Conf-911265--1

ANL/CP--73449

DE92 003408

RECYCLING OF ALUMINUM SALT CAKE

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(Prepared for presentation at and publication in the proceedings of The Seventh International Conference on Solid Waste Management and Secondary Materials)

December 1991

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ABSTRACT

The secondary aluminum industry generates more than 110 x 10^3 tons of salt-cake waste every year. This waste stream contains about 3-5% aluminum, 15-30% aluminum oxide, 30-40% sodium chloride, and 20-30% potassium chloride. As much as 50% of the content of this waste is combined salt (sodium and potassium chlorides). Salt-cake waste is currently disposed of in conventional landfills. In addition, over 50 x 10^3 tons of black dross that is not economical to reprocess in a rotary furnace for aluminum recovery ends up in landfills. The composition of the dross is similar to that of salt cake, except that it contains higher concentrations of aluminum (up to 20%) and correspondingly lower amounts of salts. Because of the high solubility of the salts in water, these residues, when put in landfills, represent a potential source of pollution to surface-water and groundwater supplies. The increasing number of environmental regulations on the generation and disposal of industrial wastes are likely to restrict the disposal of these salt-containing wastes in conventional landfills. Processes exist that employ the dissolution and recovery of the salts from the waste stream. These wet-processing methods are economical only when the aluminum concentration in that waste exceeds about 10%. Argonne National Laboratory (ANL) conducted a study in which existing technologies were reviewed and new concepts that are potentially more cost-effective than existing processes were developed and evaluated. These include freeze crystallization, solvent/antisolvent extraction, common-ion effect, high-pressure/high-temperature process, and capillary-effect systems.

This paper presents some of the technical and economic results of the aforementioned ANL study.

INTRODUCTION

Scrap-aluminum recycling (Figure 1) starts with the melting of the scrap aluminum in open-hearth reverberatory furnaces. In order to decrease the interfacial tension of the dross/aluminum and allow the aluminum to coalesce for easier separation (as well as to minimize the oxidation of the molten aluminum and to capture the aluminum oxide that forms during the melting process), a

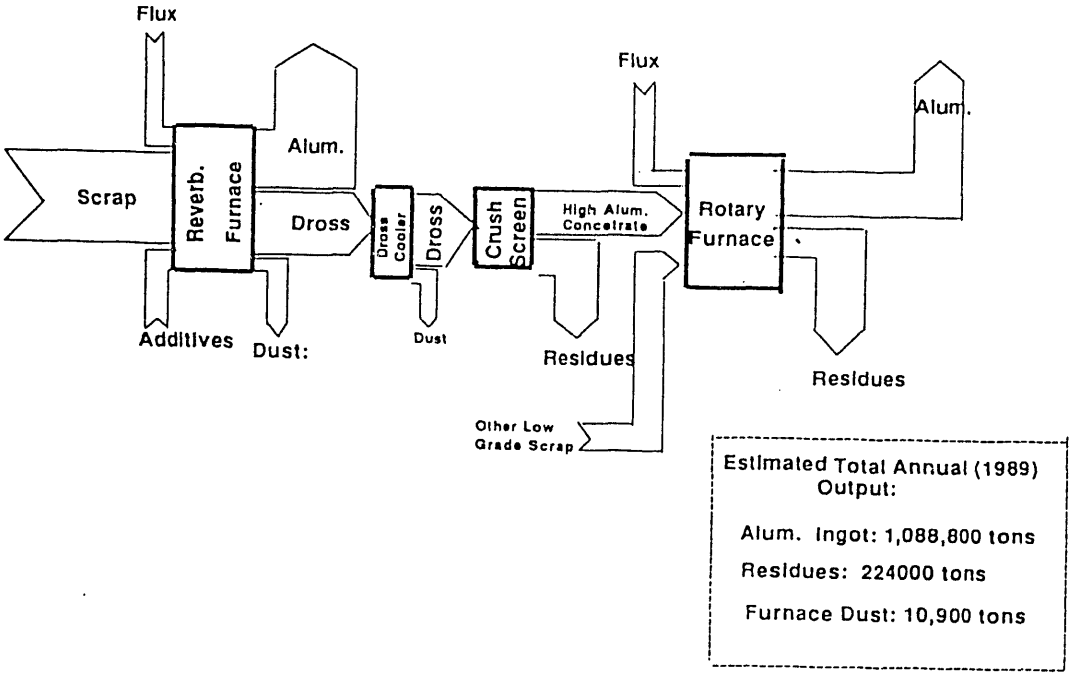

Figure 1. Scrap-Aluminum-Recycling Process

salt mixture, called flux, is spread on top of the bed. Flux typically consists of about 62% sodium chloride, 32% potassium chloride, and 6% other materials, including cryolite. This layer also serves to trap impurities that are with magnesium to form magnesium chloride, which also reports to the flux layer. After the melting process is completed, the top layer of dark-colored material, called black dross, is removed for further processing to recover some of its aluminum content. The black dross typically consists of aluminum metal (10-20%), salt-flux mixture (40-55%), and aluminum oxide (20-50%). One of two methods is used to process the black dross for the recovery of aluminum: (1) reprocessing of the black dross in a rotary furnace (along with low grade-scrap) or (2) physically separating the aluminum from the black dross using hammer mills. When the aluminum content of the dross is such that it is not economical to recover, the dross is disposed of in landfills. Rotary furnace processing requires the addition of salt fluxes in even larger quantities than that required for reverberatory furnace processing.

