Bureau of Mines Report of Investigations/1985

Hydrometallurgical Treatment of Electronic Scrap To Recover Gold and Silver

By H. E. Hilliard, B. W. Dunning, Jr., D. A. Kramer, and D. M. Soboroff

UNITED STATES DEPARTMENT OF THE INTERIOR

Report of Investigations 8940

Hydrometallurgical Treatment of Electronic Scrap To Recover Gold and Silver

By H. E. Hilliard, B. W. Dunning, Jr., D. A. Kramer, and D. M. Soboroff

UNITED STATES DEPARTMENT OF THE INTERIOR
Donald Paul Hodel, Secretary

BUREAU OF MINES
Robert C. Horton, Director

Library of Congress Cataloging in Publication Data:

CONTENTS

UNIT OF MEASURE ABBREVIATIONS USED IN THIS REPORT

A/ft^2	ampere per square foot	Mgal	thousand gallon
°Bé	degree Baumé	min	minute
°C	degree Celsius	mL	milliliter
g	gram	Mlb	thousand pound
g/L	gram per liter	pct	percent
gal	gallon	psig	pound per square inch, gauge
h	hour		
h/d	hour per day	rpm	revolution per minute
hp	horsepower	s	second
in^2	square inch	ton/d	ton per day
kg	kilogram	tr oz/ton	troy ounce per ton
kHz	kilohertz	vol pct	volume percent
kW·h	kilowatt-hour	W	watt
L	liter	W·h	watt hour
lb	pound	wt pct	weight percent
lb/d	pound per day	yr	year
lb/h	pound per hour		

HYDROMETALLURGICAL TREATMENT OF ELECTRONIC SCRAP
TO RECOVER GOLD AND SILVER

By H. E. Hilliard,[1] B. W. Dunning, Jr.,[2] D. A. Kramer,[1]
and D. M. Soboroff[3]

ABSTRACT

The Bureau of Mines investigated hydrometallurgical procedures for concentrating precious metals from a mechanically processed metal fraction of obsolete military electronic scrap. Feed material for the leaching procedures described in this report was the high-tension-separated metal concentrate obtained as a fraction from mechanically processing partially stripped electronic units. Initial pretreatment with a 20-wt-pct-NaOH solution solubilized most of the aluminum. After washing, the resulting residue was incinerated to destroy any organics attached to metal particles. The incinerated residue was leached with 20-vol-pct H_2SO_4 solution using two methods, countercurrent pressure and ultrasonic-aided leaching to solubilize the base metals, predominantly copper. Silver was recovered from the base metal leach residue using 50-vol-pct HNO_3, and gold contaminated with several percent palladium was recovered from the silver leach residue using aqua regia. The impure silver and gold plus palladium products represent about 1.5 wt pct of the high-tension-separated metal concentrate. Included is a three-step process evaluation study of mechanical processing of general electronic scrap and two hydrometallurgical procedures for leaching high-tension-separated metallic concentrates obtained from the mechanical processing step. Process evaluation concluded that only mechanical processing is economically viable.

---

[1]Chemist.
[2]Supervisory metallurgist.
[3]Research supervisor.
Avondale Research Center, Bureau of Mines, Avondale, MD.

INTRODUCTION

The recovery of precious metals from scrap supplies an important segment of domestic requirements for gold and silver. Bureau of Mines statistics estimate that U.S. refinery production of gold and silver in 1982 totaled 3.4 and 79 million tr oz, respectively (1).[4] Of this amount, 79 and 38 pct of the gold and silver, respectively, was reclaimed from secondary sources.

The Bureau of Mines investigated the recovery of precious metals from electronic scrap through a cooperative agreement with the Defense Property Disposal Service (DPDS). DPDS, an agency of the Department of Defense, is responsible for the disposal of millions of pounds of obsolete military electronics annually. The main purpose of the cooperative agreement with DPDS was to obtain an estimate of the amount and distribution of precious metals in military electronic scrap. In an initial step to obtain this information, a mechanical processing procedure was developed on a laboratory scale. Based on the laboratory results, a small pilot plant was developed for concentrating the precious metals into more valuable fractions at feed rates up to 500 lb/h (2-4). Products obtained included an iron-base fraction, an aluminum-base fraction, a nonmagnetic fraction, a tangled-wire fraction, and air classifier lights. The nonmagnetic fraction was further processed by high-tension separation (HTS) to recover metallic and nonmetallic fractions. The HTS metallic fraction represented about 17 pct of the total plant input for partially stripped general electronic scrap, yet contained approximately 51 and 56 pct of the gold and silver, respectively. This material with its abundant surface area is ideal for hydrometallurgical treatment to further concentrate the precious metals. The remaining 83 pct of the mechanically processed electronic scrap can be handled in the following fashion. The aluminum fraction (11 pct)

and the plus 1/2-in iron-base fraction (17 pct) can be recycled as aluminum and iron scrap. The minus 1/2-in iron-base fraction (11 pct) can be used as a copper cementing agent (5), recovering the precious metals from the washed residue, or as a copper blast furnace addition for the needed iron values, with the precious metals reporting to the black copper product. The remaining fractions (44 pct) would be ideal for pyrometallurgical treatment in a blast and/or converting furnace to produce copper-base bullion containing the remaining precious metals. The precious metals in the copper-base bullion would eventually be recovered from the slimes in the copper tank house circuit.

The purpose of the high-tension separation step and subsequent leaching of this fraction (6) is to further concentrate the precious metals, thus providing a more reliable sample for assay. The separation step also reduces the weight of material to ship to the refiner and the per pound toll refining charge. Traditionally, toll refiners have used several pyrometallurgical methods to recover precious metals from electronic scrap and electronic scrap concentrates. However, the small particle size resulting from mechanical processing makes the HTS metallic fraction ideal for hydrometallurgical treatment such as pressure and ultrasonic-aided leaching to remove base metals and further concentrate the precious metals. While the economics of the hydrometallurgical operations are not favorable, the operations are useful for providing a homogeneous concentrate from which an accurate precious metal assay can be determined.

The physical appearance of the HTS metallic fraction is shown in figure 1, and a chemical analysis is listed in table 1. The elemental composition of this fraction will vary depending on the type and vintage of electronic scrap processed. In earlier tests with a particlur lot of avionic scrap, the gold and silver concentrations in the HTS metallic fraction were equivalent to 42 and 505 tr oz/ton, respectively (3). The fraction used in

---

[4]Underlined numbers in parentheses refer to items in the list of references preceding the appendix.

this investigation contained 36 tr oz/ton gold and 401 tr oz/ton silver. The avionic scrap from which this fraction was recovered consisted of such items as radio receivers, amplifiers, transmitters, and decoders. Many of the items had been partially stripped of easily removed components such as circuit cards that contain precious metals. Prior to acid leaching, the HTS metallic fraction was leached with 20-wt-pct NaOH; the residue from this operation was washed and then incinerated to destroy any organics attached to the metallic particles. The caustic leaching of the HTS metallic fraction, as well as the mechanical beneficiation, are discussed in greater detail in other publications (2-4, 6).

TABLE 1. -- Elemental composition of HTS metallic fraction

	Conc, wt pct		Conc, wt pct
Ag	1.37	Fe	9.4
Al	27.2	Ni	3.2
Au	.12	Pb	.2
Cr	.1	Sn	2.1
Cu	38.4	Remainder	(1)

[1]Mostly Si as SiO_2 (fiberglass and plastic filler).

FIGURE 1. - High-tension-separated metal fraction.

EXPERIMENTAL EQUIPMENT AND PROCEDURE

Two leaching systems to solubilize the HTS metallic fraction using 20-vol-pct H_2SO_4 were studied. One system used pressure leaching and mechanical stirring; the other leach was at atmospheric pressure but stirred with ultrasonic energy.

