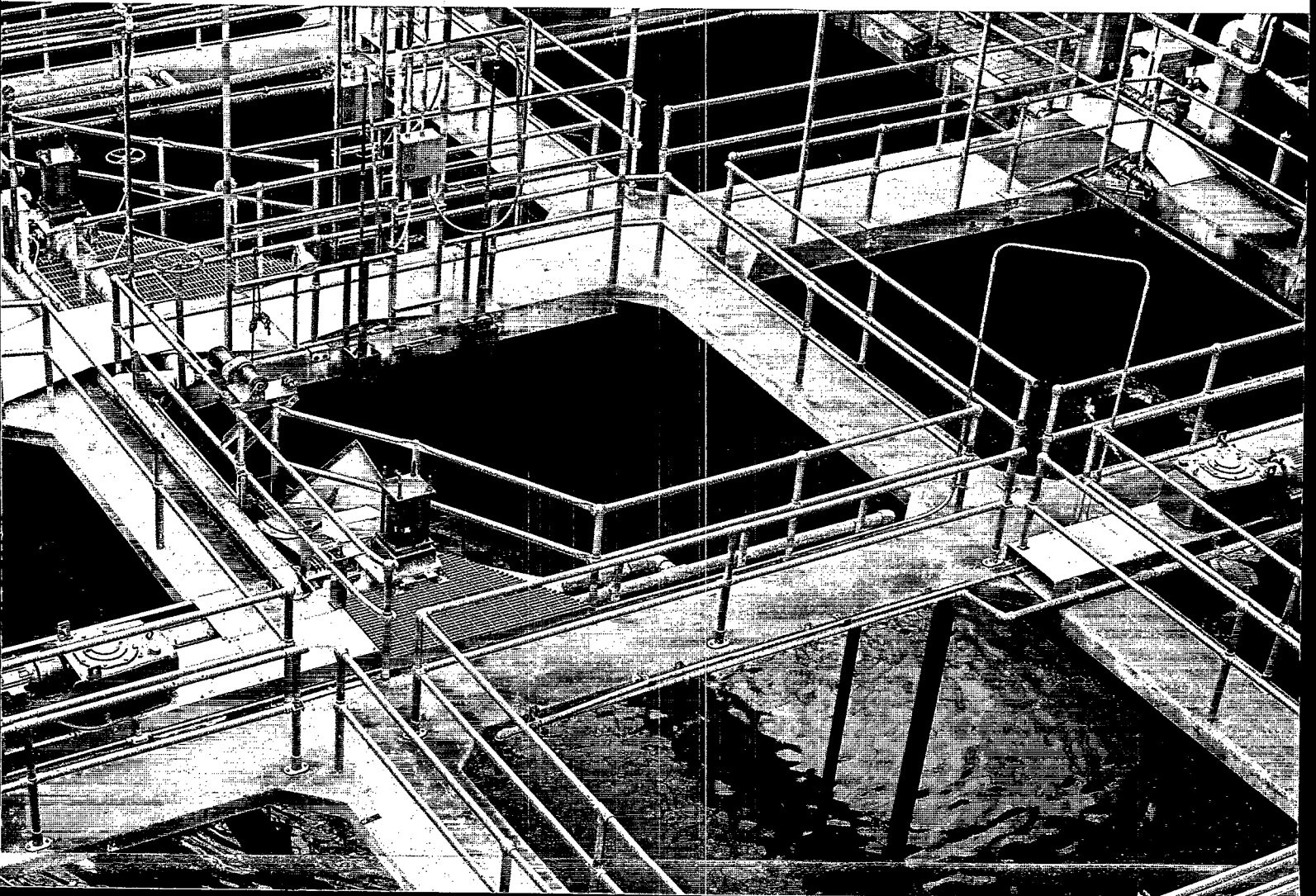




Environmental Pollution Control Alternatives

Reducing Water Pollution Control Costs