Optimisation and Modelling of the Production of High Purity Alumina from Lacustrine Ores

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Abstract

High-purity alumina (HPA) is in increasing demand for manufacture of LED lighting and electric vehicle batteries. Lacustrine ore from salt lakes in the south-west of Western Australia has properties conducive to low energy processing compared to existing mineral sources of alumina. However, its high iron content causes difficulties in attaining the target 99.99% purity. This project explores the production of HPA via a novel crystallization, which has potential for improved selectivity of alumina from iron. This work consists of process optimization through modelling in Aspen Plus and cost modelling. A comparison of solvent purification against the traditional HCl gas sparging route and the proposed process against competitors is provided. Net present value estimates and sensitivity analysis indicate that the lacustrine ore route to HPA shows promise to be profitable even at a low HPA price estimate and operating costs are comparable with HPA competitors. While the optimised SDC process appears inferior to HCl gas sparging at Playa One's experimental conditions, concentrating the pregnant leach solution and lowering solvent concentration can reduce costs.

1. Introduction

Valued at approximately 25 000 USD/tonne the HPA price is a significant markup on the smelter grade alumina price (500 USD/tonne). Due to market penetration of lithium-ion batteries and increased demand for LEDs, demand for HPA is forecasted to increase significantly in the upcoming decade. A deficiency in current production both in terms of capacity and cost-effectiveness presents an opportunity to gain a technological edge on existing producers. Entering the HPA market is an attractive option for prospective producers.

Several emerging producers have developed HPA processes with reported OPEX from 6000 to 10 000 USD/tonne produced (Alpha HPA, 2020; Altech Chemicals, 2018; King River Resources, 2021). The proposed process is potentially superior to existing processes with the lacustrine ore feedstock not requiring calcination or comminution before hydrometallurgical processing. The potential drawback of the lacustrine ore is a comparatively high iron content compared to other sources resulting in a high purification duty for hydrochloric acid gas sparging. This method carries environmental and safety concerns due to HCl corrosiveness.

An alternative purification route is a crystallisation method (Moldoveanu & Demopoulos, 2015) where a solvent is incorporated to the neutralisation of a pregnant leach solution to achieve a higher recovery and purity. This method has the potential to reduce the purification duty of HCl gas sparging or even eliminate the purification stage. Furthermore, incorporation of this technique can improve safety and reduce environmental impact. A potential drawback is high OPEX from solvent and energy costs. Optimisation of the crystallisation route and economic analysis is can ascertain whether there is benefit in incorporating the technique.

Figure 1 shows a simplified block diagram of the proposed HPA process. Like the mined kaolin processes, ore pre-processing occurs through 1) attritioning, 2) washing, and 3) filtration. The objective of these stages is to 1) create a homogeneous ore slurry, 2) leach sodium chloride from the ore mixture, and 3) filter out leached NaCl.

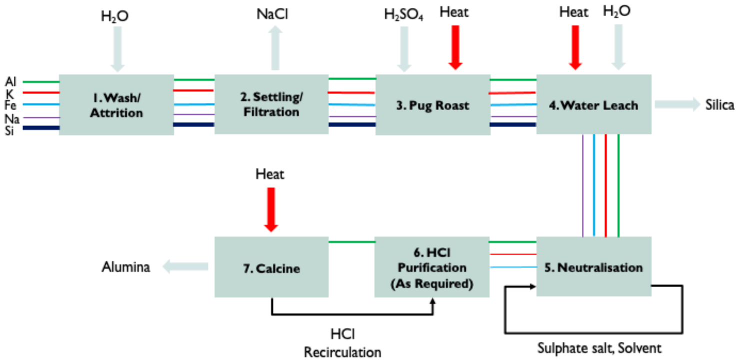

Figure 1 Simplified PFD of the proposed HPA process.

The pulp from washing is mixed with sulphuric acid at an elevated temperature in the pug roast, forming aqueous sulphates of aluminium, potassium, iron, and sodium metals in a paste. The metal sulphates are then leached in 90°C water and insoluble silicon dioxide is separated. Sulphate salt and solvent is added to the PLS and gradually cooled until the intermediate salt is precipitated. HCl gas sparging removes remaining impurities via three stages. Alumina is then formed from aluminium hydroxide by calcination.