PRESSURE LEACHING

All pressure-leaching tests were conducted in titanium autoclaves of 2-L capacity. Vigorous agitation and rapid gas dispersion were achieved using a stirring shaft with two turbine-type impellers turning at 600 rpm. An internal thermowell, support brackets, and dip tubes provided baffling so that continuous turbulence of the leaching reagent was maintained in the autoclaves. Leach tests were conducted in three-stage countercurrent steps with solid-liquid separation between each stage. Measured quantities of residue from NaOH leaching (~120 g) and 1 L second-stage depleted acid were charged into the first-stage autoclave and then heated to selected temperatures and pressures. The third-stage autoclave was charged with 1 L of fresh 20-vol-pct H_2SO_4. Leach cycle times studied were 1.5, 2, and 4 h per stage. Air was sparged through the autoclaves during all tests. After leaching, the slurry was removed from each autoclave and filtered, and the products were analyzed. A flow diagram of the process is shown in figure 2.

Fresh 20-vol-pct H_2SO_4
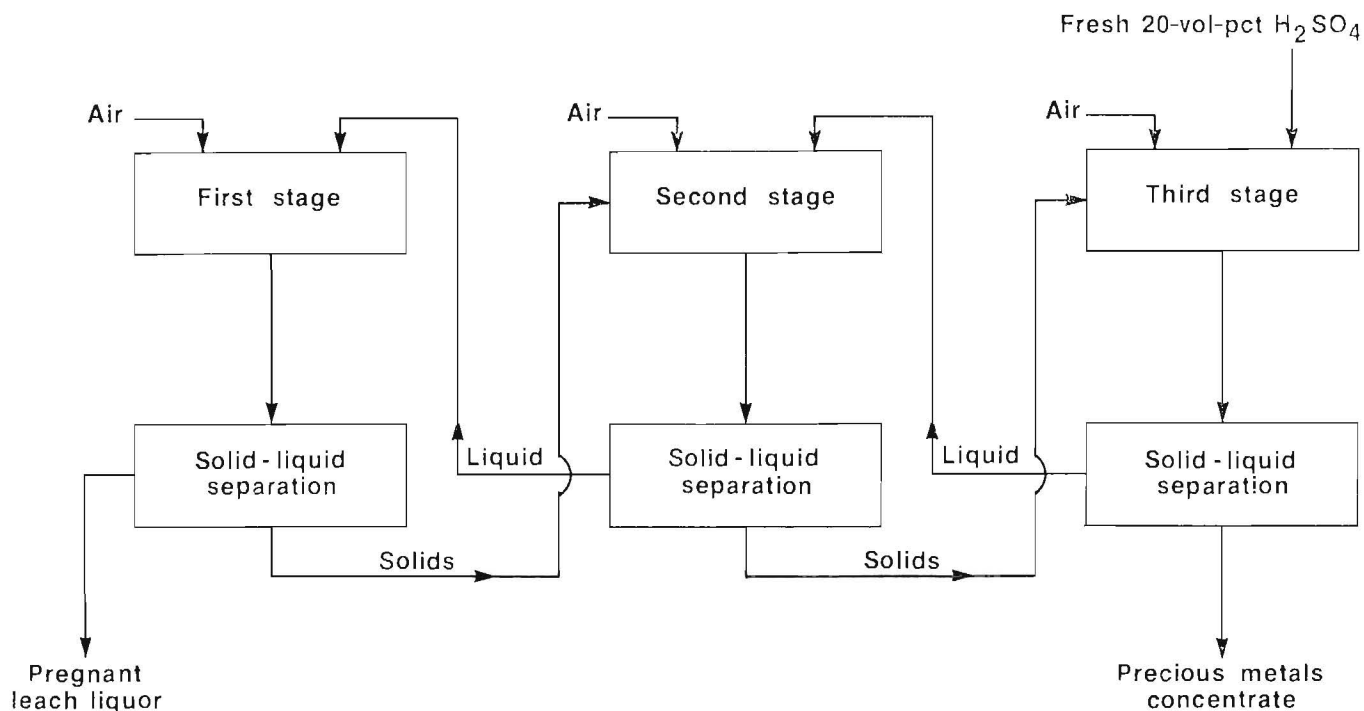
FIGURE 2. - Flow diagram for countercurrent leaching process.

ULTRASONIC--AIDED LEACHING

Ultrasonic-aided leach tests were conducted using a magnetostrictive ultrasonic supply. This unit delivered 900 W of electrical power at the resonance frequency of 20 kHz to a transducer tank remote from the generator via a connecting cable. Measured quantities of residue from NaOH leaching (~120 g) and 1 L of 20-vol-pct H_2SO_4 were added to a 2-L stainless steel beaker. The beaker was clamped inside the transducer tank, water was added to the same level as the liquid inside the beaker, and the power was turned on. All tests were single stage and conducted at atmospheric pressure. Extraction efficiency was determined for 0.5-, 1-, 2-, and 3-h leach times. The temperature was allowed to find its own level (70° to 75° C).

REGENERATION OF SULFURIC ACID AND IRON METAL RECOVERY

After copper cementation by a magnetic scrap fraction, the spent H_2SO_4 leach liquor containing mostly ferrous iron was recycled and iron metal was recovered by electrolysis in a cell containing an anode and cathode chamber separated by a cation-selective membrane. The anolyte containing aqueous ammonium sulfate and the catholyte (spent leach liquor) were circulated through the cell with a peristaltic pump. H_2SO_4 was regenerated in the anode compartment, and iron plated on the cathode. Figure 3 is a photograph of the cell. The anode was a 4- by 4-in sheet of 6 pct antimonial lead; the cathode, a 4- by 4-in sheet of stainless steel; and the membrane, a 4- by 4-in sheet of

5
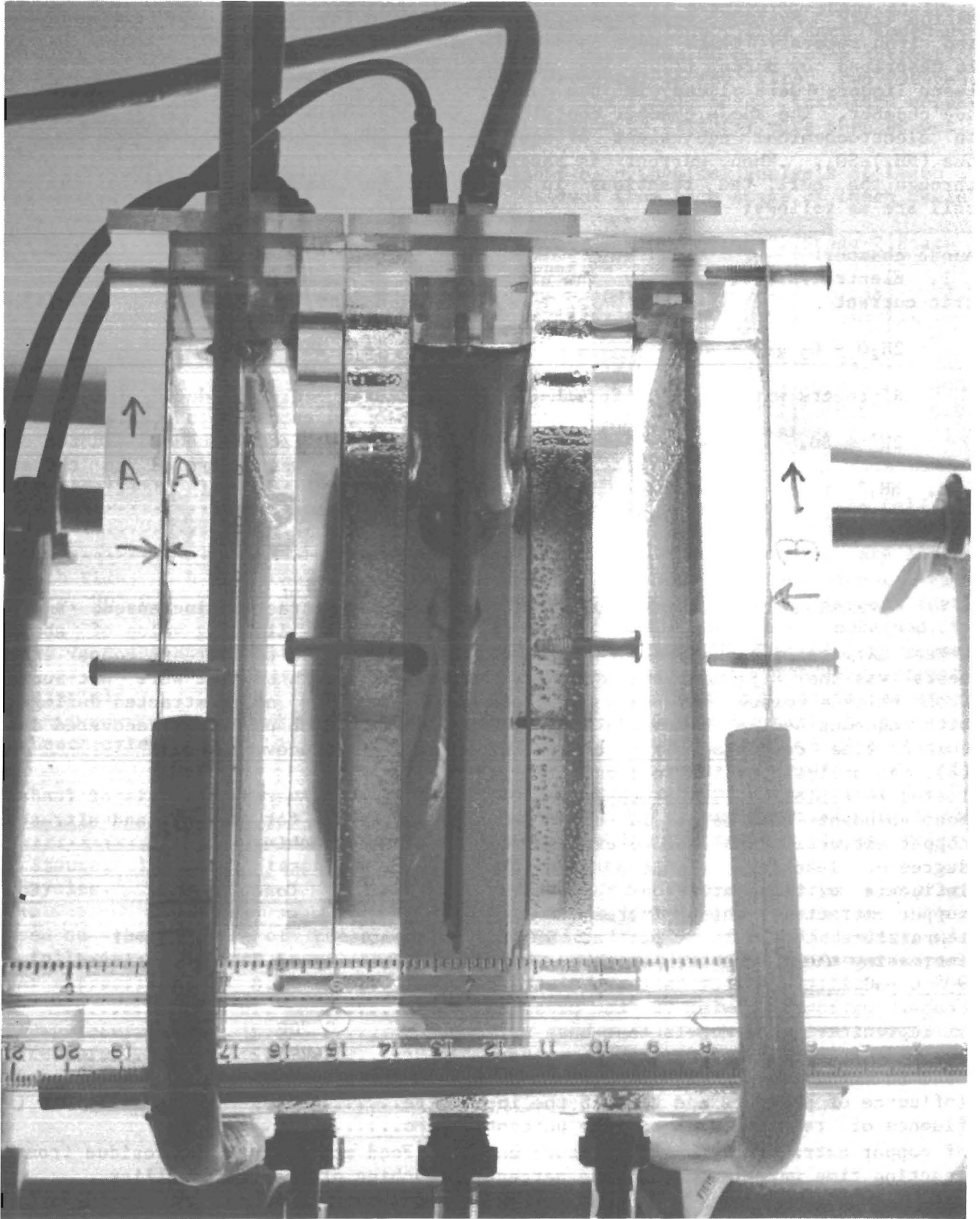
FIGURE 3. - Sulfuric acid regeneration cell.