in the Electroplating Industry



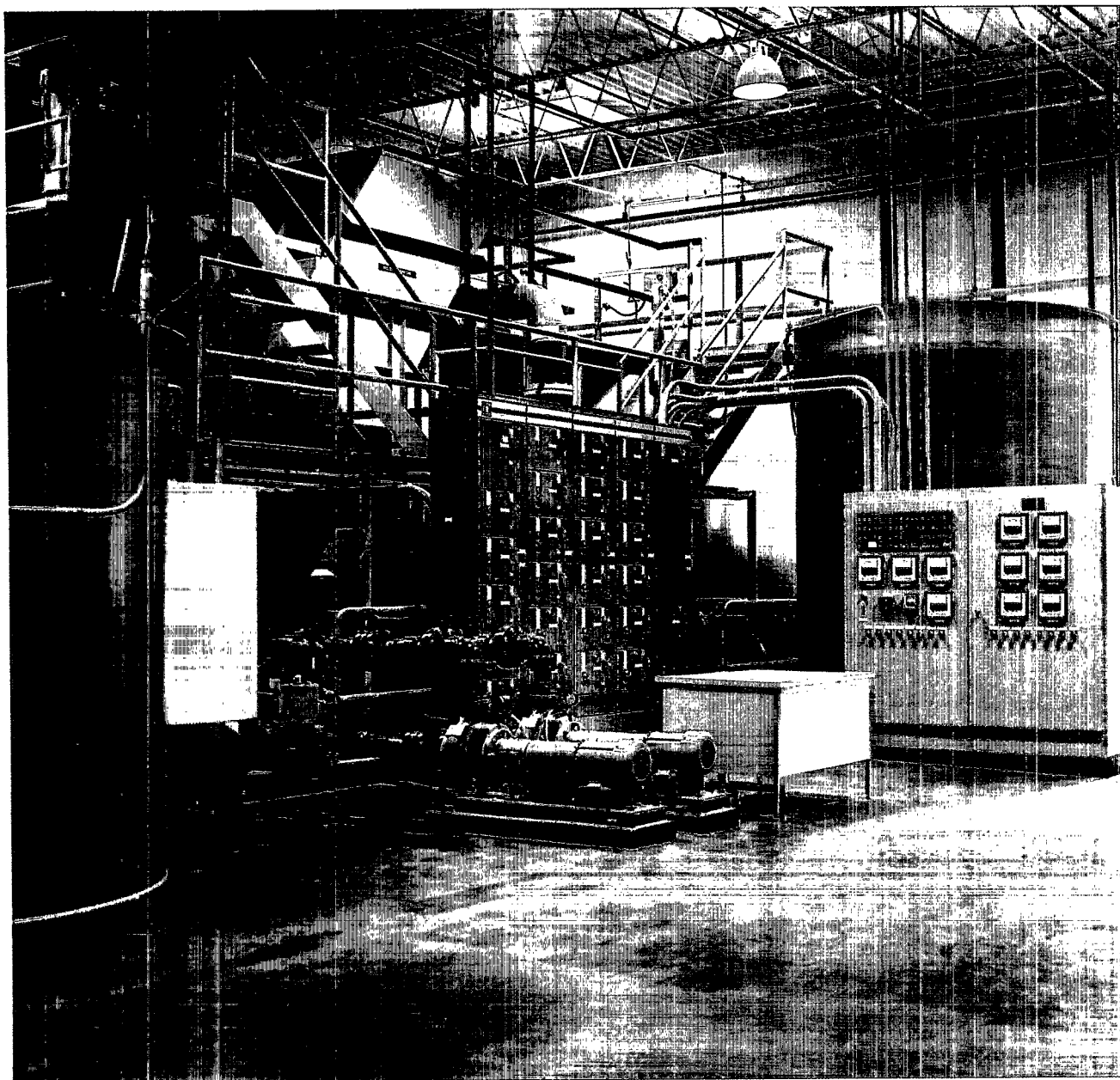
Cover Photograph:
Exterior batch treatment tanks.

