THE COMPARATIVE ECONOMICS OF PRODUCING ALUMINA
FROM U.S. NON-BAUXITIC ORES

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In September, 1976, the Bureau of Mines awarded a contract to Kaiser Engineers and Kaiser Aluminum as subcontractor, part of which required a technical and economic comparison between six processes for the manufacture of alumina from domestic, non-bauxitic ores.

The six processes studied were:

(1)   Clay/Nitric Acid

(2)   Clay/Hydrochloric Acid using evaporative crystallization

(3)   Clay/Hydrochloric Acid using HCl gas induced crystallization

(4)   Clay/Sulfurous Acid

(5)   Anorthosite-Lime Sinter

(6)   Alunite

This paper compares the capital and operating costs for the six processes, which were developed as part of that contract.

Fuel requirements were estimated and material balances were constructed for each process. From these the equipment requirements were ascertained. Equipment prices were obtained from vendors. Operating costs were developed from the material balances and fuel requirements.

The conclusion to be drawn from this study is that the clay/hydrochloric acid process using HCl gas induced crystallization has significantly lower capital and operating costs than the five other processes considered.

INTRODUCTION

There are two incentives to the search for a domestic ore for the production of alumina. One is that more than 90% of the bauxite used to produce alumina in the U.S. today must be imported from overseas. The second is that the cost of this bauxite has escalated as much as 400% since 1974.

In 1976 the Bureau of Mines awarded a contract to Kaiser Engineers and Kaiser Aluminum as subcontractor to compare six processes based on U.S. ores to produce alumina. The six processes were:

(1)   Clay/Nitric Acid

(2)   Clay/Hydrochloric Acid using evaporative crystallization

(3)   Clay/Hydrochloric Acid using HCl gas induced crystallization

(4)   Clay/Sulfurous Acid

(5)   Anorthosite-Lime Sinter

(6)   Alunite

This paper compares the capital and operating costs for the six processes at commercial scale.

A companion paper describes and compares the technological aspects of the processes.

The basic technical and process data for this study was taken from the literature or was provided by the Bureau of Mines from previous work and the recent operation of HCl/Clay and HNO_3/Clay "miniplants" at the Bureau of Mines engineering laboratory at Boulder City, Nevada. In addition, technical and economic information was drawn from the experience of Kaiser Aluminum & Chemical Corporation and Kaiser Engineers.

At the outset the Bureau of Mines recognized that certain process information was lacking but that those gaps would be filled by the contractor and subcontractor or suitable assumptions made in consultation with the Bureau. There was enough difference in the costs between the processes that a logical selection could be made with the information at hand.

The comparison of capital and operating costs for the six processes are shown in relative terms by taking the HCl/Clay process with gas induced crystallization as the base case and showing the difference in costs between the other processes and the base case.

METHODS USED TO ESTIMATE COSTS

The cost of construction and operation of a 500,000 ton per year alumina plant plus mining operation was estimated for each of the six processes.

Method Used To Develop Mining Capital Costs

Mining capital costs in Table 1 have been assembled for each ore based on the required annual tonnage. Overburden removal is required for clay and alunite, but not for anorthosite and limestone. Similar or identical equipment has been used throughout and equipment costs are based on recent user experience. Equipment operating capacities have been obtained by reducing manufacturer's design ratings by an appropriate amount based on recent user experience. The capital estimates include both mobile equipment and fixed facilities such as shops, utilities, roads, and fuel storage. Separate mining facilities have been provided for anorthosite and limestone, both of which are used in the anorthosite process. The values shown in the capital cost table represent the difference in mining cost between each process and the base case process.

Method Used To Develop Process Plant Capital Costs

The general method used to develop capital costs was as follows:

(1) Develop a material balance for each process. These balances are shown in the appendix together with the assumptions made.

(2) Develop an equipment list for the process based on the material balance. This equipment list specified the size and number of the various pieces of equipment taking into account the need for spare units to handle equipment breakdown and cleaning.

(3) Obtain quotes from vendors for all major process equipment items.

Values for process equipment in Table 1 represent the difference in installed equipment cost between each process and the base case in dollars per annual ton. Other direct capital cost differences for foundations, structures, piping, utilities, electrical, etc., are developed from percentage factors based on experience in alumina and other metallurgical processes. Indirect capital cost differences include such items as salaries, burden, overhead, personnel expenses, and office expenses. Freight costs have been included in indirect costs.

TABLE 1: CAPITAL COST COMPARISON

	Capital Cost Differences, $/A Ton Al$_2O_3$ *					
	HNO$_3$	HCl Evap.	HCl Sparging**	H$_2$SO$_3$	Anorthosite	Alunite
Mining Capital	-0-	-0-	Base Case	-0-	40.74	.08
Process Plant Capital						
Process Equipment	108.21	43.54	Base Case	196.69	151.20	93.57
Other Direct Capital (Foundations, Structures, Piping, Utilities, Electrical, Etc.)	122.33	58.41	Base Case	217.88	81.74	149.25
Indirect Capital and Freight	52.51	22.52	Base Case	94.48	53.14	55.44
Waste Disposal Capital	3.00	-0-	Base Case	-0-	36.20	1.00
Working Capital	3.68	.35	Base Case	3.89	6.68	7.24
Total Capital Cost Difference, $/A Ton	289.73	124.82	Base Case	512.94	369.70	306.58

Note: Costs shown represent difference in capital dollars between process and base case. Positive values represent higher capital requirements. Costs are calculated based on plant sized to produce 500,000 ton/yr of alumina, and reflect differences in capital spending for each ton of alumina capacity.

* $/A Ton: Dollars/Annual Ton of Al$_2$O$_3$

** Sparging: HCl gas-induced crystallization process.

(1) As of July, 1977.

The process plant capital costs exclude contingencies, escalation, and owner's costs.

Method Used To Develop Waste Disposal Capital Costs

The capital cost differences shown for waste disposal in Table 1 represent cost differentials for facilities to impound solid residues remaining after alumina removal from the ores. In all cases except anorthosite, it was assumed that the mined out area created by ore removal would be used for mud disposal. Due to the large volume of solids with anorthosite, levee construction is required. Generally the capital costs are for pumps, piping, sumps, and electrics necessary to deliver slurry to the disposal area and to recycle liquor back to the plant.

Method Used To Develop Working Capital Costs

Working capital cost differences shown in Table 1 represent the differences in costs required to maintain adequate inventories of materials, supplies, and product. Materials inventories were set at 10 days, supplies at 30 days, except oil at 10 days and coal at 45 days, and product at 5 days. An exception was the alunite byproduct, sulfate of potash, for which 45 days is used due to the highly seasonal nature of demand. All items are charged at their cost of production or purchase.

Method Used To Develop Operating Costs

Table 2 presents operating cost comparisons for all six processes considered. The comparative costs are presented as the difference, in cost per ton of alumina produced, between each process and the "base case". The HCl/Clay process using HCl gas sparging to precipitate salt is used throughout as the base case since it has the lowest total operating cost per ton of product.

The operating costs are grouped into six cost elements; a seventh element is listed for by-product credits. The first element is "Ore Cost". Unit prices for ore are based on direct and indirect mining costs including labor, repair and maintenance supplies, overhead, royalty payments, and preproduction expense. Not included are depreciation, taxes, insurance, and certain overhead functions provided by the process plant organization. All ore-mining costs are calculated on a directly comparable basis. The unit cost of clay is identical in all clay cases. Mine-to-plant haul distances is assumed to be five miles for all ores. The cost of overburden removal is significant only for clay.

Each of the six processes is dependent on an acid or base

TABLE 2: OPERATING COST COMPARISON (1)

	Operating Cost Differences, $/Ton Al$_2O_3$					
	HNO$_3$	HCl Evap.	HCl Sparging*	H$_2$SO$_3$	Anorthosite	Alunite
Ore (Excluding Mining Capital)	.38	.37	Base Case	2.74	-4.28	5.38
Reagents (Acids & Bases)	21.65	.87	Base Case	5.44	10.43	64.38
Utilities (Oil, Coal, Power, Water)	23.03	19.45	Base Case	16.37	43.09	.01
Labor (Operating, R & M, Supervision, Administration)	5.80	1.88	Base Case	5.42	.79	3.11
Supplies (R & M, Operating, Processing)	13.31	2.67	Base Case	-.22	-4.05	-1.71
Other	3.40	1.51	Base Case	6.11	3.44	3.58
Subtotal	67.57	26.75	Base Case	35.86	49.42	74.75
By-Product Credit (K$_2$SO$_4$ & H$_2$SO$_4$)						-62.63
Total Difference, $/Ton Al$_2O_3$	67.57	26.75	Base Case	35.86	49.42	12.12

* Sparging: HCl gas-induced crystallization process.