Residue generated by the rotary furnaces is referred to as salt cake, and it contains about 3-5% aluminum, 15-30% aluminum oxide, 30-45% sodium chloride, and 20-30% potassium chloride. The amount of black-dross and salt-cake residues generated from U.S. secondary aluminum-smelting plants in 1988 was approximately 229 x 10^3 tons.[1] The quantities of scrap and salt cake that are being processed are increasing (Table 1). Because the salts are readily leachable, the disposal of this material in landfills is becoming increasingly more costly and less acceptable. Therefore, alternatives to landfilling are in demand.

Table 1. Consumption and Recovery of Aluminum Scrap[2] (1985-1989)

Item	1985	1986	1987	1988	1989[a]
Scrap consumption (10^3 ton)					
Primary producers	801	861	1,063	1,130	1,133
Secondary producers	1,017	973	968	1,047	887
Other	362	355	399	411	376
Total	2,180	2,189	2,430	2,588	2,396
Scrap recovery (10^3 ton)					
from New scrap	1,005	1,090	1,250	1,087	1,150
from Old scrap	937	864	939	1,152	1,003
Total	1,942	1,954	2,189	2,339	2,253

[a]Estimated

This paper summarizes the results of a study conducted by Argonne National Laboratory (ANL) in which alternatives to the disposal of this material in landfills are presented. In the study, we (1) reviewed currently available recycling technologies and processes and (2) developed and evaluated new and innovative concepts.

TECHNOLOGIES FOR RECYCLING BLACK DROSS AND SALT CAKE

In the past, the objective of black-dross and salt-cake recycling was to maximize aluminum recovery and minimize cost. Because the disposal of the resulting waste was not a major cost factor, there was no incentive to minimize it or recycle some of its constituents. New environmental realities have forced the secondary aluminum industries in Europe and in the United States to consider recycling and waste minimization.

European Salt-Cake-Recycling Systems

In Europe, prototype salt-cake-recycling systems (less than 60 x 10^3 t*/yr) have been undergoing field testing and limited operation for a few years. Most of these systems have been installed in Italy and in Germany, countries in which the disposal of salt cake in conventional landfills is not permitted. Two organizations are marketing their systems in the United States: (1) Engitec Impianti, S.p.A. of Italy, which is represented in the United States by Lake

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*t = metric ton

Engineering, Inc; and (2) Berzelius Umwelt-Service A.G. (B.U.S.), which is represented in the United States by Horsehead Resource Development Co., Inc. These systems devised by both of these organizations employ essentially the same basic steps:[3,4] crushing and screening for aluminum recovery, salt leaching, off-gas treatment, and evaporation/crystallization for salt recovery.

In 1977, Engitec installed its first system in Italy, a system that is capable of recovering 21 x 10^3 t of salt per year. In 1988, this plant was renovated and retrofitted with an energy-efficient evaporator/crystallizer equipped with a mechanical vapor compressor. Engitec has also pilot tested a proprietary flash evaporator to be used with small recycling plants, as well as a preconcentration unit for large-capacity evaporator/crystallizer systems.

The estimated capital and operation costs of the Engitec system are given in Table 2 for two plants having capacities of about 20 x 10^3 and 66 x 10^3 t/yr.

The B.U.S. (German) system was installed in Germany about four years ago, and it has been in operation since that time. It was designed to process 50 x 10^3 short tons of salt cake and 13 x 10^3 short tons of ball mill dust per year. So far, this plant has processed over 240 x 10^3 short tons of salt cake and ball mill dust. B.U.S. is planning to have two more operating plants in Germany by the end of 1991, bringing the total installed capacity to over 260 x 10^3 short tons per year. The salt recovered from this process meets the specifications of the smelters. In the last four years, 130 x 10^3 short tons of salt have been recovered and reused successfully by eight secondary aluminum smelters. Horsehead Resource Development Co, B.U.S.'s U.S. market business partner, estimated[4] the capital cost for a 100-t/day system to be about $35,000,000.

U.S.-Developed Dross- and Salt-Cake-Recycling Systems

The U.S. Bureau of Mines[5] developed, in the late 1970s, a method to recover aluminum, aluminum oxide, and fluxing salts from dross and salt cake. In this method, dross or salt cake is leached with water at room temperature to produce a saturated brine slurry. The slurry is screened to yield an aluminum-rich fraction that can be returned to the rotary furnace. The remaining slurry

Table 2. Estimated Costs of the Engitec Salt-Cake-Recycling System[3]

Item	System A	System B
Capacity (t/d)	60	200
Salt-cake process (t/yr)	19,800	66,000
Capital cost ($/t processed)	250	150
Annual days of operation	330	330
Operating cost ($/t)	57	33

is vacuum filtered, yielding a clear brine solution and an aluminum-oxide cake. Evaporation of the clear process, over 80% of the metallic aluminum and essentially all of the fluxing salts are recovered. This method does not employ crushing of the salt cake or the dross. With the exception of the lack of crushing, the process steps are essentially the same as those employed in the European technologies.

The technical feasibility of this method was demonstrated in the laboratory. On the basis of the data generated from the laboratory experiments, a commercial-scale system was designed, and an economic analysis of the process was conducted. The commercial-scale system was designed to process 90 t/day of salt cake. The cost analysis conducted by the U.S. Bureau of Mines was based on 1977 capital and operating costs, and these costs were revised in this study to reflect current costs, as shown in Table 3.

The design and economic analysis of the U.S. Bureau of Mines system were developed in 1977. Since that time, environmental air-pollution emission regulations have become more stringent; therefore, the costs of pollution-control equipment in the Bureau of Mines study may be underestimated.