1.1 Research Objectives

The objective of this work is to develop a flexible process model in Aspen Plus for optimisation. The model and subsequent analyses address the following:

1. Economically optimise the SDC process through process and economic modelling
2. To evaluate the feasibility of applying SDC as a purification route instead of HCl gas sparging
3. To evaluate the profitability and competitiveness of the HPA production method from lacustrine ore through process modelling and project cost evaluation methodologies

2. Process

2.1 Process Model

The process model was built to accurately determine the mass and energy balances of each stage of the process, recycle streams, and utility streams. Secondly, the process model is used to optimise reagent recycling for the SDC process and developing a heat exchange network (HEN).

2.1.1 Material Balance

All feed rates were defined by the operator's process specifications. Information to calculate stream composition values is specified by applying both experimental composition data and target values for solid liquid separation. A simplified model of the lacustrine ore feed was considered. Components were chosen to replicate the compositions of all elements which are present above a 1% threshold.

The process has numerous solid-liquid separation stages. The simulated separations were represented as target values by the operator. All chemical reactions and dissolution processes were modelled by stoichiometric reactors within the Aspen package. Extents of reactions and solubility limits of key components were determined by conversions of a reference component derived from data obtained from experiments by the operator. This approach was favoured over utilising the Aspen thermodynamic and solubility models as it more accurately represented the complex natural system.

2.1.2 Energy Balance

The mass enthalpy of each stream is calculated by Aspen based on process conditions and the calculated values are used to obtain values for heat streams. Temperatures are all determined by process specifications except for the boiling temperatures in the reagent recovery streams. The majority of component specific heat capacity and enthalpy formation data is contained in the Aspen properties database. Literature data for the lacustrine ore and intermediate salt enthalpy of formation and heat capacity was found to fill the information gaps in the simulator database. Enthalpy of dissolution values were obtained from the literature.

2.1.3 Solvent Recycle

As part of the optimisation process for SDC purification, a solvent recycle stream is incorporated. The chosen recovery strategy is boiling as this was considered the most economical option. Alternatives considered were reverse osmosis and distillation. Distillation is capitally intensive, and the level of purification achieved is not necessary for the application. Reverse osmosis for mixtures containing organic solvents and is extremely novel and requires a significant amount of detail analysis alone to assess the costs and process parameters.

2.1.4 Heat Exchange Network Design (HEN)

In the optimisation of the SDC process, a HEN design was applied. The streams with the highest heating and cooling duties were considered in this analysis for simplification and identifying the most impactful heat exchange processes. A pinch analysis is conducted as described in Sinnott & Towler (2019). The heat exchange between streams is assumed to be 100% efficient.

2.2 Economic Model

The focus of the economic analysis of this study is providing a refined estimate of process operating costs and is subject to the greatest level of detail. The capital investment is derived by an order of magnitude estimate by scaling the plant capacity against comparable processes. OPEX for materials calculations apply the reagent cost to the calculated mass flow, stream quantities derived from the process model. Energy prices are directly applied to the calculated duty including heating for modelled energy streams. Quoted prices and comparable industry data are applied to estimate the unit cost of mining, pre-leach processing, desalination, ore and product transport, and costs of administration, maintenance, labour, and services. Parameters for project evaluation include a flat tax rate of 30%, a discount rate of 8%, 5% royalties of after-tax income, and a plant construction time of 2 years.

2.3 Process Optimisation

Process design and optimisation for the novel crystallisation route applied the Onion Model heuristic (Smith, 1995). First, solvent recycling was optimised and then a heat exchange network was applied.

3. Results and Discussion

3.1 Solvent Recycling

Solvent recycling was identified to be essential to process optimisation. Costing of the process without material and energy recycling estimates solvent costs to be 7600 USD/t HPA of solvent. Boiling of the solvent was chosen to be the recovery strategy. There is a trade-off between lower reagent costs and a non-linear increase in energy costs with increasing recovery percentage. The application of boiling recovery decreases the unit cost of SDC from 7600 to 6400 USD/t HPA. This is still markedly higher than the industry cost of HCl gas sparging of 2000 USD/t HPA.