Note: Costs shown represent difference between each process and the base case. Positive values represent higher costs; negative values represent lower costs. Costs are calculated at 500,000 ton/yr rate and reflect differences per ton Al$_2$O$_3$ produced.

(1) As of July, 1977.

to extract the alumina from the ore. The cost differential for these materials is reflected in the "Reagents" element. Unit costs have been obtained from potential suppliers, with the exception of limestone for the anorthosite process which is priced on the basis of mining it within five miles of the plant. Material usages have been based on the process material balance. Freight effects are considered only to the extent that alunite and anorthosite plants have been assumed to be West of the Rockies and all the clay process plants in the S.E. U.S.

The third element of comparative operating cost is "Utilities". The differences shown indicate primarily the various energy requirements for the individual processes. In all cases the cheapest practical fuel commensurate with the required product purity has been used. Therefore, coal is the fuel of choice except when direct contact with the product is required in the acid processes. In those situations requiring contact, No. 6 fuel oil is used. Natural gas is not specified for any process as the availability is questionable. A single unit cost and grade has been used for coal and fuel oil in all cases. Fuel usages are based on the process energy requirements with comparative allowances for efficiency and heat losses. A process energy comparison is shown in the appendix.

In the "Labor" element of operating cost repair and maintenance (R & M) labor cost is the most variable. R & M has been calculated as a percentage of direct process plant capital with a higher rate for highly corrosive processes (HNO_3 & HCl) and a lower rate for non-corrosive processes (alunite and anorthosite) with sulfurous acid at an intermediate level. The operating labor requirements have been compared by estimating the manpower requirements for each section of each process. Supervision and administration are calculated assuming a fixed administrative staff size and a constant ratio of supervisors to operating labor and R & M labor.

The "Supplies" element contains repair and maintenance materials, operating supplies such as gas and oil, small tools, filter cloth, etc., and processing supplies such as flocculant, lime, chlorine, etc. Most of the differences between processes in this element of cost arise from R & M materials which are based on capital cost and process corrosivity.

The "Other" element represents taxes and insurance. It is calculated as a fixed percentage of capital for all processes. No differences are implied based on location.

The alunite process produces sulfuric acid and sulfate of potash as by-products. Since H_2SO_4 production is large (1,500 ton/d) and there is minimal demand in Utah, the acid would have to be either shipped out or converted to another product on site. For this study, it is assumed H_2SO_4 would be shipped to Houston, Texas and sold on a freight-equalized basis. Sulfate of potash also presents a marketing problem. The proposed plant would be the largest producer of this product in the U.S. Since demand for the product is seasonal and depends on the state of the

agricultural economy, problems of moving the entire plant output are to be expected. For this study, an intermediate price between today's prices for sulfate of potash and muriate of potash has been assumed. For simplicity, the full sales price for the potash is given to the plant as a credit with no deductions for selling trade allowances, G & A, and overhead costs, although some costs would be incurred for the functions.

The comparative total operating cost difference represents a totaling of the seven elements of differential cost. These values represent the total additional cost of producing one ton of alumina for each process compared to the base case.

DISCUSSION OF CAPITAL AND OPERATING COST COMPARISONS

This section will discuss the major cost differences between the processes and the reasons behind these differences.

Base Case: Alumina from Clay via Hydrochloric Acid Extraction - HCl Precipitation

The base case has the lowest operating and capital costs among the six processes for several reasons.

(1) The leach liquor evaporation load is relatively low due to recycling the crystallization mother liquor back to leaching, rather than having to evaporate all of the water therein.

(2) The cost of ore and other reagents is generally lower than for other processes.

(3) The crystallization step is based on HCl gas sparging and has minimal evaporative load.

(4) The heat of decomposition of $AlCl_3 \cdot 6H_2O$ is relatively low compared to the nitrate, but higher than $Al(OH)_3$.

(5) 90% of the $AlCl_3 \cdot 6H_2O$ decomposition is done by indirect heating permitting the use of coal, which is cheaper than oil.

(6) The indirectly heated calcination facilitates the subsequent condensation of the decomposition off-gases due to the absence of fuel combustion products, resulting in considerable cost savings.

(7) The iron removal step is relatively simple and effective. This is equally true for both HCl processes.

Alumina From Clay Via Nitric Acid Process

Comparison of Operating Costs to Base Case: Table 2

(1) Reagent Costs

Reagent costs for this process are $21.65 per ton of alumina greater than the base case. The reason for this difference is the relatively high usage of nitric acid by this process in its assumed form. Nitrate losses occur in the solvent extraction section and in waste solids washing. An additional nitrate loss results from decomposition of nitrate and NO_x to elemental N_2 at elevated temperatures during the decomposition of $Al(NO_3)_3$. $9H_2O$. The cost of nitric acid is expected to increase in the future with the cost of ammonia, which is expected to increase with probable escalation of hydrocarbon feedstock costs. Nitrate losses in the thermal decomposition section possibly could be reduced, but an extended development effort would be required.

(2) Utilities

Utility costs are $23.03 per ton of alumina greater than the base case. Almost all of this is due to the greater fuel requirement for the HNO_3 process.

The major components of this greater fuel requirement are:

(a) The thermal decomposition of $Al(NO_3)_3$. $9H_2O$ which requires approximately 50% more heat per unit of alumina than is required for $AlCl_3$. $6H_2O$.

(b) A relatively high steam load is required for leach liquor evaporation and for regenerating the HCl used in iron removal.

(3) Labor, Supplies, and Other

These three items are all higher for the nitric acid process relative to the base case. Maintenance labor and materials, and taxes and insurance, which compose part of these three items, have been estimated as a fixed percentage of the capital cost, and the capital cost for this process is considerably higher than for the base case.

Comparison of Capital Costs to the Base Case: Table 1

The total plant capital costs are $289.73 per annual short ton greater than the base case.

Equipment costs for the salt decomposition system in the nitric acid process are more than double that for the base case. One reason is that provision must be made for the transfer of approximately 50% more heat than in the base case. Another reason is that the nitrate system uses the dense fluid bed mode, with lower gas velocities, while the chloride decomposition system can use an expanded bed mode with higher gas velocities. This effect plus higher gas evolution from decomposition of the $Al(NO_3)_3$, $9H_2O$ necessitates the use of more fluid beds for the nitric acid process.

The steam plant costs are about twice those for the base case because of the need for a large amount of steam to regenerate HCl used in iron removal and because of substantial process steam evaporation requirements.

Alumina from Clay via Hydrochloric Acid Extraction - Evaporative Crystallization

Comparison of Operating Costs to Base Case: Table 2

Utilities

The cost of utilities is $19.45 per ton greater than for the base case for the following reasons:

(a) The use of 20% acid for leaching in this process, as compared with the use of 26% leach acid in the base case, increases the evaporation required to bring the leach liquor to saturation with respect to $AlCl_3$.

(b) In this process, as defined, all of the mother liquor from the crystallization stage must be vaporized. In the base case the mother liquor becomes leach acid without vaporization and is recycled directly to leaching. This difference results in the use of considerably more steam in the HCl evaporative crystallization process.

(c) The evaporative crystallization approach employs direct-fired thermal decomposition of the $AlCl_3$. $6H_2O$. Oil has therefore been assumed to be the source for 100% of the energy requirement for the thermal decomposition as compared with only the

last 10% in the base case. The added cost of oil in lieu of coal for 90% of the thermal decomposition energy requirement plus the cost of the added evaporation are responsible for the higher cost of utilities for this process in comparison with the base case.

If the $AlCl_3 \cdot 6H_2O$ decomposition system were converted to the indirect-fired mode it would be possible to save about $6 per ton of alumina in fuel costs. Conversion, if the technology were available, is estimated to add approximately $30 million to the capital cost for this section. For this reason, simply changing the calcination mode does not appear worthwhile, and in fact the indirect decomposition mode becomes advantageous only when incorporated in the HCl gas induced crystallization process as defined in the base case.

Comparison of Capital Costs to the Base Case: Table 1

Capital costs for the HCl-evaporative process are the second lowest at $124.82 greater than the base case. Reasons for the differences are:

(1) Evaporation requirements to bring the purified leach liquor to saturation are substantially higher in the evaporative crystallization process than in the base case with a consequent increased capital requirement.

(2) The capital requirements for the crystallization section of the evaporative crystallization process is more than double that for the base case due to the large amount of evaporation required in this section and the large volume of vapor to be removed and condensed. In the base case there is no evaporation in this process section and only a relatively small amount of heat is rejected.