Environmental Pollution Control Alternatives

Reducing Water Pollution Control Costs in the Electroplating Industry

September 1985

This report was prepared jointly by
Industrial Technology Division
Office of Water Regulations and Standards
Office of Water
Washington, DC 20460
and
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Office of Research Program Management
Office of Research and Development
Cincinnati, OH 45268



From left to right, a lime slurry tank, flocculator feed tank, lamella solids separator, control panel, and sludge storage tank for a 300 gal/min facility.

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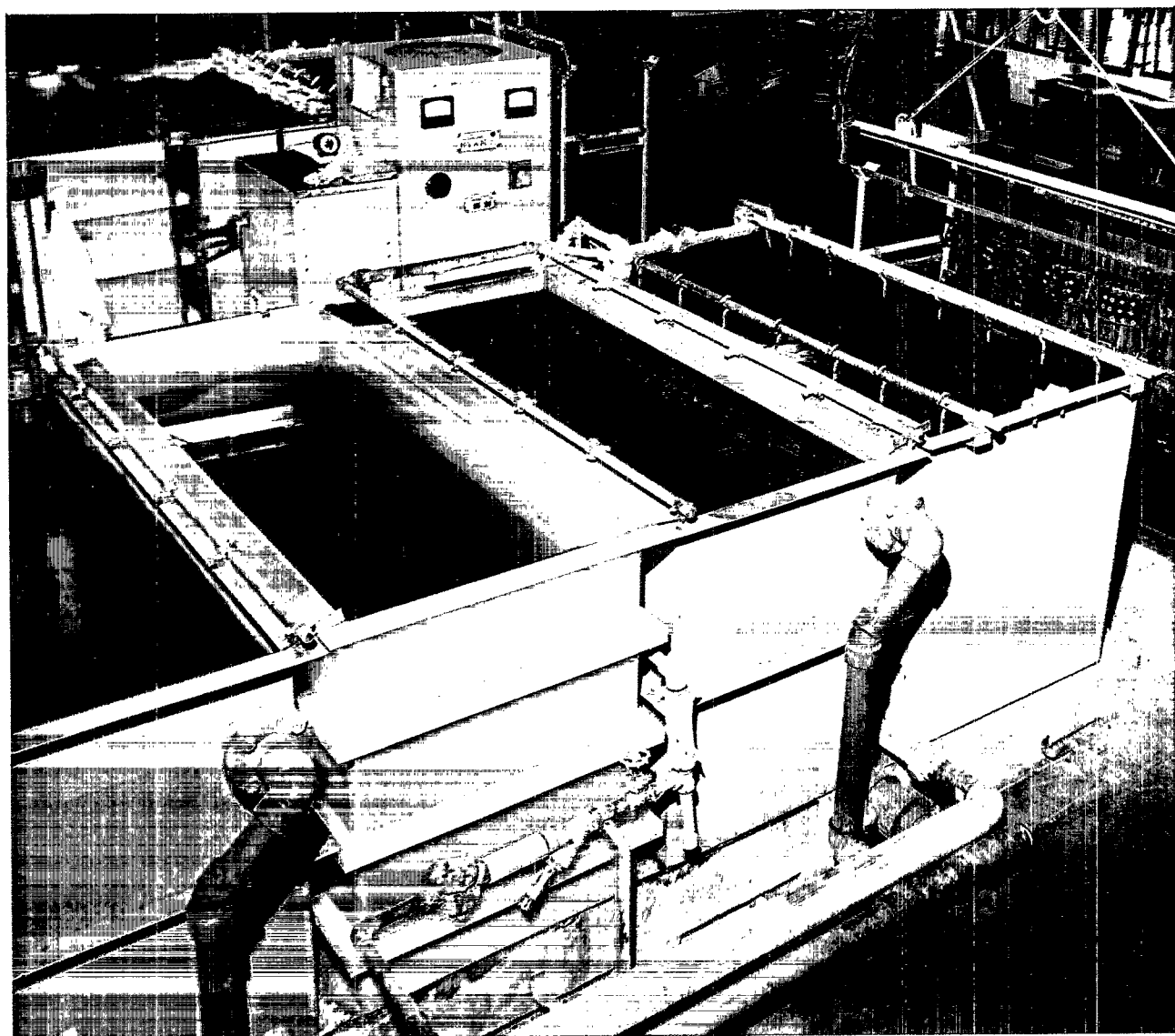
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Bath and rinse tanks for rack electroplating operation.