Nafion 423.[5] Sulfuric acid regeneration and iron recovery tests were conducted as described by Burke (7). The spent leach liquors were placed in the cathode chamber. The anode chamber contained an electrochemical equivalent of aqueous $(NH_4)_2SO_4$. When current is passed through the cell, the reactions in the cell are as follows:

Anode chamber:
1. Electrolysis of water by the electric current

$$2H_2O \rightarrow O_2 \text{ gas} + 4H^+$$

H^+ reacts with SO_4^{--} to form H_2SO_4

$$2H^+ + SO_4^{--} \rightarrow H_2SO_4$$

2. NH_4^+ ions migrate from the anode chamber through the cation-selective membrane into the cathode chamber, where they react with SO_4^{--} ions to form $(NH_4)_2SO_4$.

Cathode chamber:
3. $2NH_4^+ + SO_4^{--} \rightarrow (NH_4)_2SO_4$
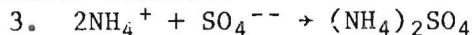
4. Fe^{++} is reduced to Fe° at the cathode.

$$Fe^{++} + 2e \rightarrow Fe^\circ$$
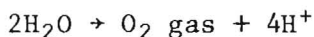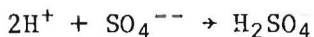
5. H_2 gas evolution

$$2e + 2H^+ \rightarrow H_2 \text{ gas}$$
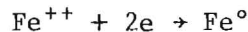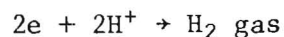
Sulfuric acid produced in the anode chamber was returned to the leach cycle; ammonium sulfate produced in the cathode chamber was recycled to the anode chamber.

RESULTS AND DISCUSSION

PRESSURE-LEACHING RESULTS

Feed material for the pressure-leaching tests was the HTS metallic concentrate from which aluminum had been extracted with aqueous sodium hydroxide and the combustibles destroyed by incineration (6). An analysis of the feed material is listed in table 2. Since copper was the most abundant base metal in the feed, copper extraction was used to express the degree of leaching. Table 3 shows the influence of temperature on the percent copper extracted, which increased with temperature to about 90 pct at 90° C. Increasing the leach temperature above 90° C had little effect on the percent of copper extracted, owing to the presence of cupronickels and Monels that have very limited solubility for the parameters used in these tests. Table 4 shows the influence of pressure and table 5 the influence of reaction time on the percent of copper extracted. As the pressure and reaction time were increased, the percent of copper extraction increased, again to the apparent limiting value of about 90 pct. Attempts to increase copper extraction beyond this point were not successful. Copper not extracted during the pressure-leach phase was recovered during the silver recovery phase.

TABLE 2. – Average analysis of feed material[1] for pressure and ultrasonic-aided leaching

Element:	Conc, wt pct		Element-- Continued	Conc, wt pct
Ag.......	1.74			
Al.......	2.8		Ni.........	4.1
Au.......	.16		Pb.........	.3
Co.......	.2		Pd.........	.1
Cr.......	.2		Sn.........	2.2
Cu.......	48.9		W..........	.1
Fe.......	14.9		Remainder..	$(^2)$
Mo.......	.3			

[1]Feed material is the residue from NaOH leaching of the HTS metallics.
[2]Si as SiO_2 and additional oxygen as metal oxides formed during the incineration step.

---

[5]Reference to specific products does not imply endorsement by the Bureau of Mines.

TABLE 3. – Influence of leach temperature on extraction of copper from feed material[1]

Temp, °C	Extraction, pct
50..........................	63
70..........................	86
90..........................	89
110..........................	90

[1]Pressure, 100 psig; leach time, 4 h per stage.

TABLE 4. – Influence of pressure on extraction of copper from feed material[1]

Pressure, psig	Extraction, pct
39..........................	69
50..........................	88
80..........................	89
100..........................	89

[1]Leach time, 4 h per stage; leach temperature, 90° C.

TABLE 5. – Influence of reaction time on extraction of copper from feed material[1]

Leach cyle time, h	Extraction, pct
1.5..........................	70
2..........................	79
4..........................	89

[1]Pressure, 100 psig; temperature, 90° C; 3 cycles.

Based on the results of these tests, the following conditions gave the best copper extraction:

Temperature	80° to 90° C
Pressure	50 to 80 psig
Leach cycles	4 h, three cycles

Table 6 lists a stage-by-stage analysis of the leach liquor from each leach cycle. More than 80-pct copper extraction was obtained in the first two stages, while the final stage increased copper extraction to slightly more than 89 pct.

TABLE 6. – Chemical analysis of leach liquor from each stage of leach cycle

	Stage 1	Stage 2	Stage 3
Extraction pct Cu..	46	36	7.8
Conc in filtrate, g/L:			
Cu.............	51.2	23.4	4.2
Ag.............	<0.0001	<0.0001	<0.0004
Al.............	1.2	0.07	0.01
Fe.............	0.5	1.4	1.4
Ni.............	0.2	0.5	0.4

Results of larger scale (16-fold increase in the incinerated residue charge used in other tests) leaching are listed in table 7. Distributions of metal values between the leach solutions and residues are similar to those obtained in small-scale tests. Most of the copper and soluble nickel and iron were extracted, while tin, lead, and stainless steel reported to the residue with the precious metals. The residue (320.6 g) contained about 29 g of precious metals. Experimental conditions for larger scale leaching were 1,900 g feed material, 16 L 20-vol-pct H_2SO_4, 3- to 4-h leach cycles, 90° C, 100 psig.

Copper was recovered from the pregnant leach liquor by cementation with a magnetic fraction of shredded electronic scrap. An analysis of the cement copper is listed in table 8. The cement copper was of acceptable quality for further commercial recovery operations, such as reverberatory smelting in a copper refinery.

TABLE 7. - Concentration and distribution of metal values in leach solution and residue

Products	Concentration	Distribution, pct
Leach solution:	g/L	
Ag.............	<0.001	<0.004
Al.............	.002	<.001
Au.............	ND	ND
Cr.............	.0004	6.2
Cu.............	36.3	89.3
Fe.............	2.9	29.0
Ni.............	1.02	38.8
Pb.............	.003	1.4
Sn.............	.14	7.9
Residue:	wt pct	
Ag.............	8.13	99.1
Al.............	.4	.2
Au.............	.74	>99.9
Cr.............	24.4	10.9
Cu.............	.42	94.0
Fe.............	39.6	71.0
Ni.............	9.0	61.1
Pb.............	1.1	98.6
Pd.............	.21	99.9
Sn.............	9.5	91.1

ND Not detected.