(3) The salt decomposition section for this process requires only about one-third the investment required for the base case because of provision in the base case for a large amount of corrosion-resistant heat transfer surface and a molten salt heating and circulating system.

(4) Acid recovery, in contrast, requires in this process a capital investment approximately four times that for the base case. The increase is due to the large amount of inert combustion product gases mixed with the hydrochloric acid to be condensed.

(5) The investment required in the steam plant for the evaporative crystallization process is almost double that for the base case because of the greater evaporation load.

Alumina from Clay via Sulfurous Acid Extraction

Comparison of Operating Costs to Base Case: Table 2

(1) Reagent Costs

One would expect sulfurous acid to be an inexpensive leaching reagent. Unfortunately, sulfur in the +4 valence state may be oxidized or reduced easily. Both of these processes occur, and the sulfur so inactivated must be replaced at substantial cost.

The process requires makeup caustic for the Bayer section, and this requirement combined with makeup sulfur brings the total primary reagent cost to a figure substantially higher than for the base case. Usages of these reagents have been extensively studied in the past, and there is believed to be little potential for any substantial reduction.

(2) Utilities

Utilities costs for this process are higher than the base case because of the need for large amounts of steam in the autoclaving sections of the sulfurous acid process. In addition, energy is required for the Bayer plant section of this process which has no equivalent in the base case.

(3) Labor and Other

These costs are higher than the base case because the maintenance and taxes components of these costs is based on equipment costs which are considerably higher for this sulfurous acid process.

Comparison of Capital Costs to the Base Case: Table 1

This process has the highest capital cost of all six processes at $512.94 greater than the base case. The main reasons for this difference are:

(1) Fifteen hours holding time under pressure are required to extract a reasonable amount of alumina from clay. This requires a very large and costly volume of pressurized reactors.

(2) The process requires both a sulfurous acid extraction plant and a modified Bayer plant to produce a suitable product which increases equipment requirements greatly.

(3) A sulfurous acid preparation system is included in the plant cost.

Alumina from Anorthosite via Lime Sinter Process

Comparison of Operating Costs to Base Case: Table 2

(1) Reagent Cost

The anorthosite-lime sinter process converts all calcium in the anorthosite to a calcium silicate and most of the aluminum to a calcium aluminate. The limestone requirement for this is high, i.e., 2 tons of limestone per ton of anorthosite, resulting in a reagent cost considerably higher than for the base case.

(2) Utilities

The cost of utilities for the anorthosite process are $43.09/ton alumina higher than the base case.

A substantial amount of electrical power is required in the anorthosite-lime sinter process for the grinding of anorthosite ore and limestone, but the overwhelmingly dominant cost element is coal for the sintering operation. There is little prospect for any substantial reduction of this requirement.

Comparison of Capital Costs with the Base Case: Table 1

Capital costs for the anorthosite process are second highest in the group of processes compared.

The mining capital cost shown for the anorthosite process is based on the capital requirements for quarrying both anorthosite and limestone. The large capital costs are due to the mining operations to produce the two materials and the large amounts of both that are required to produce a ton of alumina.

The numerous process steps, the large mass of solids involved, and low alumina concentrations in leach liquor result in a high process plant capital cost.

The waste disposal capital cost for this process is the highest of the group because of the addition of large quantities of limestone to the process. This results in the largest solid waste residue load of all the processes.

Alumina from Alunite/Reduction Roast - Bayer Extraction

Comparison of Operating Costs to the Base Case: Table 2

(1) Reagent Costs

Reagent costs for the alunite process are very high, at $64.38/ton of alumina greater than the base case, for the following reasons.

(a) Caustic potash for leaching unreduced sulfate associated with aluminum is by far the largest single reagent cost item in the manufacture of alumina from alunite via reduction roasting. It is possible in further development of the process that a way may be found to more effectively reduce sulfate associated with alumina without affecting so severely the subsequent solubility of the alumina in caustic; but the probability of this is not considered high in view of the large amount of previous work having this goal. Alternatively, NH_4OH may be used to solubilize the unreduced sulfate, but in this case either a mixed $(NH_4)_2SO_4K_2SO_4$ must be marketed or NH_3 must be regenerated by adding lime with subsequent rejection of $CaSO_4$. Neither approach is attractive, and a major reduction in the cost of reagent for dealing with unreduced sulfate is believed unlikely.

(b) A second major reagent cost is that of makeup caustic soda for the Bayer processing section. A small amount of caustic soda is lost with the product alumina. A larger amount is lost with the considerable amount of solid waste remaining after alumina extraction, but by far the largest amount is lost by reaction with reactive silica and kaolin in the ore.

(2) By-Product Credit

The alunite process is the only one in the group which produces products other than alumina. For every ton of alumina produced one ton of sulfuric acid and 0.73 tons of potassium sulfate are also

produced. There may be some difficulty in selling the large quantities of by-product produced. However, this study assumes that all the sulfuric acid can be sold in the southwestern U.S. and that the potassium sulfate can be sold as fertilizer.

Comparison of Capital Costs with the Base Case: Table 1

Capital costs for this process are third highest in the group.

The processing of alunite requires capital provision for a large number of process steps including a sulfuric acid plant, a potassium sulfate plant, and a modified Bayer plant. These operations are relatively large scale, because the entering ore is low grade with respect to alumina. The Bayer process section is required to leach and then wash a much larger amount of solids per ton of alumina than is the case in a Bayer plant operating on bauxite because of the low alumina content. The low alumina content in the ore, a relatively low overall alumina extraction efficiency, and the number of steps required to separate the components of the ore result in a process plant capital cost much higher than the base case.

CONCLUSION

In order to come to a valid conclusion in a comparison study of this type, it is necessary to estimate the costs on the same basis for all the processes, and to examine the sensitivity of the result to possible variations in process parameters.

A conscious effort was made to estimate both capital and operating costs on the same bases and using the same estimating group for all six processes. For example, the same bases were used for such items as equipment selection and sparing, working capital, work-force estimates, fuel costs, and the factors used for piping, instrumentation, foundations, electrical installations, and others.

The conclusion drawn from this study was that the Clay/Hydrochloric Acid process using HCl gas induced crystallization has the lowest capital costs and the lowest operating costs of the six processes studied.

The sensitivity of the result was examined by adopting significantly more advantageous process design assumptions for the five less economic processes, to find out if, by doing so, one of them could possibly become the most economic process.

In no instance did this happen, which leads to the supporting conclusion that, in addition, the probability of one of the other processes becoming the most economic is small.

APPENDIX

The appendix starts on the next page.

NITRIC ACID/CLAY PROCESS

Process flow diagram (line art). Process units and numbered streams:

- CLAY —[1]→ CRUSHING —[2]→ CALCINATION (FUEL; [3]→ STACK) —[4]→ GRINDING —[5]→ LEACHING (FUEL) —[6]→ THICKENING —[7]→ FILTRATION —[8]→ SOLVENT EXTRACTION —[9]→ EVAPORATION —[10]→ CRYSTALLIZATION —[11]→ CENTRIFUGE
- THICKENING —[34]→ WASHING (WATER [36]; WASTE [37]); [35]
- FILTRATION: WATER [38]; WASTE [39]
- SOLVENT EXTRACTION: ORGANIC [40]; [28]/[29]/[30]; ACID STORAGE [32]/[33]
- NITRATE WASHING (WATER [42]; WASTE [43]) —[44]
- FeCl₃ STRIPPING (HCl [54]) —[47]/[53]
- CHLORIDE WASHING (WATER [48]; WASTE [50])
- HCl REGENERATION (H₂SO₄ [51]; WASTE [52]) —[49]
- CENTRIFUGE —[12]→ CRYSTAL MELTING —[13]→ DECOMPOSITION —[14]→ —[16]→ ALUMINA; [31]→ STACK; [15]→ CONDENSER [26] → STACK [27]; ACID [55]
- CRYSTALLIZE/EVAPORATION —[19]→ CRYSTAL MELTING —[21]→ DECOMPOSITION —[23]→ ; WASTE [22]; BLEED STREAM [18]; [17]/[20]; FUEL
- CONDENSER [24] → STACK [25]

Alumina from Clay/Nitric Acid Process — Material Balance
1,000 TON/D Al₂O₃

All units are in short tons per day.

