


Figure 8. Assessing the Economic Attractivity of the chloride based SGA process

### 3.5.4. Expected IRR : One idea to fix it

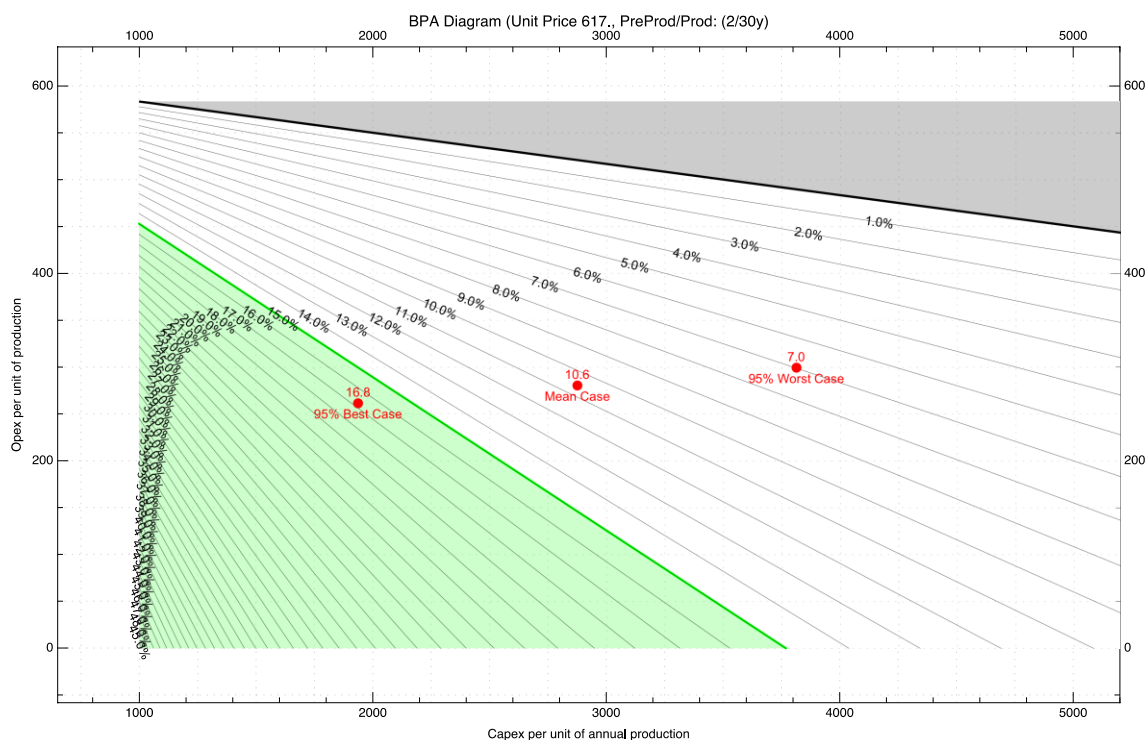
As mentioned earlier in this paper, the chloride based production rates for smelter grade alumina and high purity alumina are in many respects very similar, the later however targeting much smaller production volumes (at higher purities and better margins, as the product is readily absorbed by the fast growing electromobility and LED lighting sector). The market for HPA is growing fast (by 30% p.a. from 19.000 tonnes in 2018 to 272.000 tonnes per annum in 2028).

A company investing into a 1Mtpa chloride based SGA plant may thus reap considerable synergies by converting a part of its production into HPA.

The following assumptions shall apply:

Extra Capex for 10.000 tpa HPA capacity:	200 Mio USD
HPA opex per ton	3.000 USD
HPA sales price	30.000 USD

This results in an increase of the average sales price per ton of product (990.000 tpa of SGA and 10.000 tpa of HPA) from 320 USD/t to 617 USD/t, while the average opex per ton of product increases from 253 USD/t to 280 USD/t. The resulting expected IRR under the previous model assumptions increases to 10.6%, with the worst case yielding 7.0% and the best case yielding 16.8%. Figure 9 illustrates this improvement.



**Figure 9. Assessing the Economic Attractivity of 99% SGA and 1% HPA production**

While above assumptions are to a certain degree of arbitrary nature, they are expected to be “in the ball park”, thus strongly indicating the potential of subsidizing chloride based SGA production by the tightly integrated co-production of HPA, which could potentially be installed and operated at very attractive capex and opex, leveraging existing synergies.

### 3.5.5. Expected IRR : Improvement by increasing plant capacity

Apparently, a higher plant capacity – such as 2Mtpa - is supposed to yield a more attractive capex per ton of alumina produced. However, it appears highly likely, that plants for such extremely high capacities would consist of several parallel trains, which cannot leverage the effects of economies of scale and benefit from capacity factored price depression to the extent a single, scaled-up plant can. On another note, the capex estimate for a 1Mtpa unit as developed in the present paper is (in absolute numbers) already at the upper limit of what (with a look at their balance sheets) present-day alumina and aluminum producing companies are expected to be ready to invest into a single project.

## 4. Conclusion

As outlined in previous sections of the present paper, the idea to produce smelter grade alumina from aluminous clays such as Kaolin may have significant potential merit from the perspective of particular Alumina or Aluminum companies – especially those, who incur prohibitively high transportation cost for their feedstock. However, extrapolating a capex and opex estimate based on presently available information from high purity alumina projects, shows the potential of non-satisfactory returns into the respective investment.

It has to be duly noted, though, that the method used to derive the estimate is not based on actual equipment quotations and related front end / basic engineering does not take into account particular location-specific factors, such as the potential to substantially decrease investment cost

by placing such a plant in a location with favorable labor and construction cost or particularly well suited existing infrastructure.

On another note, the presented economic analysis is based on an SGA price of 320 USD/t. Higher prices will yield higher expected IRRs.

*It may obviously also be very worthwhile exploring (leveraging substantial synergies) to upgrade at least a part of the alumina production to HPA and sell it into the fast-growing market for high purity alumina (at the date of writing selling at 30.000 USD/t) at healthy margins, which could well support the case for the basic SGA business.*

Finally, the technical challenges posed by the – unprecedentedly for a HCl leach based operation – large scale of an installation of an industrially meaningful capacity are not to be underestimated.

The required scale-up from known laboratory test plants and/or similar plants for the production of high purity alumina shouldn't be undertaken without one or two intermediate steps of piloting and optimizing the underlying concept at different intermediate sizes while at the same time getting more certainty on the true expected capital and operating cost structure.

## 5. References

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