TABLE 8. - Chemical analysis of cement copper

Element	Conc, wt pct
Ag................	0.001
Al................	<.01
Au................	<.0001
Cu................	91
Fe................	1.0
Pb................	.01
Sn................	.6

PRECIOUS METAL SEPARATION

Silver was recovered from the pressure-leaching residue by dissolution in 50-vol-pct HNO_3 and subsequently precipitated with sodium chloride as silver chloride. Silver chloride was reduced to silver metal by mixing with sodium carbonate and heating the mixture to 600° C (8). Copper not extracted during H_2SO_4 leaching was recovered from the silver-free HNO_3 solution by cementation with iron scrap. Finally, gold was extracted

from the HNO_3 leaching residue with aqua regia and precipitated with sodium bisulfite. An analysis of the gold and silver products and the final residue is listed in table 9. The final residue, weighing 194 g, consisted mostly of acid-insoluble noncombustibles (mostly silica) used in the manufacture of plastics and printed circuit boards. Stainless steel and tin made up the remainder.

ULTRASONIC-AIDED-LEACHING RESULTS

Figure 4 shows a plot of percent copper extracted versus reaction time during ultrasonic-aided leaching. About 38-pct extraction was obtained in the first 30 min, but subsequent dissolution of copper was very slow. The coextraction of silver is also plotted. Silver extraction was less than 1 pct after 240 min of leaching. In exploratory tests, HNO_3 was added to the H_2SO_4 solution as an oxidizing agent. Figure 5 shows the influence of HNO_3 concentration on the extraction of copper and silver. An extraction of 94 pct copper was obtained after a 2-h leach time. Coextraction of silver was about 7 pct. Treatment of leach residues and leach liquors was the same as discussed under pressure leaching.
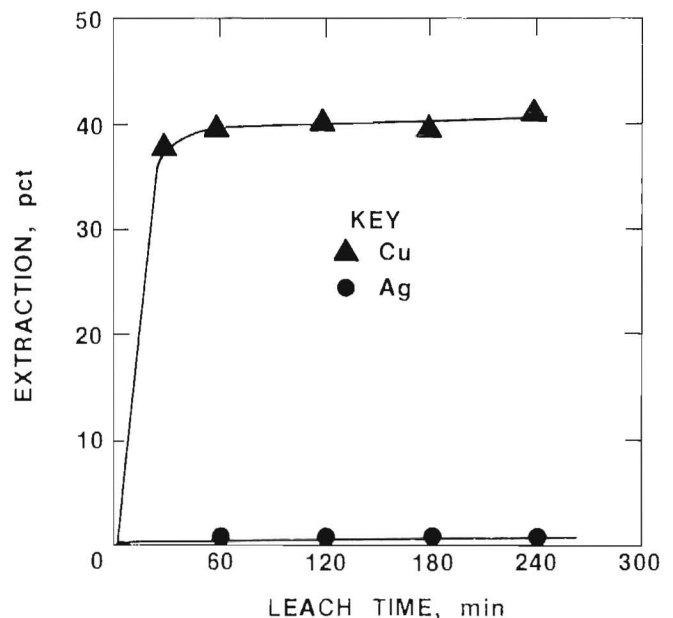
FIGURE 4. - Influence of reaction time on copper extraction.

TABLE 9. - Spectrochemical analysis of products, weight percent

Product	Gold	Cement silver[1]	Final residue
Ag................................	0.03 - 0.3	>10	0.0003- 0.003
Al................................	.003- .03	.03 -0.3	.3 - 3
Au................................	>10	.0003- .003	ND
Cr................................	.003- .03	ND	.3 - 3
Cu................................	.003- .03	.03 - .3	.1 - 1
Fe................................	.03 - .3	.03 - .3	.3 - 3
Ni................................	.003- .03	.003 - .3	.01 - .1
Pb................................	.01 - .1	.3 --3	.1 - 1
Pd................................	1 -10	ND	ND
Si................................	.1 - 1	.1 -1	>10
Sn................................	.03 - .3	.3 -3	1 -10

ND Not detected.

[1]Quantitative analysis of cement silver: 97.5 Ag, 0.028 Al, 0.53 Cu, 0.12 Pb, and 1.82 Sn.

REGENERATION OF SULFURIC ACID

Burke (7) investigated the regeneration of H_2SO_4 and the recovery of iron metal from spent pickle liquors at current densities from 32 to 129 A/ft^2. A current density of about 60 A/ft^2 was reported to
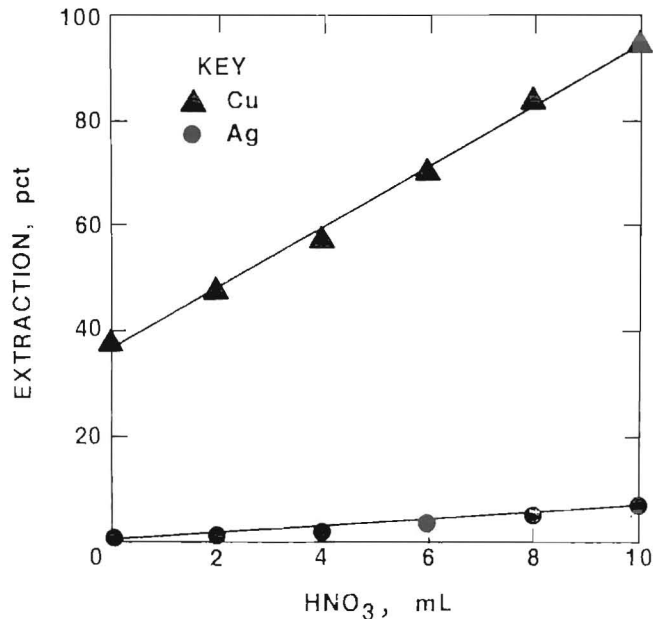
FIGURE 5. Influence of nitric acid addition to leach solution on copper and silver extraction.

be most economical. Burke determined that electrolysis at 32 A/ft^2 consumed less power (501 W·h versus 466 W·h). However, electrolysis at about 60 A/ft^2 was preferable because the current efficiency is much higher and electrolysis is completed in less than half the time required at 32 A/ft^2 (12 h versus 28 h). This was confirmed in our laboratory. Consequently, all results reported here were obtained at a current density of 58 A/ft^2. The catholyte was spent leach solution containing 107 g/L iron and 3.4 g/L H_2SO_4. The anolyte contained 368 g/L ammonium sulfate. The electrolytes were circulated through the anode and cathode chamber and electrolyzed until essentially all iron was deposited. The pH in the cathode compartment was maintained between 1.3 and 3.0 with dilute H_2SO_4. Electrolysis was carried out at room temperature. Results from three tests are reported in table 10. In practice, electrolysis is continued until the H_2SO_4 concentration in the anode chamber reaches the desired level. This is accomplished by adjusting the ammonium ion content of the anolyte and the ferrous ion content of the catholyte.

TABLE 10. - Cathode efficiency for iron deposition and H_2SO_4 regeneration[1]

Test	Run 1	Run 2	Run 3
Membrane current density..........................A/ft^2..	57.6	57.6	57.6
Ambient temperature....................................°C..	30	30	30
Cell temperature.......................................°C..	40	40	40
Hours to completion...	8	4	15
Cathode efficiency.....................................pct..	87	86	95.1
W·h/L of solution..	518	518	485.6
kW·h/kg Fe deposited.......................................	5.0	4.9	4.5
kW·h/kg H_2SO_4 regenerated..............................	3.2	3.6	2.3
H_2SO_4 in final electrolyte.........................g/L..	161.2	72.3	226.9
Iron deposited...g..	104.6	52.3	213.7

[1]Leach solution: Run 1, 1 L; run 2; run 2, 500 mL; run 3, 2 L.

ECONOMICS

An economic evaluation was made for the mechanical processing of electronic scrap followed by hydrometallurgical leaching procedures with HTS metallic concentrate. In the evaluation, the process was divided into three steps, and the economics of each considered individually. Detailed process descriptions and economics are presented in the appendix.