Component	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28
Al_2O_3	1131	1131	12	1119	1119	56								986		1000												
$Al(NO_3)_3$						4441	4397	4395	4351	6354	6242						2003											
$Al(NO_3)_3 \cdot 9H_2O$											2153	7396	7396															
$AlCl_3$																												
$Al(OH)_3$																												
Fe_2O_3	27	27		27	27	9																	114					
$Fe(NO_3)_3$						53	53	53				195	195					150	150		150							
$Fe(OH)_3$																												
$FeCl_3$																		435	338		338		338		319			
SiO_2	1436	1436	14	1422	1422	1422	12	6	6																			
LOI	402	402	402	83	83	83	7							148														
Other	699	699	699							73	73					1	67											2
H_2O	84	84	1			4736	6350	6353	6339	9437	7396				3032		2990	224	221		221		221			3732		854
HNO_3						185	183	183	229	3214	3214	195	195		3326		5807											
N_2O_5												1	1	1	195													
HCl																		5	5	3	5	5				2821		
Organic																				97								
CO_2																												
O_2															73												73	
N_2															26												26	
H_2SO_4																						36			354			
Total	3779	3779	1128	2651	2651	10985	11002	10990	10925	19078	19078	7787	7787	1135	6652	1001	10867	814	714	100	714	41	673	—	673	6553	99	856

Alumina from Clay/Nitric Acid Process — Material Balance

Component	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45	46	47	48	49	50	51	52	53	54	55
Al_2O_3																											
$Al(NO_3)_3$																											
$Al(NO_3)_3 \cdot 9H_2O$																											
$AlCl_3$																											
$Al(OH)_3$																											
Fe_2O_3																											
$Fe(NO_3)_3$																											
$Fe(OH)_3$																											
$FeCl_3$																											
SiO_2																											
LOI						1410	1410		1410																		
H_2O	2705	1851	63	76	4167	2066	3081	3361	3598	17				3481	3481	2811	2073	2073	2811	6108	2810	6108		205	1869	204	195
Other																											
HNO_3	5		5	76	4167																						
N_2O_5															14												
HCl						1467																					
Organic																											
CO_2																											
O_2			49																								
N_2		3	17																								
H_2SO_4																											
Total	2710	1854	134	152	8334	5195	5212	3361	5198	17	25	1	2076	3481	3495	2662	2495	2529	2928	6108	2810	6126	52	346	2235	260	390

HYDROCHLORIC ACID/CLAY PROCESS — EVAPORATIVE CRYSTALLIZATION

Process flow diagram with the following unit operations and streams:

- CLAY → [1] → CRUSHING → [2] → CALCINATION (FUEL; STACK [60]) → [3] → GRINDING → [4] → LEACHING → [7] → THICKENING → [8] → WASHING (WATER [9]; WASTE [10])
- LEACHING (FUEL [5]; HCl CONDENSER [6])
- THICKENING → [11], [12] → FILTRATION → [14] → WASTE; [13] →
- SOLVENT EXTRACTION (Cl_2 [16]; ORGANIC; [15])
- IRON STRIPPING → [52] → EVAPORATION → [54] / [55] / [56]; $FeCl_3$ DECOMPOSER (COMBUSTION WATER, FUEL [58]; WASTE [59])
- EVAPORATION [17] → EVAPORATION [18] → CRYSTALLIZATION [19] → SEPARATION & WASHING [20]
- DECOMPOSITION AND CALCINATION → ALUMINA [23]; COMBUSTION WATER [22]; STACK [24]; [25]
- HCl CONDENSER [27] (STACK), H_2O [26]
- HCl STRIPPING [31] / [43] / [44]
- ABSORPTION [33] / [49]
- WASTE CHLORIDE DECOMPOSITION (FUEL; COMBUSTION WATER [48]; H_2SO_4; WASTE [50])
- HCl CONDENSER [39] / [40] / [41] / [47]
- ACID [29]; FUEL

Alumina from Clay/Hydrochloric Acid Process
Evaporative Crystallization
Material Balance
1,000 TON/D Al₂O₃

Component	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22
Al₂O₃	1127	1127	1116	1116			56	56		56												
AlCl₃							2776	557		27	530	2749		1	2748		2738	2738	1506	1392	20	
AlCl₃·6H₂O									4719										4746		4746	
Fe₂O₃	27	27	27	27																		
FeCl₃					2	2	52	11			11	52			52							
HCl							229	46			46	229			229		240	149	1971	1822	268	
L.O.I.	418	418								1411				8								
H₂O	701	701			198	198	10734	2491		3147	4063	12306	6	6	12306		12306	2691	7523	6954	549	983
Other	84	84	83	83			83	78		77	8	6			6		6	6	74	68		
SiO₂	1433	1433	1419	1419			1419	1419														
Organic							1	1	1	1	1	8									1	
H₂SO₄																11						
Cl₂																						
Total	3790	3790	2645	2645	200	200	15350	4659	4720	4720	4659	15350	6	15	15341	11	15290	5584	15820	10236	5584	983

Alumina from Clay/Hydrochloric Acid Process
Evaporative Crystallization
Material Balance

Component	23	24	25	26	27	28	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44
Al₂O₃	1000							94														
AlCl₃																						
AlCl₃·6H₂O																					1435	1341
Fe₂O₃	1																					
FeCl₃																						
HCl		10	2435		24	2467	39	174	3	254		15		1			91	89	74	2541	88	71
L.O.I.																						
H₂O			3120	2089	31	9868	73	565	157	472	281	60	4189	64	9259	9259	9615	356	296	10164	2374	1812
Other																						
SiO₂	1				1																	
Organic								5							2	2						
H₂SO₄																						
Cl₂																					76	71
Total	1002	10	5555	2089	56	12335	112	838	160	726	281	75	4189	65	9261	9261	9706	445	370	12705	3973	3295

All units are in short tons per day.

Alumina from Clay/Hydrochloric Acid Process
Evaporative Crystallization
Material Balance

Component	45	46	47	48	49	50	51	52	53	54	55	56	57	58	59	60	61
Al_2O_3							10			10			10				
$AlCl_3$		94													4		
$AlCl_3 \cdot 6H_2O$																418	
Fe_2O_3		5															
$FeCl_3$							59	7							26		44
HCl	12	5			82	12				52		1	52				
$L.O.I.$																	
H_2O	435	127		20	128					1249	1185	1249	64	13		701	66
Other																	
SiO_2						36											
Organic			7				3781	3780	1	1			1			14	1
H_2SO_4																1	
Cl_2																11	
Total	447	231	7	20	210	48	3850	3787	1	1313	1185	1250	128	13	30	1145	111

HYDROCHLORIC ACID/CLAY PROCESS — HCl GAS INDUCED CRYSTALLIZATION

Process flow diagram with the following unit operations:

CLAY → CRUSHING → CALCINATION (STACK, FUEL) → GRINDING → LEACHING (ACID, WATER) → THICKENING → WASHING (WATER, WASTE) → FILTRATION (WATER, WASTE, H_2O, Cl_2)

THICKENING → SOLVENT EXTRACTION → FeCl3 STRIPPING → FeCl3 EVAPORATION → FeCl3 DECOMPOSITION (COMBUSTION WATER, FUEL, H_2SO_4, WASTE) → CONDENSER

EVAPORATION → CRYSTALLIZATION → CENTRIFUGING (ALUMINA)

INDIRECT FIRED DECOMPOSITION, DIRECT FIRED DECOMPOSITION (FUEL, COMBUSTION WATER, LOSSES, ALUMINA)

LEACH ACID PREPARATION, WASH ACID PREPARATION, DILUTE ACID RECOVERY (LOSSES), PARTIAL CONDENSING (H_2O)

HCl STRIPPING, WASTE CHLORIDE EVAPORATION, WASTE CHLORIDE DECOMPOSITION (WASTE)

Stream numbers: 1a, 1b, 1c, 2, 3, 4, 5, 6, 8a, 8b, 8c, 8d, 8e, 10, 11, 12, 13, 14a, 14b, 15, 16, 20, 23, 24, 25, 27, 28, 29a, 29b, 30, 31, 32, 33, 34, 35, 40, 44, 47, 48, 49, 50, 51, 55, 56, 57, 58, 69, 70, 71, 72, 87a, 87b, 87c, 89, 89a, 90, 91, 92, 95, 96, 97, 98, 104, 111

Alumina from Clay/Hydrochloric Acid Process
Hydrochloric Gas Precipitation
Material Balance

All units are in short tons per day.