1. Overview

Under the Federal Water Pollution Control Act Amendments of 1972, the Environmental Protection Agency (EPA) was directed to issue effluent limitations and performance standards for industrial dischargers to waterways or publicly owned treatment works (POTWs). Subsequent legislation, the Clean Water Act Amendments of 1977, required that EPA adhere to a new schedule for promulgating those limitations and standards, particularly with regard to a list of toxic pollutants referred to as "priority" pollutants. A number of these pollutants are associated with the electroplating industry. Regulations under this Act appear as the Effluent Guidelines and Standards for Metal Finishing, promulgated July 15, 1983 in 48 Federal Register 32485, and in subsequent amendments and corrections. These are incorporated into Title 40, Part 433 of the Code of Federal Regulations. These regulations apply to all new metal finishing and electroplating facilities and all captive and integrated existing facilities. Regulations for existing electroplating job shops that discharge to POTWs appear in Part 413 of Title 40 of the Code of Federal Regulations.

In addition to these effluent-related requirements, electroplaters are also affected by the requirements of the Resource Conservation and Recovery Act of 1976 (RCRA). Under RCRA and the subsequent amendments of 1984, EPA has promulgated and will continue to promulgate regulations controlling the management and disposal of industrial sludges, particularly those containing hazardous materials.

Operating costs in the electroplating industry have been affected by the economic burden of meeting these pollution control requirements, as well as by steadily rising raw material and utility costs. Table 1 illustrates the increase in unit prices over a 12-year period for electroplating chemicals, wastewater treatment chemicals, and utilities used in the industry. Three significant features of the electroplating chemicals listed in the Table are worthy of mention:

- Cyanide, chromium, and cadmium compounds are classified as toxic substances on the EPA list of priority pollutants.
- Almost all the chemicals listed are typically present in the wastewater from electroplating processes.
- The price of each has risen dramatically since 1972.

As indicated in Table 1, prices of wastewater treatment chemicals and utility costs, including water and sewer fees, have also risen sharply. These rising costs of materials and services, as well as increasingly stringent EPA regulations, suggest that all of these costs, and particularly the costs of water pollution control treatment for the electroplating industry, need to be reevaluated.

A review of the costs of conventional treatment systems for electroplating wastewater quickly comes to focus on two factors:

- The volume of wastewater passing through the system
- The concentration of pollutants in the wastewater.

The capital and operating costs are both directly dependent on these factors. Consequently, costs of complying with EPA regulations can be reduced by using technologies that minimize the volume of wastewater generated, thereby reducing the quantity of treatment chemicals used as well as reducing a firm's water, sewer, and sludge removal expenses. Costs can be further reduced by using technologies which allow for the recovery and reuse of valuable plating materials and process chemicals.

This publication discusses the cost tradeoffs of wastewater reduction and materials recovery technologies for the electroplating industry in the context of the EPA regulations. Although it discusses sludge briefly, its primary focus is on wastewater treatment. It is designed for those who will be selecting an optimum control system for their operation. Chapter 2 presents a description and capital costs for each of six components of a conventional treatment system. The chapter also includes information on determining the operating costs for the system. Chapter 3 presents process modifications technologies which offer potential for cost savings by minimizing water use or by reducing drag-out loss. These include several materials recovery processes. For each, examples or worksheets are provided to aid the user in estimating costs or in making cost/benefit assessments of treatment technologies.