3.2 Heat Exchange Network (HEN)

Further process optimisation is achieved via design of a heat exchange network. 66% of energy consumption was from the boiling of solvent and therefore the focus of the HEN was to supply heat to this boiling process from high temperature process streams. To supplement this heat, the two hot streams with the greatest cooling duty are utilised: the solvent recondensation heat, and the pug roast cooling heat. A pinch analysis is used to design the HEN as shown in Figure 2.

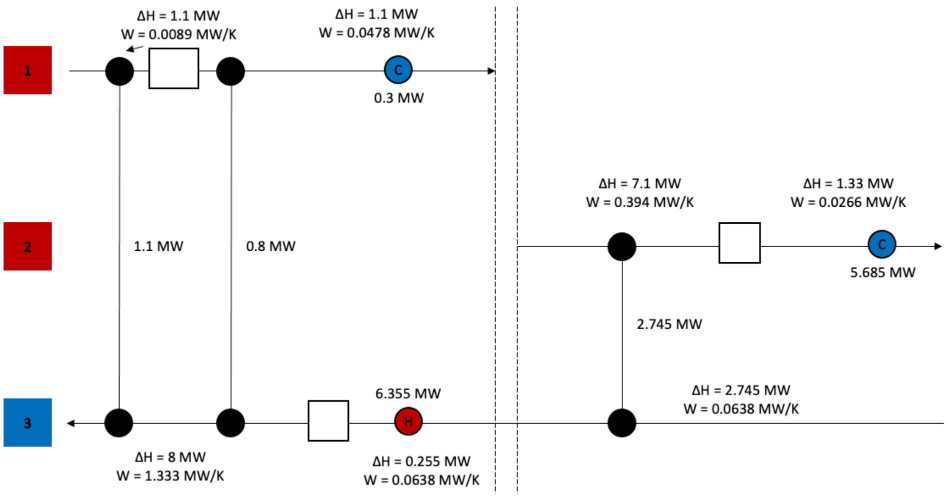

Figure 2 Heat exchange network. Stream 1 = cooling and condensation of pug roast product, Stream 2 = condensation and cooling of solvent, Stream 3 = heating and boiling of neutralisation product stream.

A total of 4.6 MW of energy is estimated to be recoverable from a HEN. This reduces the operating costs of SDC to 4200 USD/tonne.

3.3 Adjustment of Process Conditions

The total crystallisation costs are calculated at a PLS concentration ranging from 15 g/L (Playa One experiment conditions) to 40 g/L and a solvent volume fraction was reduced from 30% (experimental conditions) to 20%. The results are shown in Figure 4.

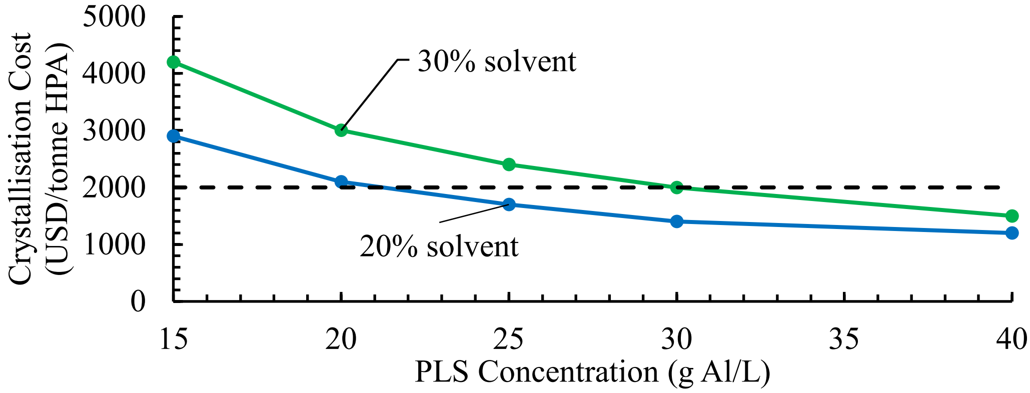

Figure 3 Heat exchange network. Stream 1 = cooling and condensation of pug roast product, Stream 2 = condensation and cooling of solvent, Stream 3 = heating and boiling of neutralisation product stream.

By revising the process conditions, the operating costs of solvent crystallisation is estimated to be reduced to 1200 USD/tonne. With an incorporated HEN, SDC becomes competitive with the costs of HCl Gas Sparging above 25 g/L PLS concentration.