1,000 TON/D Al₂O₃

Component	1a	1b	1c	2	3	4	5	6	8a	8b	8c	8d	9	10	11	12	13	14a	14b	15	16	20
Al₂O₃	1091	1091	1080	1080				11	54			54	54					54				
AlCl₃																	4820					
AlCl₃·6H₂O	25	25	25	25		444			3130.5	2285.5	3101.5	845	816	3100.5	3090.5	3111	2288.5	29	1	1812	37	444
Fe₂O₃									1		1	1	1					1				
FeCl₃						2348			49	36	49	13	13	116	116	107	2857			989		2006
HCl			4560		4560	6680	6	679	7431.5	5423.5	8960.5	2008	3527	8950.5	8950.5	6990	10998	3041	6	6163.5	2	5100
L.O.I.	405	405						405										1365				
H₂O	679	679				292			67		67	67	67					67				
HCl																						
Other								1	311	227	308	84		308	308	308	630.5	3		322.5		
Other as Soluble Salts										8	8	8	81						8		2	
SiO₂	1387	1387	1373	1373				14	1373		1365							1365				292
Organic	81	81	80	80					105	77	105											
H₂SO₄																						
Cl₂																					40	
Total	3668	3668	2558	2558	4560	9964	6	1110	12522	8057	12522	4465	4465	12524	12465	10516	21564	4560	15	9287	40	7842

Alumina from Clay/Hydrochloric Acid Process
Hydrochloric Gas Precipitation
Material Balance

Component	23	24	25	26	27	28	29a	29b	30	31	32	33	34	35	40	44	47	48	49	50	51	54
Al₂O₃		10															2.5			99.5	97	2.5
AlCl₃																						
AlCl₃·6H₂O	4783		960	1000																		
Fe₂O₃																						
FeCl₃			237																			
HCl	88				106			2140	834	1315					352	153	224.5		8	8	811	
L.O.I.				1																		
H₂O	163			1	132	301	52	1698	655	1033	512	1249	1761	30	655	420	125	117	201	1033.5		21.5
Other																						
Other as Soluble Salts																	15					
SiO₂																						
Organic	1		1															2		8		15
H₂SO₄																				582	582	
Cl₂																						
Total	6035	10	1198	1002	238	301	52	3837	1489	2348	512	1249	1761	30	1007	573	367	119	209	1723	1475	39

Alumina from Clay/Hydrochloric Acid Process
Hydrochloric Gas Precipitation
Material Balance

Component	55	56	57	58	69	70	71	72	87a	87b	87c	89a	89	90	91	92	95	96	97	98	102	111
Al₂O₃	1		2																			
AlCl₃		8			10		10	59	1	1	1	1	8	55	1	1	1	1	1		1	
AlCl₃·6H₂O														10		10	10	41		4	12	
Fe₂O₃				4			40									49	49	64				
FeCl₃																	61		13		17	
HCl		24		4	1979	1939		305	1021	61	960				960	1021				24		
L.O.I.	14																					
H₂O																						
Other																						
Other as Soluble Salts			2																			
SO₃																						
Organic								1				1										
H₂SO₄																						
Cl₂													3081	3082								11
Total	16	30	2	4	1989	1939	50	305	1022	61	961	1	3087	3147	961	1082	121	106	13	28	30	11

SULFUROUS ACID/CLAY PROCESS

Process flow diagram blocks and stream numbers:

- CLAY [1] → CRUSHING [2] → CALCINATION [3 STACK, FUEL] [4] → GRINDING [5] → LEACHING [6] → BLOWDOWN [7] → [8] → FILTRATION [9 WATER] [10 WASTE] [11]
- 1ST AUTOCLAVING [12] [13 STEAM] → Al₂(SO₃)₃ THICKENING [14] [15]
- 2ND AUTOCLAVING [16] [17] [18] [STEAM]
- THICKENING FILTRATION [19 WATER] [20] [21]
- DIGESTION [22 NaOH] [23] → RED MUD FILTRATION [24 WATER] [25 WASTE] [26] → PRECIPITATION [27] → CLASSIFICATION [28] [29]
- FILTRATION [30 WATER] [31] [32] → CALCINATION [33 FUEL, LOSSES] [34 ALUMINA]
- THICKENING [35] [36] → EVAPORATION [37] [38 H₂O]
- SO₂ ABSORPTION [44] → SULFUR BURNING [39 SULFUR AIR] [40] [41]
- SO₂ ABSORPTION [42 WATER] [43 LOSSES]

Alumina from Clay/Sulfurous Acid Process — Material Balance
1,000 TON/D Al₂O₃

Component	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24
Al_2O_3	1364	1364	14	1350	1350	1382		1382		270	1112	253	125	1112	32	1080	2991	1080		31	1049		2069	
Fe_2O_3	32	32		32	32	58		58		6	52			52										
NaOH															26	26		26		2	24		24	
SiO_3	1734	1734	18	1716	1716	1716		1716		1711	5			5		5		5			5	10	5	
Other	101	101	1	100	100	175		175		25	150			150						72	3		3	
L.O.I.	506	506	506																					
H_2O	843	843	843						1726						75	75		75						
S							114																	
SO_3						28599		29485		863	29348			29476	14738	14738		11747	800	11747	800	10	25458	235
SO_2							2265			1			1330				1358			121			2879	
O_3						5697		3432			3431			2101	488	1613		255		112	22		22	
Total	4580	4580	1382	3198	3198	37627	2379	35248	1726	2876	34098	253	1455	32896	15359	17537	4349	13188	800	12085	1903	20	30450	235

All units are in short tons per day.

Alumina from Clay/Sulfurous Acid Process — Material Balance

Component	25	26	27	28	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45
Al_2O_3	39	2020	2558	1504	1054		44	1010	10	1000	538	1010		1010							32
Fe_2O_3	24																				
NaOH	4	2875	2968	2843	125		6	6		6	83	2869		2869							26
SiO_3																					
Other	5	3	3		3		119	3		3											
L.O.I.																					
H_2O	115	25578	26665	25031	1634	1010	1921	723	723		1087	25865	1217	24648			2				75
S	22															129					
SO_3															129			258	10631	10631	28599
SO_2																				256	
O_3																					5697
Total	209	30476	32194	29378	2816	1010	2084	1742	733	1009	1718	29744	1217	28527	129	129	2	258	10631	10887	34429

Alumina from Anorthosite/Lime Sinter Process
Material Balance
1,000 TON/D Al₂O₃

All units are in short tons per day.

Component	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22
Al_2O_3	1318	81	1423	14	1409	1591		375	1216	2161	2161		1192	969	945	24	188	5		1000	181	
$Al_2O_3 \cdot 3H_2O$																	1930	1545	1545		385	385
Na_2O	202	5	221	2	219	2360		199	2161	2857	2857		2147	710	696	14	2196	65	6	6	2131	49
K_2O	50	12	62	1	61	61		61														
SiO_2	2687	136	2823	28	2823	2824		2800	24	851	851		0.8	850	827	23	0.8	0.3	0.3	0.3	0.5	
CaO	583	5448	6031	60	5971	5971		5971														
MgO	15	21	36	1	36	36		36														
Fe_2O_3	102	19	121		120	120		120														
CO_2		4278	4278	10328		1317	16942	747	570	619	619		570	49	49		1350	39	3	3	1311	30
Other	112	85	197	2	195	195		195				136				75						
H_2O	104	532	2142	3346		17719		11217	23444	25789	25789		23505	2420	2345		23570	890	521		22680	525
O_2				386																		
N_2				18424																		
Total	5173	10622	17362	32992	10834	32194	16942	21721	27415	32277	32277	136	27415	4998	4882	136	29233	2547	2075	1009	26689	963

ANORTHOSITE/LIME SINTER PROCESS

Process flow diagram (box labels and stream identifiers):

Inputs/feeds: ANORTHOSITE [1], LIMESTONE [2], WATER, HEAT, SODA ASH, FLOC

Unit operations (boxes):
- CRUSHING
- GRINDING
- CLASSIFICATION
- THICKENING
- FILTRATION
- PARTIAL DRYING
- PELLETIZING
- SINTERING
- SORKING
- DUSTING & SCREENING
- LEACHING
- DESILICATION
- FILTRATION
- CARBONATION
- CLASSIFICATION
- THICKENING
- FILTRATION
- CALCINATION
- WASHING & COOLING
- SCRUBBING
- MIXING
- EVAPORATION

Output labels: WASTE, WATER VAPOR, STACK, ALUMINA, WASTE HEAT

Stream numbers shown on diagram: 1–55

Alumina from Anorthosite/Lime Sinter Process
Material Balance

Component	23	24	25	26	27	28	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44
Al_2O_3	177	5					1	13		1	13				1399		1646	1399	1399			
$Al_2O_3 \cdot 3H_2O$																						
Na_2O	2082	59	2141	2141		2141	1	2	1	1	2				207		244	207	207			
K_2O			182	182		182					2				62		73	62	62			
SiO_2	0.5		0.5	0.5	0	0.5	3	25		3	25				2828		3327	2828	2828			
CaO							6	54		6	54				6031		7095	6031	6031			
MgO								1			1				36		42	36	36			
Fe_2O_3								1			1				121		142	121	121			
CO_2	1281		1317	1317		1317	938	9390	938			9390			4278		5053	4278	4278			
Other		36			0		304	3042	265		2	2649			197		232	197	197			
H_2O	22155	2028	24183	17719	0	17719		2		490	4802		451	4409	636	2469	5780	50647	15159	12054	35688	47742
O_2							35	351	35			351										
N_2							1673	16751	1673			16751										
Total*	25696	2128	27824	21360	0	21360	2960	29632	2911	500	4900	29141	451	4409	15795	2469	23614	66006	30318	12054	35688	47742

*Rounded to nearest whole number.