American Recovery Technology Systems (ARTS) has designed, built, and is currently operating its first dross and salt-cake recycling system in Cleveland, Ohio. The plant has a capacity of processing about 65 x 10^3 t/yr. The closed-loop system consists essentially of the same processes employed in the European technologies: crushing, leaching, and evaporation/crystallization. The evaporation is carried out at temperature of about 238-240°F.

Table 3. Estimated Costs of the Bureau of Mines System (current dollars)

Estimated capital cost: $7,000,000

Capacity: 90 t/day

Operation: 330 days/yr, 24 h, 7 day/week

Direct labor: three workers/shift

Item	Unit Consumption	Cost Per Unit	Total Annual Cost ($)	Unit Cost ($/t)
Personnel	0.8 h/t	$17.00/h	407,150	13.60
Electricity	33.6 kWh/t	$0.05/kWh	50,290	1.68
Process steam	2.42 10^3 lb/t	$9.50/$10^3$ lb	688,250	22.99
Oil	6.96 gal/t	$0.50/gal	104,030	3.48
Water	0.59 m^3/t	$0.26/$m^3$	7,780	0.15
Maintenance	—	—	449,055	15.00
Depreciation (11-yr life)	—	—	610,275	20.40
Insurance (1% of capital equip.)	—	—	67,180	2.24
Total			2,384,010	79.64

Alcan International Limited and Plasma Energy Corporation are currently developing a dross processing technology that does not require salt. This process uses a specially designed rotary furnace heated by a plasma-arc gas heater. Inside the plasma torch are two tubular electrodes placed end-to-end but separated by a small gap. During operation, a process gas, such as air or nitrogen, is injected into the small gap between the electrodes. This arc heats the incoming gas at temperatures in excess of 5,000°C. At this temperature, the gas is dissociated and partially ionized. The ionized gas (plasma) is ejected out of the torch and into the rotary furnace to heat the dross.

The pilot dross-plasma-treatment plant was installed in 1988. The plasma torch is rated at 1 MW, and the capacity of the rotary furnace is about 3 t/h. The 1-MW torch uses 1.5-3 m^3 of gas per minute. It is designed to process industrial-scale samples of dross, and blocks up to 1 t can be charged (as can very dusty dross). The rotary furnace is designed to be airtight, and during its operation, the inside pressure is always maintained positive to prevent outside air infiltration. To date, about 400 t of dross, representing a total of about 150 batches, have been processed. The dross was supplied from various primary- and secondary-aluminum production plants in the United States, Canada, and Europe. The aluminum content in these samples ranged from 10% to 80%. The typical electrical energy input into the system per metric ton of dross processed was estimated by Alcan to be about 844 kWh for the nitrogen plasma and 475 kWh for the air plasma.

The plasma dross-treatment process is an innovative and potentially cost-effective approach for the recovery of aluminum from the drosses of the primary aluminum industry. The key advantage of this process is that it eliminates the salt-flux requirements in dross processing. As a result, it can generate salt-free by-products, if the dross charged into the furnace did not contain any salt, as is typically the case in some of the drosses of the primary aluminum industry. If the dross processed contains salt flux, as in black dross, the salt will remain in the vetrified residual waste, which may not be leachable.

Alcan is building a 15 x 10^3-t/year plasma dross-treatment plant in Canada. This commercial-scale facility will provide reliable data that can be used to evaluate the applicability of this process, as well as its technical feasibility and cost-effectiveness for addressing the waste-minimization and disposal issues facing the secondary and primary aluminum Industries.

INNOVATIVE CONCEPTS FOR RECOVERING SALT FROM DROSS AND SALT CAKE

As discussed earlier, several commercial processes are being developed for the leaching of chlorides from aluminum salt cake using water. These methods rely on the evaporation of water to concentrate and recover the salts after they are leached. This process is energy-intensive; it requires a minimum of about 0.64 kWh/kg water processed using a single-stage evaporator. The use of multiple-effect evaporators could reduce the amount of energy required by about 50%. Licon, Inc., in Pensacola, Florida, stated in its literature that similar solutions can be concentrated by using about 50-100 Btu/lb (0.032-0.065 kWh/kg) of water removed. Licon's method uses vapor compression to recycle the latent heat and operates at reduced pressures in the range of

20-26 in. of mercury. The estimated cost of the evaporation equipment is about $500,000-$750,000 for a system that can treat about 500-1,000 gal/h. The corrosive nature of the salt cake and dross may result in a higher cost, because special construction materials may be required.

As part of our effort, we developed and evaluated the following concepts in search of a more efficient and economical method for procuring salt cake: (1) freeze crystallization, (2) solvent/antisolvent extraction, (3) common-ion effect, (4) high-pressure/high-temperature process, and (5) capillary-effect systems. These methods are discussed below.

Freeze Crystallization

This process involves cooling the solution beyond its saturation point. At this point, the salts start to crystallize and precipitate out of the solution. Cooling may be continued until pure ice crystals start to form and rise to the surface, resulting in the precipitation of more salts. The process may be operated at reduced pressures (near the triple point of water), the point at which water evaporates and ice forms simultaneously. The cooling of the solution can be achieved by means of indirect or direct heat exchange. Indirect cooling involves the cooling of the solution without its coming in contact with the refrigerant. Direct cooling involves the direct addition of a refrigerant to the solution. A recent article[6] described this method using direct contact of the water with a refrigerant. The author reported that the process is cost-effective for wastewater from any industry with recoverable by-products worth more than a cent a gallon. The applicability of this process to water/salt solutions generated from the processing of aluminum salt cake was evaluated. The conclusions are summarized below:

1. This method will work on separating the salt-cake salts and will produce pure water and may also enable the recovery of magnesium chloride from the solution in a marketable purity. Recovering the magnesium chloride is highly desirable to prevent its buildup in the recycled flux as an impurity.