The first step is a mechanical beneficiation technique, which divides the bulk scrap into a variety of upgraded products (2-5). The second step utilizes one product, the HTS metallic concentrate, which contains greater than 50 pct of the gold and silver found in the original sample, to obtain a precious metal concentrate through chemical separation techniques. In the final step, the precious metal concentrate from the second step is chemically treated to separate gold and silver and recover these as relatively pure metals.

Of the three sequential steps in the proposed process, the maximum return on investment after taxes (73 pct) is obtained by performing the first step and sending the product to a precious metal refiner. Precious metal concentration by the second step produces a return on investment after taxes of 6 pct, and the third step operates at a loss. Proposed modifications (seeding the aluminum hydroxide to reduce precipitation time and using fresh sulfuric acid instead of regenerating the acid) to the second step do not reduce the processing cost sufficiently to permit the estimated profit to reach a 15-pct return on investment.

A potential problem in processing this type of heterogeneous scrap is obtaining a reliable analysis of the precious metal content. Since the product from step 2 is more homogeneous than the product from step 1, a more reliable precious metal assay can be determined from the step 2 product. If the precious metal composition of the scrap is reproducible, constraints (i.e., minimum precious metal recoveries) can be placed on a refiner so that the maximum value is obtained for the precious metals. These constraints generally cannot be imposed if the precious metal content of the scrap is not known. Therefore, performing step 2 may increase the revenue returned from a precious metal refiner. A similar case can be made for performing step 3. However, no data are available to quantify the additional value realized by increasing product homogeneity.

SUMMARY

Mechanically processed HTS metallic fraction that has been pretreated with aqueous NaOH to remove the aluminum and incinerated to destroy the organics was leached with 20-vol-pct H_2SO_4 to remove most of the soluble base metals. Relatively pure silver was removed from the H_2SO_4 leach residue using nitric acid. Impure gold contaminated with palladium was recovered from the silver leaching residue with aqua regia. The impure silver and gold products represent about 1.5 pct of the original weight of the HTS metallic fraction. The final leach residue is primarily silica, but does contain recoverable amounts of tin. While the economics of the hydrometallurgical processing are not favor-able, it provides a way to assess the value of the scrap treated.

REFERENCES

1. U.S. Bureau of Mines. Mineral Commodity Summaries 1983. Pp. 60-61, 140-141.

2. Ambrose, F., and B. W. Dunning, Jr. Precious Metals Recovery From Electronic Scrap. Paper in Proceedings of the Seventh Mineral Waste Utilization Symposium (Chicago, IL, Oct. 20-21, 1980). IIT Res. Inst., Chicago, IL, 1980, pp. 184-197.

3. _____. Mechanical Processing of Electronic Scrap To Recover Precious-Metal-Bearing Concentrates. Ch. in Precious Metals, ed. by R. O. McGachie and A. G. Bradley. Pergamon, 1981, pp. 67-76.

4. Dunning, B. W., Jr., F. Ambrose, and H. W. Makar. Distribution and Analyses of Gold and Silver in Mechanically Processed Mixed Electronic Scrap. BuMines RI 8788, 1983, 17 pp.

5. Salisbury, H. B., L. J. Duchene, and J. H. Bilbrey, Jr. Recovery of Copper and Associated Precious Metals From Electronic Scrap. BuMines RI 8561, 1981, 16 pp.

6. Hilliard, H. E., B. W. Dunning, Jr., and H. V. Makar. Hydrometallurgical Treatment of Electronic Scrap Concentrates Containing Precious Metals. BuMines RI 8757, 1983, 15 pp.

7. Burke, Z. L. Recovery of Spent Pickle Liquor and Iron Metal. U.S. Pat. 4,149,946, Apr. 1979.

8. Kunda, W. Treatment of Complex Silver Arsenide Concentrate in Nitric Acid System. Can. J. Chem. Eng., v. 59, June 1981, pp. 347-356.

9. Weaver, J. B., and H. C. Bauman. Cost and Profitability Estimation. Sec. 25 in Perry's Chemical Engineers' Handbook, ed. by R. H. Perry and C. H. Chilton. McGraw-Hill, 5th ed., 1973, p. 47.

12

APPENDIX.--ECONOMIC EVALUATION

PROCESS DESCRIPTIONS

Brief descriptions of each of the three steps of the proposed process are

included here. Process flowsheets for each of the three steps are provided in figures A-1 through A-3.

Bulk electronic scrap

SHREDDING → DUST COLLECTION → Light fraction

AIR CLASSIFICATION → Wire fraction

MAGNETIC SEPARATION → Magnetic fraction

SCREENING

Minus 1/2-in nonmagnetic scrap

Plus 1/2-in nonmagnetic scrap

CRUSHING

SCREENING → Wire fraction

Minus 1/4-in scrap → HIGH-TENSION SEPARATION → Metallic fraction

Nonmetallic fraction

Plus 1/4-in scrap

EDDY-CURRENT SEPARATION

Mixed metallic and nonmetallic fraction

Copper alloys and cast aluminum fraction

Aluminum fraction

FIGURE A-1. - Bulk electronic scrap separation section (step 1).

Metallic fraction
of high-tension
separator scrap   Air

CAUSTIC LEACHING

Air   Offgases

Wash
water   FILTRATION AND WASHING   Filter cake   INCINERATION

Makeup
NaOH

AGING   SULFURIC ACID LEACHING   Air

Wash
water   FILTRATION AND WASHING   Washings

Wash
water   FILTRATION AND WASHING   Precious metals
concentrate

NaOH

Fe   COPPER CEMENTATION   Air

Water
vapor   DRYING

Water
vapor

Wash
water   FILTRATION AND WASHING   Filter cake   DRYING   Cement
copper

Al(OH)$_3$

Makeup
H$_2$SO$_4$

Fe   ELECTROLYSIS   O$_2$

Sludge

H$_2$SO$_4$ CONCENTRATION   Water
vapor

FIGURE A-2. - Precious metal concentration section (step 2).

Bulk Electronic Scrap Separation Section (Step 1)

As-received electronic scrap is shredded in a 75-hp hammer mill at a feed rate of 6 ton/d to reduce the size to about 1-1/2 in, and then fed to an air classifier where it is separated by differences in density. Light material entrained in the air stream is removed by a dust collector. The heavy material is conveyed to a magnetic separator where the magnetic fraction is separated from the bulk scrap and recovered as a product. Nonmagnetic material from the magnetic separator is sized on a 1/2-in vibrating screen. Plus 1/2-in nonmagnetic scrap is stored as an intermediate product, while the minus 1/2-in material is fed to a 15-hp roll crusher.

After the ceramic and nonmetallic portions of the minus 1/2-in scrap are broken up in the roll crusher, the scrap falls onto a 1/4-in vibrating screen.

Wire collects in bunches on the screen and is easily removed by hand. The plus 1/4-in fraction from the screen and the plus 1/2-in nonmagnetic scrap intermediate product are fed to an eddy-current separator, which is a batch operation where three products are recovered--an aluminum fraction, a copper alloy and cast aluminum fraction, and a mixed metallic-nonmetallic fraction.

Minus 1/4-in scrap from the vibrating screen is combined with the mixed metallic and nonmetallic fraction from the eddy-current separator. These are fed to a high-tension separator, operating batchwise, where the material is separated by its ability to hold an induced charge. Two products are recovered from this separation--a nonmetallic fraction and a metallic fraction. About 1,200 lb/d of the metallic fraction is recovered, which contains significant quantities of precious metals and is used as feed to the second step.