Alumina from Anorthosite/Lime Sinter Process
Material Balance

Component	45	46	47	48	49	50	51	52	53	54	55
Al_2O_3	1399		1399	1483	1483	74		10		247	
$Al_2O_3 \cdot 3H_2O$											
Na_2O	207		207	231	231	12				37	
K_2O	62		62	64	64	3				11	
SiO_2	2828		2828	2972	2972	149				499	
CaO	6031		6031	6285	6285	314				1064	
MgO	36		36	38	38	2				6	
Fe_2O_3	121		121	126	126	6				21	
CO_2	4278		4278							755	188
Other	197		197	205	205	10				35	
H_2O	3105	1038	2067				1659	1056	6464	2675	190
O_2											35
N_2											1673
Total	18264	1038	17226	11404	11404	570	1659	1066	6464	5350	2086

Alumina from Alunite via Reduction Roasting — Modified Bayer Processing — Material Balance

1,000 TON/D Al_2O_3

All units are in short tons per day.

Component / Process Stream	1	2	3	4	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24
Al_2O_3	788	788	785	798	1130	1161	1161	2629	2498	2479	1011	1000	131	19	600	1468	1468	1468				285	
SO_3	4483	4483	4465	4438	4394	4394	4394	4394					4372	30									
Fe_2O_3	280	280	278	276	273	273	273	273	22				273	22									
$Al_2(SO_4)_3$	1320	1320	1314	1246	104																		
K_2SO_4	671	671	668	664	657	2543																	
Na_2SO_4	30	30	30	30	30	980	7	7	7														
KOH						2	1	1		1			1										
NaOH								2773	2773	2683	10	10				2673	2673	2773	100				
H_2O	417	417	416	17	12	13334	1031	17642	21145	21386	685		7106			21301	16511	16611	100		10609	285	
SO_2														44	600								
SO_3																							
H_2SO_4																							
O_2														7									
H_2													1										
CO													60										
CO_2														23								23	2216
N_2																							668
CaO																							
Coal				85	88	88	88	88					88										
-Fuel Oil																							
Total	7989	7989	7956	7554	6688	22775	6955	27807	26385	26548	1706	1010	12031	145	600	25442	20052	20852	200		10609	308	2884

ALUNITE PROCESS (flow diagram)

ORE [1] → CRUSHING [2] → FINE CRUSH & DRY GRIND (STACK [57]) [3] → 1ST ROAST (STACK [56]; COAL [55], AIR [54]) [4] → REDUCING ROAST / REOXIDATION (STEAM [39], COAL [36], AIR [37]) [6] → DISSOLVE K_2SO_4 (KOH [58]) [8] → WASH & REMOVE SOLIDS (WATER [28])

WASH & REMOVE SOLIDS [9] → DISSOLVE ALUMINA (WATER [22], LIME [23]) [10] → RESIDUE SEPARATION (WASTE [14]) [11] → PRECIPITATE SiO_2 (SEED [15], WASTE) [17] → PRECIPITATE $Al(OH)_3$ (WATER [16], SEED) → CALCINATION (AIR [24], FUEL [25]; STACK [26]) [13] → ALUMINA

EVAPORATION (H₂O [27]) [18]; MAKE-UP NaOH [20]; [19]; [29]

EVAPORATION [35] → CRYSTALLIZATION CENTRIFUGE [32] → DRY & SIZE K_2SO_4 (PRODUCT [33]; LOSSES [38]) [31]; WASTE [30], [34]

COOL & WASH SO_2 GAS (WATER [52], WASTE [53]) [40], [41] → SO_2 PRECIPITATION [51] (WASTE) [42] → SO_2 DRYING & ACID DILUTE [43] → SO_2 CATALYTIC CONVERSION (AIR [45]) [44] → SO_2 ABSORPTION (SCRUBBER [46]) [47], [48], [49]; [50] → 93% H_2SO_4

Alumina from Alunite via Reduction Roasting — Modified Bayer Processing Material Balance

Component	25	26	27	28	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45	46	47	48
Al_2O_3	200											23				8								
SiO_2																44								
Fe_2O_3																3								
$Al_2(SO_4)_3$										67						14								
K_2SO_4					2536	2536	1728									7								
Na_2SO_4					980	980	950																	
KOH																								
NaOH					1	1																		
H_2O		877	4790	1545	13848	12716	12112	124	25	470	1132			99	14	202	282	259			180		96	355
SO_2		5						741	731	30										4		4		
SO_3		11								1														
H_2SO_4																							4721	4721
O_2		34																						
H_2																								
CO													6309			6267	6267	6267	6267	6887		6887		
CO_2		637												10		667	662	657	657	816	603			
N_2		2216																						
CaO																								
Coal																								
Fuel Oil																								
Total	200	3780	4790	1545	17365	16233	14800	865	756	568	1132	23	6309	109	14	7212	7211	7183	6924	7707	783	6891	4817	5076

Alumina from Alunite via Reduction Roasting — Modified Bayer Processing Material Balance

Component	49	50	51	52	53	54	55	56	57	58	59
Al_2O_3					8			5	3		
SiO_2					44			27	18		
Fe_2O_3					3			2	2		
$Al_2(SO_4)_3$					14			9	6		
K_2SO_4					7			4	3		
Na_2SO_4											
KOH										104	
NaOH											
H_2O	280	75	23	1896	1816			399	1	1183	129
SO_2			5		5			41			
SO_3											
H_2SO_4	3721	1000									
O_2						1674					
H_2											
CO								7847			
CO_2											
N_2						5603					
CaO											
Coal							655				
Fuel Oil											
Total	4001	1075	28	1896	1897	7277	655	8334	33	1287	129

PROCESS ASSUMPTIONS USED TO ESTIMATE THE HEAT AND MATERIAL BALANCES

Clay/Nitric Acid Process

(1) The heat and material balance is based on 1,000 ton/d of alumina.

(2) The chemical analysis of the clay feed (dry basis) is as follows:

Al_2O_3	36.5%
Fe_2O_3	0.86
SiO_2	46.4
L. O. I.	13.54
Other*	2.7
Total	100.00%

*Other is primarily TiO_2

(3) The raw clay feed to the process contains 18.5% free moisture.

(4) A rotary kiln is used for the dehydration of the clay feedstock.

(5) There is a 1% dust loss (calcined basis) from the dehydration rotary kilns.

(6) 50% by weight nitric acid is utilized in the process (makeup acid and that recycled from the acid recovery section).

(7) 95% extraction efficiency of Al_2O_3 is achieved in the leaching step.

(8) 8.4% of "other" is solubilized in the leaching step.

(9) 67% of Fe_2O_3 is solubilized in the leaching step.

(10) The underflow from the settling and washing units contains 30% solids by weight.

(11) 1% of the soluble alumina is lost in the waste residues.

(12) The iron content in the pregnant solution from solvent extraction is reduced to about 0.001 g/l.

(13) Approximately 1.0% of the dissolved alumina is lost in the solvent extraction step.

(14) 1 ton/d of organic is lost in the solvent extraction step.

(15) 17% HCl by weight is used as the stripping acid in the solvent extraction section.

(16) Triple effect evaporators are used for the concentration of the main aluminum nitrate solution and the bleed stream of aluminum nitrate.

(17) The $Al(NO_3)_3 \cdot 9H_2O$ crystals from the centrifugation of the slurry of crystals will contain 5% liquor by weight.

(18) Fluid bed roasters are used to decompose $Al(NO_3)_3 \cdot 9H_2O$ for both the main and bleed streams.

(19) There is a 0.5% dust loss (calcined basis) from the final calcination of alumina.

(20) 98% decomposition of $Al(NO_3)_3 \cdot 9H_2O$ is achieved in the indirect thermal decomposition step.

(21) There is a 3% loss of nitrate as N_2 in the indirect thermal decomposition step.

(22) There is a 2% loss of nitrate as N_2 in the direct-fired product calciner.

Clay/HCl Process (Evaporative Crystallizations)

(1) The heat and material balance is based on 1,000 ton/d of alumina.