2. Although considerably less energy is expended on freezing water than boiling it, the energy used in the cooling process is generally electric, while heating can be achieved with less-expensive means. Furthermore, the cost of the cooling equipment is generally higher than the cost of the heating equipment. Therefore, the freeze-drying process may be more expensive to apply than the evaporation process. It is estimated that the treatment cost will be about 10-20¢/gal. The value of the recovered salts ($100/ton) will be about $0.15/gal of water processed when the dissolution process takes place at room temperature.

Solvent/Antisolvent Extraction

Precipitation of a solute from a solvent using another solvent is a well-known technique that is widely practiced in the chemical and petrochemical industries. We experimented with this technique to separate the salts from water. In the laboratory, we tested the use of water-miscible organic solvents to precipitate salts from a saturated salt-cake solution by using several organic solvents that have high solubility in water (acetone, isopropyl alcohol, ethyl methyl ketone, methyl alcohol, tetrahydrofuran, and dioxane). The inorganic salts, commonly found in salt cake, have negligible solubility in these organic solvents. Because of their high affinity for water and their relative insensitivity to the salts, these solvents can displace the salts in solution, thereby causing the salts to precipitate. Usually a second phase of an essentially salt-free organic solvent/water mixture also separates out. After the precipitate is filtered, the liquid mixture can be separated by distillation, with both the solvent and the water recovered for reuse. The latent heat of vaporization of organic solvents is considerably lower than that of water. Furthermore, the boiling points of the organic solvents are generally lower than that of water. Therefore, lower-grade heat can be used in the distillation process.

Experiments were conducted using a saturated salt-cake solution prepared with the salts recovered from an actual sample of salt-cake slag. The salt content of the solution was found to be 31.52% by weight, and it contained about 45% potassium chloride and 55% sodium chloride. Various quantities of solvents were added to 50-mL samples of salt-saturated water solution. As a result, various amounts of salts precipitated from solution. After about an hour, the precipitates were filtered, dried, and weighed. In the samples where excess solvent was added, two liquid phases separated, because the solubility of the water-miscible solvents decreased in the salt solution (as compared with salt-free water). We found that the excess layer can be decanted and reused without additional treatment because the level of contamination is low. The results obtained with different solvents are discussed below.

Preliminary tests of mutual compatibility eliminated methyl alcohol, methyl ethyl ketone, and tetrahydrofuran from further consideration. Methyl alcohol, even though it was compatible with the salt solution, caused negligible salt precipitation. The other two solvents showed negligible solubility in the salt solution. Warranting further consideration were acetone, isopropyl alcohol, and dioxane. The results obtained with these water-miscible solvents are summarized in Table 4 and Figure 2.

Table 4. Salt Precipitation with Water-Miscible Solvents

Volume of Saturated Water Solution (mL)	Volume of Solvent Added (mL)	Mass Percent of Salts Recovered		
		Isopropyl Acetone	Isopropyl Alcohol	Isopropyl Dioxane
50	5	5.0	5.8	0
50	10	9.3	6.7	9.7
50	20	15.0	9.0	18.3
50	30	17.0	11.4	19.7
50	40	19.2	17.0	20.5
50	50	22.5	20.0	23.8
50	100	32.9	33.1	31.6

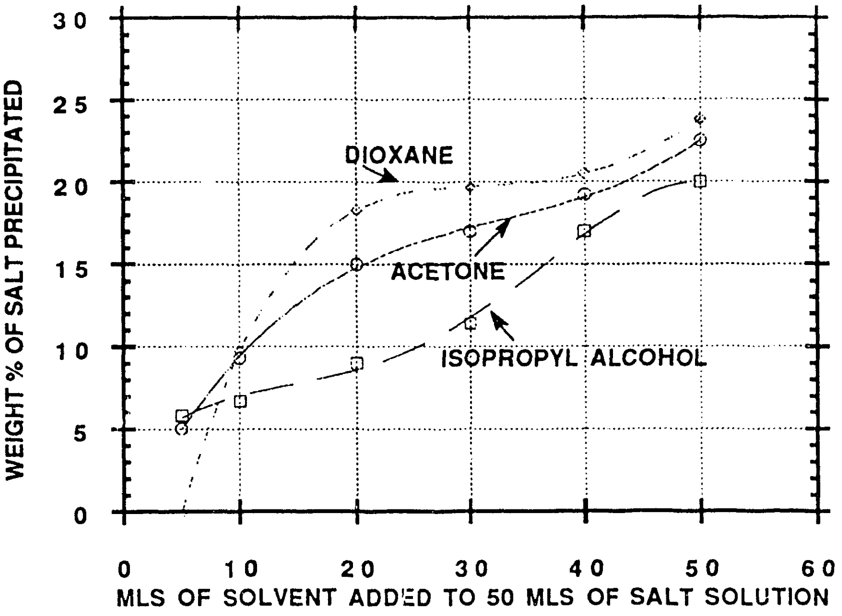

Figure 2. Salt Precipitation using Various Water-Miscible Solvents

As expected, we observed that the solubility of these organic water-miscible solvents in the salt solution was significantly decreased (as compared with salt-free water), so that a separate upper phase, rich in organic water-miscible solvents, was formed. We investigated whether this upper phase could be used after decanting for the desalting of fresh saturated solutions. In the case of acetone, 30 mL of the supernatant liquor was added to 50 mL of fresh saturated solution. This procedure resulted in the precipitation of 1.65 g of salt (8.6%), which is about half what was precipitated when fresh acetone is used. Similarly, when 30 mL of isopropyl alcohol was added to 50 mL of fresh saturated solution, it precipitated about 1.14 g of salt (5.9%). This quantity of salt represents about half of what was precipitated when fresh isopropyl alcohol was used.