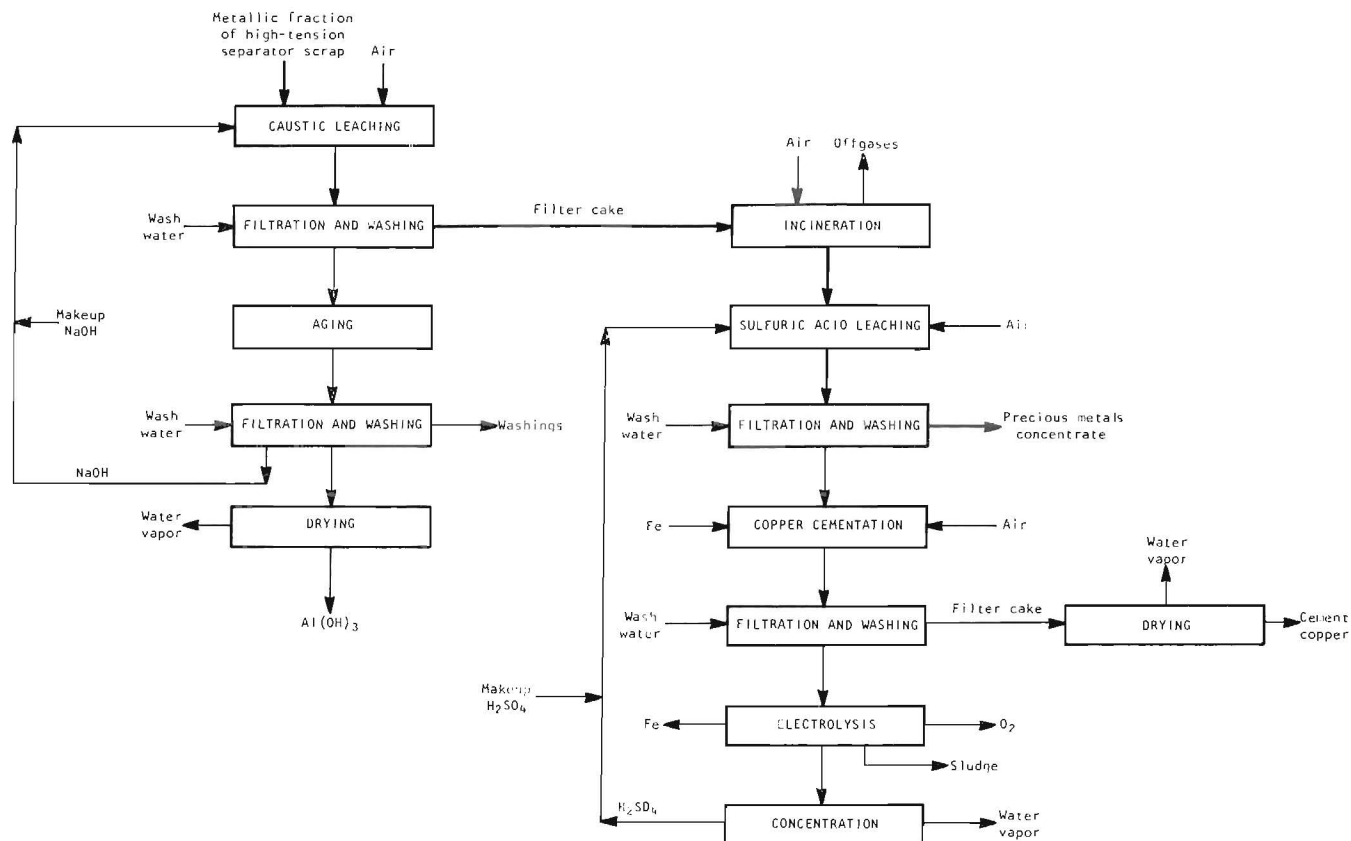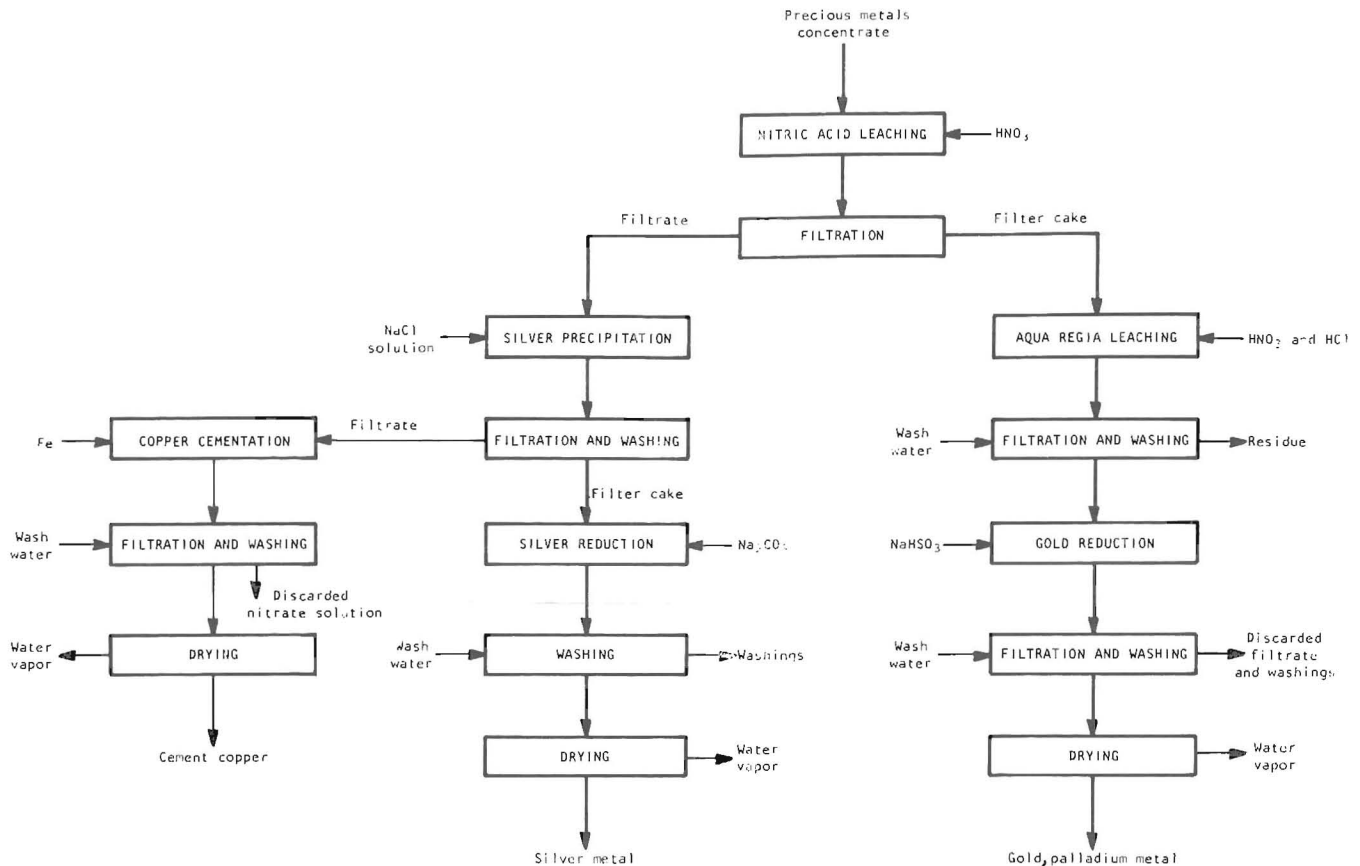FIGURE A-3. - Precious metal separation section (step 3).

Precious Metal Concentration
Section (Step 2)

In the second process step, the metallic fraction recovered from the high-tension separator in step 1 is leached in an 800-gal tank with a 20-wt-pct caustic solution to dissolve most of the aluminum. The leach slurry is then filtered, and the solids are washed. Aging the combined filtrate and washings from 10 to 12 weeks causes aluminum hydroxide to precipitate. Aluminum hydroxide solids are recovered from the slurry by filtration, dried, and stored. The caustic filtrate is recycled to leaching after makeup caustic is added to replace losses.

The filter cake from austic leaching is incinerated in an electrically heated furnace to burn off any organics that are left in the solids. Pressure leaching the remaining solids with a 20-vol-pct H_2SO_4 solution in a 550-gal autoclave dissolves copper. This leach slurry is

filtered, and the filter cake, containing most of the precious metals, is washed. These solids are recovered as a precious metal concentrate (1,250 lb/d), which may be sold as a product or further processed in step 3. Copper-containing filtrate and washings are contacted with the minus 1/2-in metallic scrap recovered in step 1 to precipitate about 415 lb/d of a cement copper coproduct.