(2) The chemical analysis of the clay feed (dry basis) is as follows:

Al_2O_3	36.5%
Fe_2O_3	0.86
SiO_2	46.4
L.O.I.	13.54
Other*	2.7
Total	100.00%

*Other is primarily TiO_2.

(3) The raw clay feed to the process contains 18.5% free moisture.

(4) A rotary kiln is used for the calcination of the clay feedstock.

(5) There is a 1% dust loss (calcined basis) from the rotary kilns used to calcine the raw clay.

(6) 20% by weight hydrochloric acid is utilized as leach acid.

(7) 15% of the heat of reaction in the leaching tanks is

released in the form of vapors. These vapors contain 1% of HCl and are condensed and returned to the leach tanks.

(8) 95% extraction efficiency of Al_2O_3 is achieved in the leaching step.

(9) 95% of Fe_2O_3 is solubilized in the leaching step.

(10) 7.2% of "other" (approximately the same as in the HNO_3 process) is solubilized in the leaching step. This soluble fraction mainly consists of sodium, potassium, calcium, and magnesium salts plus phosphate and sulfate.

(11) Approximately 10% excess HCl is contained in the leach acid.

(12) The underflow from the settling and washing units contains 33% solids by weight.

(13) Filter press solids from pregnant liquor polish filtration contains 60% solids.

(14) 1% of the soluble alumina is lost in the leach waste residues.

(15) Chlorine is added to the pregnant liquor before solvent extraction based upon one third of soluble iron content being present as ferrous iron.

(16) 100% of the chlorine added is converted to HCl.

(17) A solution of 10% Alamine 336 in kerosene is used as the organic extractant for ferric iron.

(18) The iron content in the pregnant solution from solvent extraction is reduced to about 0.001 grams per liter.

(19) A volume ratio of aqueous/organic of 3/1 is present during extraction.

(20) A volume ratio of organic/aqueous of 3.5/1 is present during stripping.

(21) The strip acid will contain approximately .03% HCl.

(22) The recycled organic solution to extraction will contain about 0.5 grams per liter Fe.

(23) The organic loss is taken as 2.1 lb per ton Al_2O_3 or 1 ton/d.

(24) Use of the distribution coefficient of .004 for aluminum when using Alamine 336 is equivalent to a loss of 10 ton/d of $AlCl_3$ in the solvent extraction section.

(25) Multiple effect evaporators (3-effect) and evaporative crystallizers are used for the concentration and crystallization of $AlCl_3.6H_2O$ intermediate product.

(26) The slurry from the $AlCl_3$ crystallizers contains 30% solids by weight and a residual 13.6% $AlCl_3$ in solution.

(27) The $AlCl_3.6H_2O$ crystal slurry from the crystallizer is filtered and washed on vacuum filters yielding a cake containing 85% solids.

(28) Filtrate mother liquor is recycled to the crystallizer.

(29) The bleed stream for minor impurities purging ("other") is the washings filtrate from vacuum filtration.

(30) One displacement wash of 35% HCl at an 83% washing efficiency is used on the product $AlCl_3.6H_2O$ crystals.

(31) The washed $AlCl_3.6H_2O$ crystals are decomposed in direct fired rotary kilns.

(32) There is a 1% dust loss (calcined basis) from the decomposition and calcination of $AlCl_3.6H_2O$ crystals.

(33) The product alumina will contain 0.01% residual Cl_2.

(34) Only combustion water has been included in the material balance of direct-fired heating units.

(35) 1% of the HCl and H_2O from the product decomposition is lost in the subsequent acid recovery operation.

(36) The washings filtrate bleed is stripped to remove 93% of free HCl and evaporated in multiple effect evaporators to a 45% solution of chlorides.

(37) Sufficient liquor from the evaporator is recycled to maintain a 40% solution of chlorides in the HCl stripping step.

(38) A stoichiometric quantity of H_2SO_4 for "other" is added before the waste chlorides decomposition.

(39) The stripping solution is concentrated to a 50% solution of chlorides in a multiple effect evaporator before decomposition in a direct-fired fluid bed decomposer.

(40) Makeup HCl is added as a 35% acid wash to the $AlCl_3.6H_2O$ vacuum filter.

(41) A $53^{\circ}F$ temperature rise occurs in cooling water.

(42) Theoretical heat requirements divided by .85 to estimate actual heat requirements. This method used only when no actual data was available.

Clay/HCl Process (HCl Gas Induced Crystallizations)

(1) The heat and material balance is based on 1,000 ton/d of alumina.

(2) The chemical analysis of the clay feed (dry basis) is as follows:

Al_2O_3	36.5%
Fe_2O_3	0.86
SiO_2	46.4
L.O.I.	13.54
Other*	2.7
Total	100.00%

*Other is primarily TiO_2.

(3) The raw clay feed to the process contains 18.5% free moisture.

(4) A rotary kiln is used for the calcination of the clay feedstock.

(5) There is a 1% dust loss (calcined basis) from the rotary kilns used to calcine the raw clay.

(6) 15% of the heat of reaction in the leaching tanks is released in the form of vapors. These vapors contain 1% HCl and are condensed and returned to the leach tanks.

(7) 95% extraction efficiency of Al_2O_3 is achieved in the leaching step.

(8) 95% of Fe_2O_3 is solubilized in the leaching step.

(9) 16.3% of "other" is solubilized in the leaching step. This soluble fraction mainly consists of sodium, potassium, calcium, and magnesium salts, plus phosphates and sulfates.

(10) Approximately 5% excess HCl is contained in the leach acid.

(11) The underflow from the settling and washing units contains 33% solids by weight.

(12) Filter press solids from pregnant liquor polish filtration contains 60% solids.

(13) 1% of the soluble alumina is lost in the leach waste residues.

(14) Chlorine is added to the pregnant liquor before solvent extraction based upon one third of soluble iron content being present as ferrous iron.

(15) 100% of the chlorine added is converted to HCl.

(16) A solution of 10% Alamine 336 in kerosene is used as the organic extractant for ferric iron.

(17) The iron content in the pregnant solution from solvent extraction is reduced to about 0.001 grams per liter.

(18) A volume ratio of aqueous/organic of 3/1 is present during extraction.

(19) A volume ratio of organic/aqueous of 3.5/1 is present during stripping.

(20) The strip acid will contain approximately 0.03% HCl.

(21) The recycled organic solution to extraction will contain about 0.5 grams per liter Fe.

(22) The organic loss is taken as 2.1 lb per ton Al_2O_3 or 1 ton per day.

(23) Use of the distribution coefficient of .004 for aluminum when using Alamine 336 is equivalent to a loss of 10 tons per day of $AlCl_3$ in the solvent extraction section.

(24) A multiple effect evaporator (3-effect) is used to concentrate the crystallizer feed solution.

(25) The $AlCl_3 \cdot 6H_2O$ crystal slurry from the crystallizers is filtered and washed on centrifuges yielding a cake containing 95% solids.

(26) Mother liquor from centrifugation is recycled to the crystallizer.

(27) The bleed system for minor impurities purging ("other") is 5% of recycle acid stream.

(28) 0.2 lb of 35% acid wash is used per 1 lb of $AlCl_3 \cdot 6H_2O$ crystals.

(29) The washed $AlCl_3 \cdot 6H_2O$ crystals are decomposed in indirect fired kilns.

(30) There is a 1% dust loss (calcined basis) from the decomposition and calcination of $AlCl_3 \cdot 6H_2O$ crystals.

(31) The product alumina will contain 0.01% residual Cl_2.

(32) Only combustion water has been included in the material balance of direct-fired heating units.

(33) 12 ton/d of HCl is lost in the acid recovery section.

(34) The bleed liquor is stripped to remove 93% of free HCl and evaporated in multiple effect evaporators to a 45% solution of chlorides.

(35) Sufficient liquor from the evaporator is recycled to maintain a 40% solution of chlorides in the HCl stripping step.

(36) A stoichiometric quantity of H_2SO_4 for "other" is added before the waste chlorides decomposition.

(37) The stripping solution is concentrated to a 50% solution of chlorides in a multiple effect evaporator before decomposition in a direct-fired fluid bed decomposer.

(38) Makeup HCl is added as HCl gas to the acid recovery section.

(39) A 53% temperature rise occurs in cooling water.

Clay/H_2SO_3 Process

(1) The heat and material balance is based on 1,000 ton/d of alumina.

(2) The chemical analysis of the clay feed (dry basis) is as follows:

Al_2O_3	36.5%
Fe_2O_3	1.86
SiO_2	46.4
L.O.I.	13.54
Other*	2.7
Total	100.00%

*Other is primarily TiO_2.

(3) The raw clay feed to the process contains 18.5% free moisture.