Examination of the data in Table 4 and Figure 2 indicates that on the basis of the amount of salt precipitated per milliliter of solvent added, dioxane appears to be the most effective, followed by acetone. However, dioxane forms an azeotrope with water, which makes it more difficult to separate. Moreover, in both cases, the energy required to regenerate the solvent is more than the energy required to separate the same amount of salt by boiling the water. For example, to boil off 1 lb of acetone, about 22% of the energy required to boil 1 lb of water is needed, but it will precipitate about 20% of the salt that can be precipitated by boiling the water. Therefore, larger evaporation equipment will be required for the evaporation of the acetone. However, because the acetone/water solution contains very little salts, it is less corrosive.

We also evaluated a hybrid system in which water is boiled off in the first section and acetone is used in the second section. The steam generated in the first section (leaving a multiple-effect evaporator at about 1 atm) is used to regenerate the solvent in the second section. This system takes advantage of the lower boiling temperature of the acetone. This system is illustrated schematically in Figure 3. The solvent section can be sized to match the capacity of the effluent steam from the multiple-effect evaporator, and thus it could be operated energy-free. Typically, the solvent section should be able to handle about 25% of the total saturated water solution, which corresponds to a 25% reduction in the energy requirement of the process employed to separate the water from the salts.

Common-Ion Effect

We conducted a preliminary study on the concept of salting out the salt-cake salts from a saturated solution by the addition of compounds that contain chloride ions. Two sources of chloride ions were evaluated: hydrochloric acid and lithium chloride. The results are discussed below.

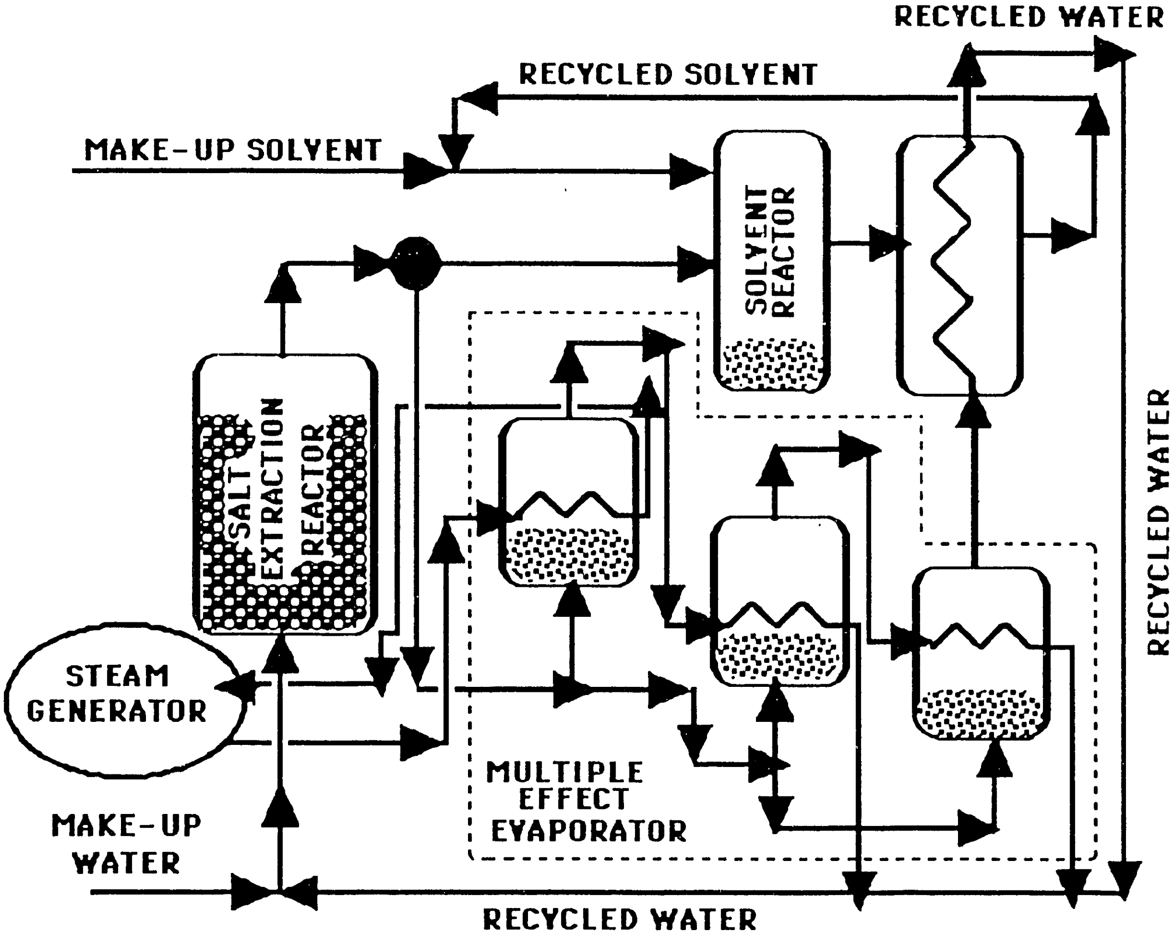

Figure 3. Schematic Diagram of the Hybrid (Evaporation/Solvent Precipitation) System

Hydrochloric Acid

When 50 mL of commercial hydrochloric acid containing about 38% by weight hydrogen chloride was added to 50 mL of saturated salt solution, 55.2% (10.6 g) of its salt content was precipitated. This level of salt precipitation was significantly higher than that achieved by the water-miscible solvent-precipitation process. Clearly, it would be more efficient to bubble 100% hydrogen chloride gas directly into the salt solution.