The copper-free solution is treated by electrolysis to remove the dissolved iron as a solid and regenerate the sulfuric acid leach solution. This solution is concentrated by evaporation and recycled to acid leaching.

Precious Metal Separation
Section (Step 3)

The precious metals concentrate produced in step 2 is leached with HNO_3, in a 165-gal tank to dissolve silver and copper. This leach slurry is filtered, and the filtrate is mixed with a sodium

chloride solution to precipitate the silver as a chloride. Silver chloride solids are separated from the copper-containing solution and washed. These solids are then heated with soda ash, which reduces the silver chloride to metal, and any remaining solids salts are dissolved by washing the silver metal. About 18 lb/d of this silver precipitate is dried and stored as a product for shipment to a precious metal refiner. Cement copper is recovered from the remaining solution as in step 2 (49 lb/d).

Filter cake from nitric acid leaching, which contains gold and palladium, is leached in a 170-gal tank with aqua regia, dissolving these metals as chlorides. The solids are separated by filtration and discarded. Saturating the chloride filtrate with sodium bisulfite precipitates gold and palladium as metals. These metals are recovered from the solution by filtration and washed. The mixed gold and palladium metal precipitate (2 lb/d) is dried and stored as a product for shipment to a precious metal refiner.

ECONOMICS

Estimated capital and operating costs are based on the preceding process descriptions and are presented separately for each step.

Capital Costs

The capital cost estimate is of the general type called a study estimate by Weaver and Bauman (9).[1] This type of estimate, prepared from a flowsheet and a minimum of equipment data, can be expected to be within 30 pct of the actual cost for the plant described. The estimated capital costs for each of the three steps are shown in table A-1 on a third quarter 1983 basis (Marshall and Swift (M and S) index of 764.7). Plants designed for each of the three steps operate 1 shift per day, 5 days per week.

---

[1]Underlined numbers in parentheses refer to items in the list of references preceding the appendix.

Equipment costs for the process are based on cost-capacity data and manufacturers' cost quotation. Costs data are brought up to date by use of inflation indices. In developing the plant capital costs, corrosion-resistant materials of construction are used where appropriate.

Factors for piping, etc., except for the foundation and electrical factors, are assigned to each step, using as a basis the effect fluids, solids, or a combination of fluids and solids may have on the process equipment. Foundation costs are estimated for each piece of equipment individually, and a factor for the entire step is calculated from the totals. The electrical factor is based on motor horsepower requirements for each step. A factor of 10 pct, referred to as miscellaneous, is added to each step to cover minor equipment and construction costs that are not shown with the equipment listed.

For each step, the field indirect cost, which covers field supervision, inspection, temporary construction, equipment rental, and payroll overhead, is estimated at 10 pct of the direct cost. Engineering cost is estimated at 10 pct, and administration and overhead cost is estimated at 5 pct of the construction cost. A contingency allowance of 15 pct and a contractor's fee of 5 pct are included in the step cost.

The costs of plant facilities and plant utilities are estimated as 5 and 1 pct, respectively, of the total process costs and include the same field indirect costs, engineering, administration and overhead, contingency allowance, and contractor's fee as are included in the section costs. Included under plant facilities are the costs of material and labor for auxiliary buildings such as offices, shops, laboratories, and cafeterias, and the cost of nonprocess equipment such as office furniture, together with safety, shop, and laboratory equipment. Also included are labor and material costs for site preparation such as site clearing, grading, drainage, roads, and fences. The cost of water, power, and steam distribution systems is included under plant utilities.

TABLE A-1. - Estimated capital costs[1]

	Step 1	Step 2	Step 3
Fixed capital:			
Erected equipment cost	$392,600	$644,900	$107,900
Foundations	51,500	61,300	5,500
Buildings	103,400	77,500	8,300
Structures	15,300	50,600	5,500
Insulation	0	5,500	0
Instrumentation	15,300	1,500	500
Electrical	50,000	47,100	31,000
Piping	46,000	30,600	2,700
Painting	9,200	4,600	500
Miscellaneous	30,700	15,300	5,500
Total direct cost	714,000	938,900	167,400
Field indirect	71,400	93,900	16,700
Total construction cost	785,400	1,032,800	184,100
Engineering	117,800	154,900	27,600
Administration and overhead	39,300	51,600	9,200
Subtotal	942,500	1,239,300	220,900
Contingency	94,300	123,900	22,100
Subtotal	1,036,800	1,363,200	243,000
Contractor's fee	51,800	68,200	12,200
Subtotal	1,088,600	1,431,400	255,200
Plant facilities, 5 pct of above subtotal	54,400	71,600	12,800
Plant utilities, 1 pct of above subtotal	10,900	14,300	2,600
Total plant cost	1,153,900	1,517,300	270,600
Land cost	0	0	0
Subtotal	1,153,900	1,517,300	270,600
Interest during construction period	66,600	87,500	15,700
Fixed capital cost	1,220,500	1,604,800	286,300
Working capital:			
Raw material and supplies	1,000	15,400	5,400
Product and in-process inventory	27,200	27,600	16,200
Accounts receivable	27,200	27,600	16,200
Available cash	15,700	15,700	12,500
Working capital cost	71,100	86,300	50,300
Capitalized startup costs	12,200	16,000	2,900
Subtotal	83,300	102,300	53,200
Total capital cost	1,303,800	1,707,200	339,500

[1]Basis: M and S equipment cost index of 764.7.

The cost for interest on the capital borrowed for construction is included as interest during construction. Land investment is not included in this estimate. Cost for the plant owner's supervision is not included in the capital cost of the proposed plant.

Working capital is defined as the funds in addition to fixed capital, land investment, and startup costs that must be provided to operate the plant. Working capital, also shown in table A-1, is estimated from the following items: (1) Raw material and supplies inventory (cost for raw material and operating supplies for 30 days in steps 1 and 3 and 90 days in step 2), (2) product and in-process inventory (total operating cost for 30 days), and (3) available cash (direct expenses for 30 days).

Capitalized startup costs are estimated as 1 pct of the fixed capital, which is shown in table A-1.

Operating Costs

The estimated operating costs are based on an average of 250 days of operation per year over the life of the plant. The operating costs are divided into direct, indirect, and fixed costs.

Direct costs include raw materials, utilities, direct labor, plant maintenance, payroll overhead, and operating supplies. The direct labor cost is estimated by assigning employees where needed. It has been assumed that no supervision will be required and that these employees will perform routine maintenance tasks. Thus, the only cost component for plant maintenance is the required materials.

Payroll overhead, estimated as 35 pct of direct labor and maintenance labor, includes vacation, sick leave, social security, and fringe benefits. The cost of operating supplies is estimated as 10 pct of the cost of the plant maintenance.

Indirect costs are estimated as 25 pct of the direct labor and maintenance costs. The indirect costs include the expenses of control laboratories, accounting, plant protection and safety, plant administration, marketing, and company overhead. Research and overall company administrative costs outside the plant are not included.

Fixed costs include the cost of taxes (excluding income taxes), insurance, and depreciation. The costs of both taxes and insurance are each estimated as 1 pct of the plant construction cost. Depreciation is based on a straight-line, 20-yr period.

The estimated operating costs for the three steps are shown in table A-2. No credit is included for other product fractions recovered from the bulk scrap in step 1, nor is a credit included for the cement copper coproduct recovered in steps 2 and 3. In the precious metals concentration section (step 2), no credit has been included for the byproduct aluminum hydroxide. It is possible that this material could have a value to the cement industry, but this is not presently known. In addition, no waste disposal costs have been included in either step 2 or step 3. Solid residue must be disposed of, and the waste liquids must be either neutralized or recycled. Inclusion of these costs will increase the operating costs for both steps.