(4) A rotary kiln is used for the dehydration of the clay feedstock.

(5) There is a 1% dust loss (calcined basis) from the dehydration rotary kilns.

(6) 80% extraction efficiency of Al_2O_3 is achieved in the SO_2 leaching step.

(7) 75% of other metal impurities is solubilized in the leaching step.

(8) 81.25% of Fe_2O_3 is solubilized in the leaching step.

(9) 0.3% of SiO_2 is solubilized in the leaching step.

(10) The waste leach mud after filtration will contain 30% moisture.

(11) One-half of the liquor into the thickener following the first autoclaving step (precipitation of monobasic aluminum sulfite) will exit via overflow and one-half will exit with underflow solids (equivalent to approximately 20% solids in the underflow).

(12) The decomposition of monobasic aluminum sulfite to alumina will be accomplished at a 97% efficiency (7 atm autoclaving).

(13) The product stream from the monobasic aluminum sulfite decomposition is used to preheat the feed to this process step.

(14) Approximately 9% of contained SO_2 is lost as sulfur and sulfate during the monobasic aluminum sulfite decomposition.

(15) Crude alumina product from filtration will contain approximately 42% free and combined water.

(16) In the acid recovery section, sulfur is burned with air to produce a stoichiometric amount of SO_2 for makeup.

(17) There is a 2-ton/d SO_2 loss from the 3 atm absorber in the acid recovery system.

(18) Waste heat is recovered from the sulfur burner.

(19) There is a 100% absorption efficiency of acid vapors in the 9 atm absorber in the acid recovery system.

(20) The purification of crude alumina product is accomplished by leaching with caustic in a modified Bayer plant.

(21) Sulfur and Fe_2O_3 compounds precipitate during caustic digestion and are removed with the red mud waste solids.

(22) The filtered and washed red mud will contain 55% moisture.

(23) No conversion of NaOH to Na_2CO_3 is considered.

(24) Approximately half of the alumina precipitates while the balance is recycled.

(25) Caustic concentration during digestion is approximately 9.5% by weight as NaOH.

(26) A 5-effect evaporator is used to concentrate the recycle caustic liquor stream.

(27) The trihydrate alumina feed to calcination will contain approximately 11% free moisture.

(28) A rotary kiln is used to calcine the alumina product.

(29) There is a 1% dust loss (calcined basis) from the rotary kiln.

(30) The over-all recovery of alumina in the caustic purification section is approximately 95%.

Anorthosite Lime-Sinter Process

(1) The heat and material balance is based on 1,000 ton/d of alumina.

(2) The chemical analysis of the anorthosite feed (dry basis) is as follows:

Al_2O_3	26.0%
Na_2O	4.0
K_2O	1.0
SiO_2	53.0
CaO	11.5
MgO	0.3
Fe_2O_3	2.0
CO_2	–
Other	2.2
Total	100.00%

(3) The raw anorthosite feed to the process contains 2% free moisture.

(4) The chemical analysis of the limestone feed (dry basis) is as follows:

Al_2O_3	0.8%
Na_2O	0.04
K_2O	0.12
SiO_2	1.4
CaO	54.0
MgO	0.2
Fe_2O_3	0.2
CO_2	42.4
Other	0.84
Total	100.00%

(5) The raw limestone feed to the process contains 5% free moisture.

(6) The weight ratio of limestone/anorthosite (dry basis) in the feed is 2/1.

(7) 15% of ground slurry will be +200 mesh and will be recycled as 50% solids from wet classification to wet grinding.

(8) The -200 mesh solids underflow from the thickener will contain 50% solids.

(9) The -200 mesh solids filter cake will contain 17% moisture.

(10) The 17% moisture filter cake of limestone and anorthosite will be reduced to 12% in a waste heat dryer. The heat will be supplied by the hot sinter kiln gases.

(11) The sinter kiln will be a rotary kiln.

(12) There is a 1% dust loss (calcined basis) from the sinter kiln. No Na_2O is vaporized and lost from the kiln.

(13) Coal with a heating value of 13,000 Btu/lb is used as the sinter kiln fuel. It is burned using 1% excess oxygen.

(14) 5% of the sinter out of the kiln will not "dust" and will be recycled as +20 mesh material to the sinter kiln.

(15) 75% extraction efficiency of Al_2O_3 is achieved in the leaching step.

(16) 24% extraction of Na_2O and 0% K_2O from the limestone/anorthosite mix is achieved.

(17) No CaO, Fe_2O_3, other, or MgO are extracted in the leaching step.

(18) Vacuum filtration is used to separate the pregnant liquor from the leach residue and 1.5 displacements of wash water are used to yield a 95% washing efficiency.

(19) Sufficient SiO_2 to yield 1 gram per liter SiO_2 in the pregnant liquor is extracted in the leaching step.

(20) The leach residue will contain 48% solids.

(21) 2% soluble Al_2O_3 is lost in the leach residue.

(22) DSP seeded desilication will be used.

(23) DSP will be $3Na_2O.3Al_2O_3.5SiO_2.5H_2O$.

(24) SiO_2 from desilication will be 0.035 gram per liter.

(25) DSP filter cake will contain 50% solids.

(26) Recycle DSP filter cake to desilication will not be washed.

(27) 100 gram per liter DSP seed will be recycled to desilication.

(28) DSP bleed filter cake to the sinter kiln will be completely washed.

(29) 80% CO_2 absorption efficiency will be achieved in carbonation.

(30) Scrubbed carbonation gases will be saturated with water at 130°F.

(31) Alumina hydrate slurry will settle to 60% solids.

(32) Alumina hydrate filter cake will contain 75% solids.

(33) There is a 1% dust loss (calcined basis) from the alumina hydrate calcination kiln. A rotary kiln is used for this calcination.

(34) Sinter kiln gas scrubbers will remove 100% of contained solids.

(35) The scrubber water bleed will be at 2% solids.

(36) A multiple-effect evaporator (3-effect) is used to concentrate the spent liquor solution.

Alunite Process

(1) The heat and material balance is based on 1,000 ton/d of alumina.

(2) Ore recovery (Al_2O_3 basis) through K_2SO_4 leach is 97%.

(3) Al_2O_3 is 90% soluble in NaOH leach.

(4) Al_2O_3 recovery in caustic refining section is 97%.

(5) Alunite ore contains 5.2% H_2O of which 5% is removed in drying.

(6) During ore drying 4.5% of $Al_2(SO_4)_3$ in the alunite decomposes.

(7) The reduction of the $Al_2(SO_4)_3$ is performed by coal gas.

(8) The coal gas used in 7 above is burned in excess air to provide the heat and O_2 for reoxidation of the sulfides formed in 7 above.

(9) The redox operation converts 92% of the residual $Al_2(SO_4)_3$ to Al_2O_3.

(10) Coal ash (13%) from ore drying and redox remains in the ore and is inert.

(11) A 2% KOH excess is used to convert residual $Al_2(SO_4)_3$ to Al_2O_3.

(12) SiO_2 contained in the ore dissolves to the extent of 0.5% in the caustic digest.

(13) All soda in the caustic refining section is assumed to be caustic soda.

(14) Dissolved SiO_2 is converted to $2 NaOH.Al_2O_3.SiO_2$ (DSP) for removal.

(15) DSP filter cake contains 33% H_2O.

(16) Residual ore (primarily SiO_2) is rejected at 40% solids.

(17) Solids feed to caustic digest contain 15% free H_2O.

(18) Alumina/caustic (A/C, caustic as Na_2CO_3) is 0.4 in caustic digest feed liquor and 0.65 in caustic digest exit liquor.

(19) Caustic concentration (as Na_2CO_3) is 220 g/l in digestion.

(20) The chemical analysis of the alunite feed is as follows:

Al_2O_3	9.86%
SiO_2	56.11
Fe_2O_3	3.50
$Al_2(SO_4)_3$	16.53
K_2SO_4	8.40
Na_2SO_4	0.38
H_2O	5.22
Total	100.00%

PROCESS ENERGY COMPARISON
Millions BTU's/T Al_2O_3

	HNO_3	HCl Evap.	HCl Sparging	H_2SO_3	Anorthosite	Alunite
Net Heat to Process (1)	37.04	31.42	23.52	28.16	61.90	26.12
Gross Fuel (2)	46.90	32.63	27.36	37.45	61.90	29.33
% of Gross Fuel from Coal	87%	38%	66%	72%	94%	87%
from Fuel Oil	13%	62%	34%	28%	6%	13%

(1) Total heat actually delivered to process stream. Does not include heat losses from auxiliaries such as steam boilers, molten salt boilers, and steam distribution system.

(2) Total fuel burned.