The residual acidified solution can be filtered to remove the precipitated salts and reconcentrated by removing some of its water content by boiling or freeze crystallization. A concentration of about 38% can be attained by the use of such methods. Further concentration would require more elaborate techniques, because the acid/water azeotrope would have to be broken. Therefore, we do not expect this technique to offer any energy or cost savings over the simple removal of water by boiling or freezing.

Lithium Chloride

Lithium chloride is worth investigating for two reasons: (1) its high solubility in water and (2) the strong and extensive hydration sphere for the small Li^+ ion, which might help "squeeze out" the salts from the total aqueous environment. The significant disadvantages of this approach include the following: (1) the high cost of lithium chloride, (2) possible coprecipitation of lithium chloride with the potassium and sodium salts and the resulting contamination of the recovered product, and (3) recovery and reuse problems associated with water and lithium chloride.

Fifty milliliters of saturated lithium chloride solution (40% by weight) was added to 50 mL of saturated salt solution. This procedure resulted in the precipitation of 11.2 g of salt (58.2%). Although this is a slightly better performance than that achieved with hydrochloric acid, we do not believe it is enough to overcome the disadvantages described above. Therefore, this approach was not investigated further.

High-Temperature and High-Pressure Process

We conceived a closed-system process that involves leaching the salts at elevated temperatures using pressurized liquid water. This process exploits the dependence of salt solubility on temperature (Figure 4). Increasing the temperature from about 20°C to 300°C increases the solubility of the sodium and potassium chlorides in liquid water by about 40% and 110%, respectively. Therefore, more salts can be leached at higher temperatures by each kilogram of water. For a 50/50 mixture, as what is encountered in aluminum salt cake, an overall increase of over 75% is expected. This value is consistent with our measured values for the mixture in the range of 20-100°C.

Figure 5 is a schematic diagram of the process. The process involves leaching of the salts at elevated temperatures using compressed liquid water and cooling the pregnant solution to effect the crystallization of the salts from the water. The leaching process can be carried out at subcritical or supercritical conditions. The approximate material and energy data shown in Figure 5 are for

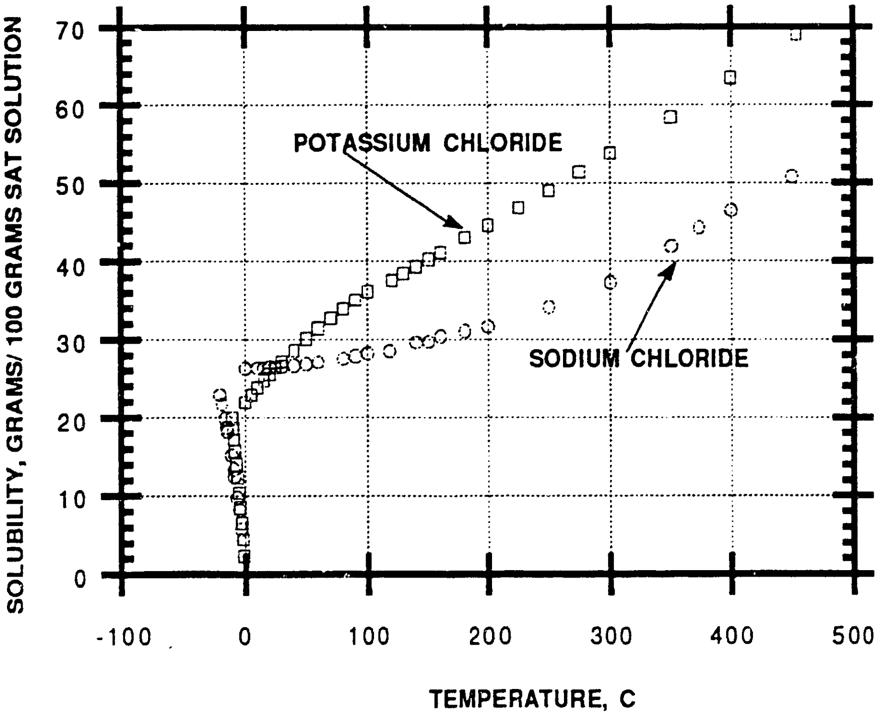

Figure 4. High-Temperature Solubility Data[7] of Sodium and Potassium Chlorides

leaching at 300°C and are intended for illustration purposes only. At this condition, the energy consumption of the process will be about 8% of that required by the evaporation process using a multiple-effect evaporator. This process requires additional equipment, such as a chiller, heat exchangers, and a centrifuge or an equivalent solid-/liquid-separation system.

The concept of leaching the salts at elevated temperatures can be integrated with more efficient separation methods, such as the Licon and freeze-crystallization processes that were discussed earlier. The large temperature gradient in the liquid/liquid heat exchanger enables the efficient recovery of a substantial portion of the heat in the salt-loaded stream. Another advantage of this process is that it conserves water. Except for the venting of gases generated during the leaching process (which will include small amounts of steam), the process is closed and thus the water is recycled.

Reaction of the high-temperature water with aluminum is not expected to be a major problem because aluminum constitutes about 3% of the salt cake. The possibility of hydrating the aluminum oxide at the elevated temperatures is a welcomed effect if it proceeds at a practical rate under practical operating conditions. The hydrated aluminum oxide could be used to produce aluminum, and this process will eliminate the salt cake problem altogether because every component of it will be recycled.