3.4 Economic Evaluation

The versions of the solvent crystallisation process studied are the operator specified case (30% Solvent, 15 g/L PLS) and the most economical case (20% Solvent, 40 g/L PLS. Figure 4 displays the cumulative NPV curves of the three mentioned purification cases.

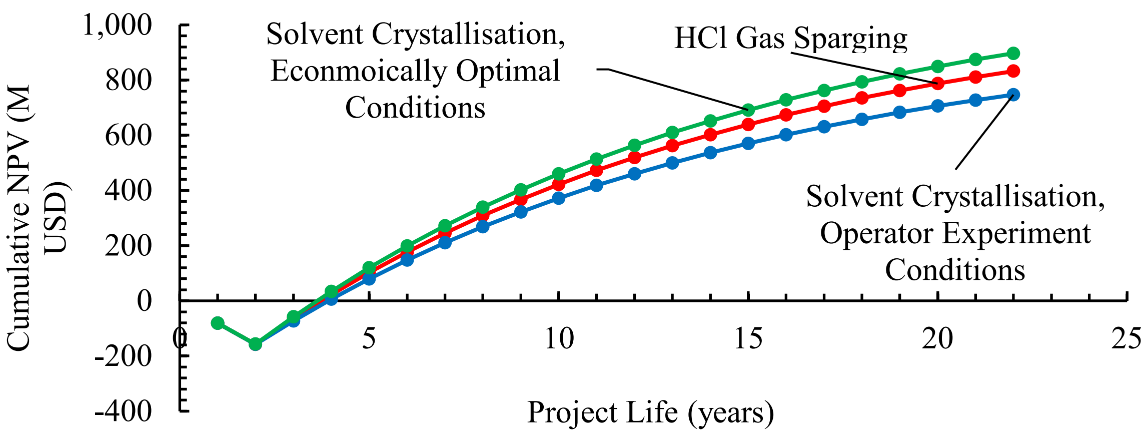

Figure 4 Cumulative NPV case study.

All cases are profitable with the HCl Purification, operator conditions, and economically optimal conditions having a payback time of 3.50, 3.65, and 3.40 years respectively and a cumulative NPV of 832, 746, and 896 M AUD respectively at an 8% discount rate. This demonstrates that the HPA route from lacustrine ore is profitable with both the HCl and solvent purification routes. Sensitivity analysis suggests that all studied cases are profitable at the minimum expected HPA price of 15 000 USD/tonne with calculated NPV values of 248, 135, and 306 M USD for HCl purification, operator conditions, and economically optimal solvent conditions respectively.

4. Conclusions and Future Work

Applying solvent assisted crystallisation with recovery by boiling and HEN integration, a unit OPEX of 4200 USD/tonne is predicted. Heat from solvent condensation heat is suggested to pre-heat the solvent solution and pug roast heat to boil the solvent. Ultimately, the optimised SDC process was not competitive with HCl gas sparging at the PLS concentration and solvent vol% in the operator's experiments. Higher PLS concentrations (40 g Al/L) and lower solvent vol% (20%) decreased the unit OPEX to 1500 USD/tonne. SDC purification becomes competitive with HCl gas sparging above a PLS concentration of 25 g Al/L at 30 vol% solvent, and above 20 g Al/L at 20 vol% solvent.

The OPEX estimate of the lacustrine ore processes is 7300 USD/tonne for HCl gas sparging, 6100-8800 USD/tonne for SDC purification. For both purification routes, this shows promise that HPA production from lacustrine ore is competitive with emerging producers, which range between 6000-10 000 USD/tonne. Calculation of cumulative NPV, and sensitivity analysis indicate that the lacustrine ore route to HPA shows promise to be profitable at a low HPA price estimate. Ultimately, HPA production via lacustrine ore appears to be an economical route to HPA and SDC shows promise to be a viable alternative for achieving the target purity.

Further work required includes a refined CAPEX estimate with equipment sizing. This gives greater insight to comparing HCl sparging with solvent crystallisation. Refining the OPEX estimate for mechanical separation units is suggested to improve the OPEX estimate. Selection of HEN equipment and associated efficiency and approach temperature will give greater insight in how heat transfer can be used to optimise the solvent process.

5. Acknowledgements

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