ECONOMIC EVALUATION

In this three-step process, each step produces an upgraded product from which precious metals can be recovered. Based on data currently available, the precious metal losses in the proposed process are equivalent to those that a precious metal refiner would have. Each step reduced the volume of material to be treated by a precious metal refiner. The smaller the volume of material, the lower the shipping and processing charges. After each of the processing steps, the product can either be sold to a refinery to recover the precious metals or used as feed material for the next step. Thus, the cost of each processing step must be considered individually and compared to the cost of sending the feed material for that step directly to a refinery. Table A-3 presents the estimated daily processing cost of each of the three steps.

TABLE A-2. - Estimated operating costs

	Cost per pound precious metal concentrate		
	Step 1	Step 2	Step 3
Direct cost:			
Raw materials:			
Caustic soda, flake at $0.19/lb..........................	$0.000	$0.003	$0.000
Sulfuric acid, 100 pct at $0.03/lb......................	.000	.038	.000
Nitric acid, 42° Bé at $0.10/lb.........................	.000	.000	.550
Hydrochloric acid, 22° Bé at $0.03/lb...................	.000	.000	.136
Soda ash at $0.06/lb....................................	.000	.000	.002
Sodium bisulfite at $0.27/lb............................	.000	.000	.019
Sodium chloride at $0.04/lb.............................	.000	.000	.002
Drums at $40 each.......................................	.000	.791	.316
Total...	.000	.832	1.025
Utilities:			
Electric power at $0.06/kW·h............................	.237	.359	.025
Process water at $0.25/Mgal.............................	.000	.002	.000
Steam, 125 psig at $10/Mlb..............................	.000	.111	.000
Total...	.237	.472	.025
Direct labor: Labor at $9/h..............................	1.184	.591	.889
Plant maintenance: Materials.............................	.993	.770	.110
Payroll overhead, 35 pct of above payroll.................	.414	.207	.330
Operating supplies, 20 pct of plant maintenance...........	.199	.153	.022
Total direct cost.................................	3.027	3.025	2.401
Indirect cost, 40 pct of direct labor and maintenance.......	.871	.544	.400
Fixed cost:			
Taxes, 1 pct of total plant cost........................	.182	.240	.043
Insurance, 1 pct of total plant cost....................	.182	.240	.043
Depreciation, 20-yr life................................	.964	1.268	.226
Total operating cost..............................	5.226	5.317	3.113

TABLE A-3. - Estimated daily processing costs

	Process step		
	1	2	3
Operating cost of processing step.....	$1,322	$1,346	$787
Cost to complete processing by refiner[1].	1,945	410	32
Total processing cost..............	3,267	1,756	819
Byproduct credit.....	0	360	14
Net processing cost.............	3,267	1,396	805

[1]Quoted refinery charges to recover pure metals, including shipping charges.

The net daily processing cost is composed of three items. The first is the operating cost of the processing step as described in this report. The second item is the cost charged by a refiner to recover pure metals from the product of each step, because none of these steps produce pure metal products in ingot form. Refinery treatment charges for each of the feed and product materials were obtained from a precious metal refiner. For the bulk electronic scrap (feed to step 1) the treatment charge is $0.90/lb. A charge of $1.50/lb is used for the high-tension separator scrap, the precious metal concentrate, and the gold and silver (products from steps 1,

2, and 3, respectively). In addition, a shipping charge of $241/ton is added to determine the total refinery processing charges (6). This charge is for a 250-mile shipping distance. The third item is a credit for byproducts recovered in steps 2 and 3. (Cement copper and un-reacted metallic scrap from copper cementation, both containing significant quantities of precious metals, are recovered in step 2. Cement copper is recovered in step 3.) The copper coproduct is valued at $0.30/lb, gold is valued at $350/tr oz, and silver is valued at $8.50/tr oz. Therefore, the net daily cost of processing for each step is the daily cost of the processing step, plus the additional refinery processing charges, minus the byproduct credit.

To determine the economic merits of the proposed processing step, the net processing cost for each step must be compared to the cost of recovering the precious metals, without processing, from the feed material by a refiner. Table A-4 shows the comparison. Since steps 1 and 2 show a profit, it can be concluded that it is less costly to process the scrap by these steps rather than sending the material directly to a precious metal refiner. Step 3 operates at a loss; therefore, including this step would not be worthwhile.

One more factor to be considered is to determine if the return on investment is sufficient to justify the investment. Steps 1 and 2 require investment of about $1.3 million and $1.7 million, respectively. In order for this investment to be worthwhile, each step must have an individual return on investment after taxes of 15 pct, which is considered the minimum acceptable for a new venture. Table A-5 shows the profit required for a 15-pct return on investment for steps 1 and 2 as compared to the estimated profit shown in table A-4. Step 3 is not shown because it operates at a loss. Table A-6 shows the estimated return on investment for each processing step.

TABLE A-4. - Comparison of daily costs of proposed processing steps and refinery processing charges

	Process step		
	1	2	3
Refiner's processing charge[1]	$12,246	$1,945	$410
Net processing cost[2]	3,267	1,396	805
Estimated profit per step	8,979	549	-395

[1]Quoted refinery charges to process feed material, including shipping charges.
[2]From table A-3.

TABLE A-5. - Comparison of estimated daily profit with profit required for a 15-pct return on investment

Processing step	Estimated profit[1]	Profit required for a 15-pct return on investment
1	$8,979	$3,127
2	549	2,968

[1]From table A-4.

TABLE A-6. - Estimated return on investment after taxes, percent

Processing step

1	73
2	6
3	<0

For step 1, the estimated profit is greater than that required for a 15-pct return on investment, and for step 2 the estimated profit is much less. Therefore, only step 1 yields sufficient return on investment to be considered a worthwhile investment.

A potential problem in processing this type of heterogeneous scrap is obtaining a reliable analysis of the precious metal content. Since the product from step 2 is more homogeneous than the product from step 1, a more reliable precious metal assay can be determined from the step 2

product. If the precious metal assay of the scrap is reliable, constraints (i.e., minimum precious metal recoveries) can be placed on a refiner so that the maximum value is obtained for the precious metals. These constraints generally cannot be imposed if the precious metal content of the scrap is not known. Therefore, performing step 2 may increase the revenue returned from a precious metal refiner. A similar case can be made for performing step 3. Because no data are available to quantify the additional value realized by increasing product homogeneity, it cannot be considered in this study.

TECHNICAL EVALUATION

Modifications to the proposed process have the potential to reduce costs. At present, modifications to the second step of the proposed process are the only ones that should be considered. Costs of the third step cannot be considered until the second step is shown to be viable.

There are two high-cost operations in the second step of the proposed process: aluminum hydroxide precipitation and H_2SO_4 regeneration.

The aging time of 10 to 12 weeks for the aluminum hydroxide precipitation requires a large number of tanks for storing the solution, and these are a significant portion of the capital cost. In normal Bayer process operation, this same type of precipitation takes 36 to 48 h using recycled seed slurry. If the precipitation time for the proposed process

can be reduced to the length of time used in the Bayer process, the cost of the tanks will be reduced significantly. This would reduce the depreciation charges in the operating cost by roughly 15 pct, or $0.20/lb of precious metal concentrate.

H_2SO_4 regeneration by the electrolytic procedure described in this report is also a costly item. It accounts for about 17 pct or $0.90 of the total operating cost. The electrolytic cell is a high-capital-cost item, and it consumes most of the electrical energy required in this step. Instead of regenerating the H_2SO_4, the solution could be neutralized with lime and discarded, and fresh acid could be used for each batch. Neutralizing and disposing of the H_2SO_4 would cost about $0.50/lb of precious metal concentrate, which would be less expensive than electrolytic regeneration by about $0.40/lb of precious metal concentrate.

Consequently, if both these modifications are employed, the total cost savings would be about $0.60/lb of precious metal concentrate, or 11 pct of the total operating cost for this step. However, even with this savings, the reduction in the operating cost would not be sufficient to yield a 15-pct return on investment after taxes. Any other process modifications that could be considered represent minor savings compared to the two detailed above. Thus, even with cost-saving modifications, the estimated profit on the proposed process will not produce a 15-pct return on investment after taxes.