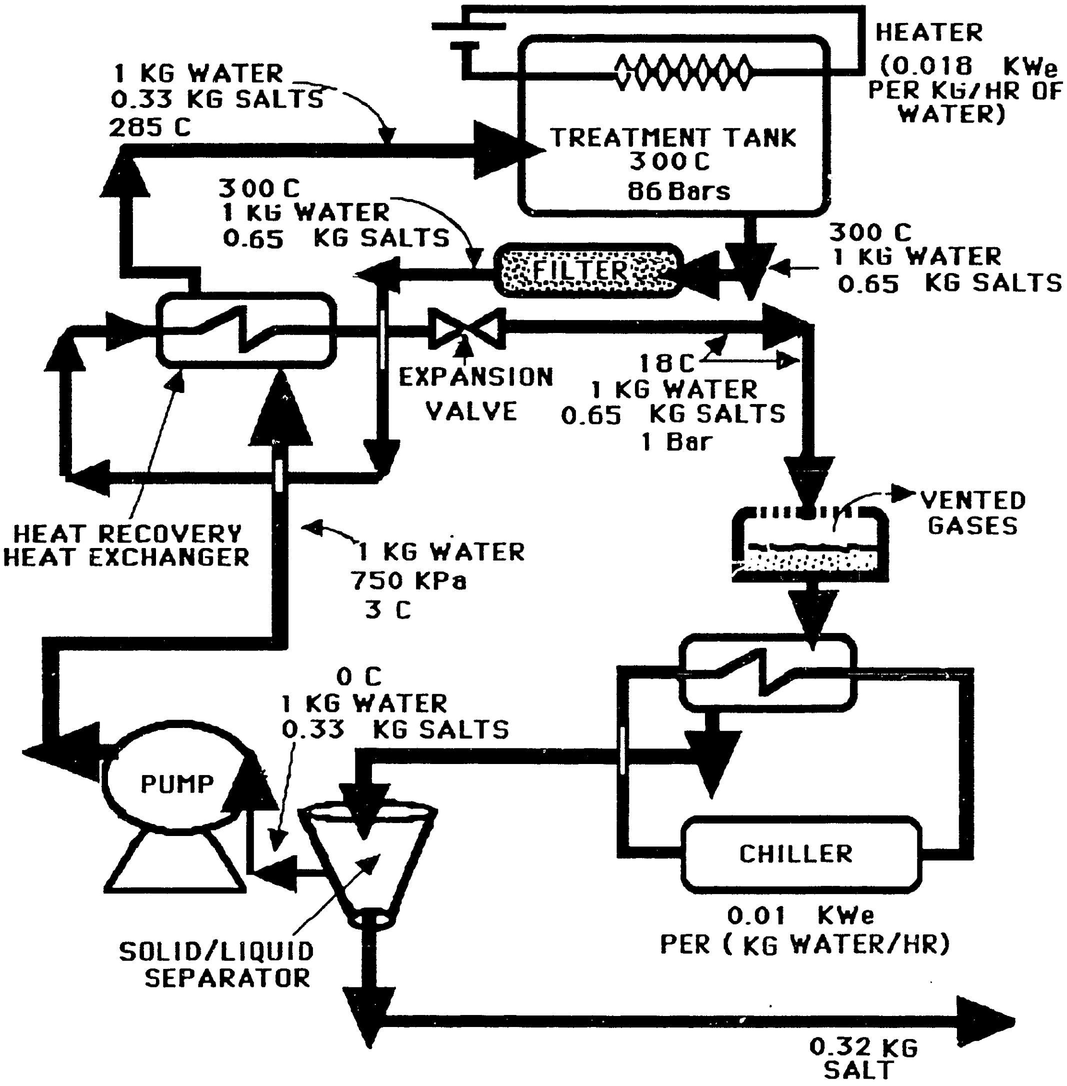

Figure 5. A Schematic Diagram of the High-Temperature, High-Pressure Process

Capillary-Effect Systems

While conducting various experiments on separating dissolved salts from aqueous solutions, we used smooth glass beakers that were less than half filled with salt solutions. As expected, when we left these beakers for some time in the hood, water evaporated from the solution, leaving behind the salts. We also observed the following:

1. Although the beakers were initially less than half filled with salt solutions, the entire internal surface of the beakers was covered with salt crystals. The thickness of the salt-crystal layer on the sides of the beaker continued to build up as long as an aqueous solution existed in the beaker.

2. The thickness of the salt-crystals layer appeared uniform, except on the bottom of the beakers where it was thicker. The thickness of the salt layer appeared to be a function of the salt concentration in solution.

3. One could easily collect the salt crystals by scraping the sides of the beaker. The inner crystals appeared more strongly attached than the outer ones.

From these observations, we developed a concept involving the use of capillary-effect and diffusion-enhanced dryers and heat-transfer equipment.

Process Concept

Porous or semiporous fibers can be used to facilitate the diffusion of water and salt ions and the eventual rapid drying of the salts. The increased surface area for heat and mass transfer promotes the rapid drying of the salts. The increased surface area can be designed to more than compensate for the lack of a large temperature gradient, which is normally the driving force in heat-exchange processes (such as drying operations). The mass-transfer process (water evaporation) can be accelerated further with the use of forced-convective techniques, such as blowing air across the diffusing front. Should waste heat be available, it can be used to further accelerate the rate of mass transfer.

One can collect the dried salts by scraping the outer layers or by agitating the porous media. Or, the dried salts can be air-blown into a collection vessel. Although this method results in the loss of the water, its operating cost should be very low. Because we have yet to perform quantitative experiments to confirm this concept's validity, we cannot discuss its merits in detail.

ECONOMICS OF BLACK-DROSS- AND SALT-CAKE-RECYCLING TECHNOLOGIES

The cost of recycling reusable constituents in black dross and salt cake, using currently available recycling systems, is very high. The key factors influencing the economics of these systems are the landfill disposal cost, amount and value of recyclable materials, and the capital and operating costs of the recycling process.

The concentration of salt and aluminum in the black dross has a direct impact on the energy consumption of the process and on the revenues generated from the sale of recycled materials. The analysis indicates that if the aluminum concentration is greater than 10%, the net revenues from the recovered aluminum and salts are sufficient to cover the operating and capital cost of the wet process. Capital charges ($/ton) are estimated on the basis of a charge rate of 40% of total capital cost. The two key constituents, aluminum and salt, are the major contributors to the potential revenues derived from the sale of the recyclable constituents. Figure 6 shows the economics of a typical wet-process system at various levels of aluminum concentrate.

As the concentration of aluminum decreases, revenues alone are insufficient to justify the costs of the process, and the economics become dependent on the credit for avoided landfill costs. The avoided landfill costs will have to be above $100/ton to economically justify recycling black dross or salt cake (in which case the aluminum is not recoverable) with an aluminum concentration of 3-5%. Figure 7 shows the break-even landfill costs at low concentrations of aluminum.

CONCLUSIONS

The following conclusions were made on the basis of the study:

1. Existing salt-recycling technology is economical in the United States when the aluminum content of the salt cake is more than about 10% or the landfill costs exceed $100/ton.
2. A considerable portion of the operating cost of the recycling process is the cost of energy required to evaporate the water in order to recover the salts.
3. The high-temperature and high-pressure process conserves energy and could result in the hydration of the aluminum oxide

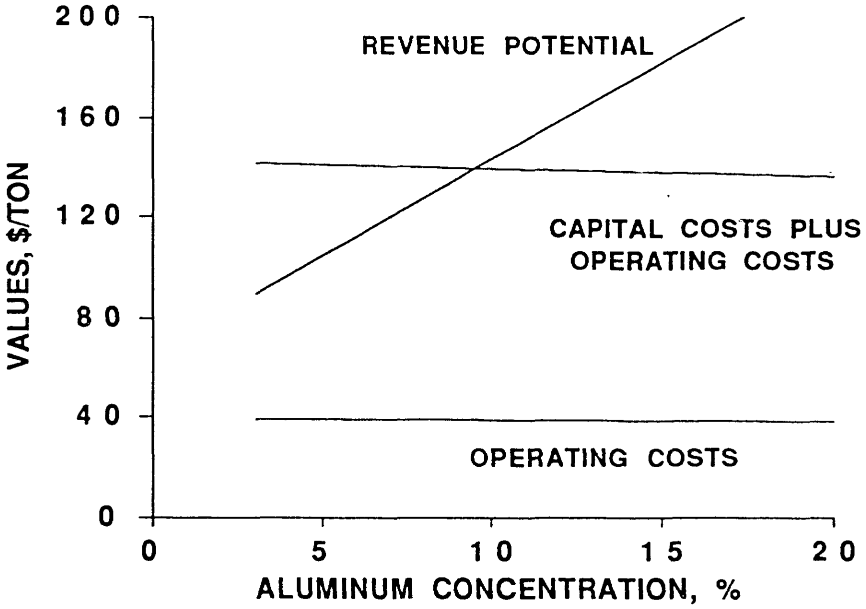

Figure 6. Wet-Process Economics versus Aluminum Concentration

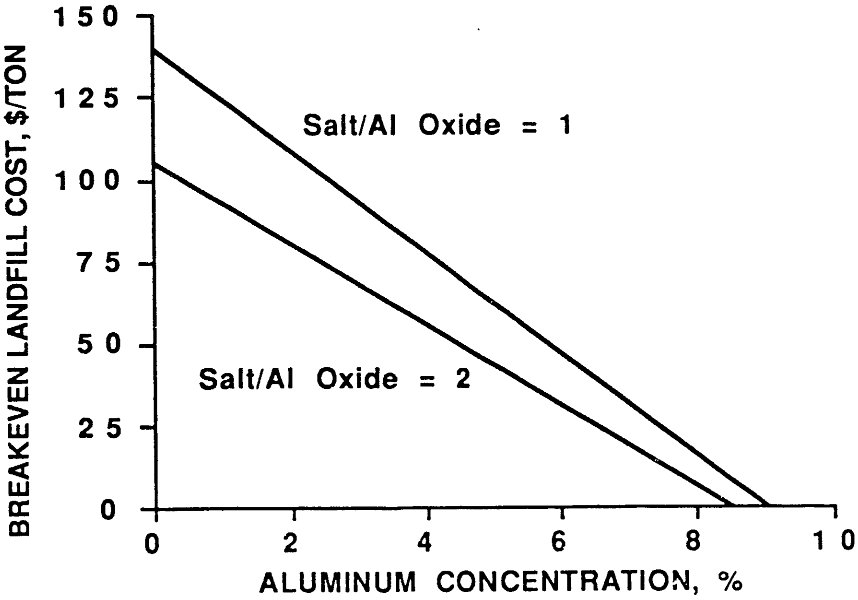

Figure 7. Break-Even Landfill Costs for Recycling Low-Aluminum-Concentration Black Dross/Salt Cake

ACKNOWLEDGMENT

This work is supported by the U.S. Department of Energy, Assistant Secretary for Conservation and Renewable Energy, under contract W-31-109-Eng-38.

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