This publication is an update of a 1979 EPA publication: *Environmental Pollution Control Alternatives: Economics of Wastewater Treatment Alternatives for the Electroplating Industry*; EPA publication number 625/5-79-016. It has been revised to reflect changes in technologies and prices and in the EPA regulations that directly affect costs in the electroplating industry.

A companion document, *Environmental Regulations and Technology: The Electroplating Industry*, EPA publication number 625/10-85-001a, gives detailed information on water and waste regulations, case histories, an overview of control technologies for both water and solid waste, financial information, and sources for further information.

Table 1.

Prices of Electroplating Chemicals, Wastewater Treatment Chemicals, and Utilities Used by the Electroplating Industry

	Price		
	1972	1978	1984
Electroplating chemicals (\$/lb):^{a,b}			
Boric acid (H_3BO_3)	0.069	0.176	0.30
Cadmium chloride ($CdCl_2$) ^c	—	2.60	3.73
Chromic acid (H_2CrO_4)	0.37	0.78	1.18
Copper cyanide ($Cu(CN)_2$) ^c	1.05	1.95	2.62
Copper sulfate ($CuSO_4$) ^c	0.47	0.88	0.88
Nickel chloride ($NiCl_2$) ^c	0.67	1.04	1.05
Nickel sulfate ($NiSO_4$) ^c	0.50	0.76	1.19
Sodium cyanide ($NaCN$) ^c	0.21	0.40	0.68
Zinc (metallic) ^c	0.18	0.31	0.51
Zinc cyanide ($Zn(CN)_2$) ^c	0.64	1.41	2.00
Wastewater treatment chemicals (\$/lb):^{a,b}			
Calcium hydroxide ($Ca(OH)_2$)	0.010	0.017	0.025
Calcium oxide (CaO , quicklime)	0.009	0.016	0.02
Chlorine (Cl_2)	0.038	0.075	0.09
Ferrous sulfide (FeS)	—	0.40	0.50
Hydrochloric acid (28% HCl)	0.0135	0.023	0.35
Sodium bisulfite ($NaHSO_3$)	0.066	0.13	0.30
Sodium carbonate (58% Na_2CO_3)	0.018	0.03	0.06
Sodium hydroxide (98% $NaOH$ equiv.)	0.036	0.08	0.13
Sodium hypochlorite ($NaOCl$)	—	0.40	0.60
Sodium sulfide (Na_2S)	0.07	0.12	0.22
Sulfur dioxide (SO_2)	0.038	0.085	0.14
Sulfuric acid (H_2SO_4)	0.017	0.023	0.05
Utilities:^d			
Electricity (\$/kWhr)	0.028	0.045	0.08
Steam by energy source (\$/MMBtu):			
Natural gas	1.03	2.07	6.00
Oil	1.39	3.53	7.50
Water (\$/1,000 gal):			
Use fee	0.25	0.50	0.80
Sewer fee ^e	0.25	0.60	1.20

^aPrices are for bulk shipments of chemicals; prices for smaller quantities or specially packaged quantities may be 10% to 50% higher. Plating chemicals purchased with proprietary additives are from 20% to 40% higher.

^bPrices from *Chemical Marketing Reporter*, Oct. 30, 1972; Feb. 20, 1978; and Feb. 6, 1984.

^cSubstance is on EPA list of priority pollutants.

^dAverage prices.

^eTypical of a metropolitan area.

2. Costs of Conventional Wastewater Treatment Systems

Certain wastewater treatment technologies have become so widely accepted in the electroplating industry that they are usually referred to as "conventional" treatment technologies. These include chromium reduction and cyanide oxidation (as needed), neutralization to precipitate suspended solids followed by clarification of the wastewater, and thickening and filtration of the resulting sludge. Figure 1 illustrates a conventional treatment system for wastes containing, among other pollutants, chromium and cyanides.

Of course, the size and complexity of a particular system can vary significantly. Wastewater flow rate is a major factor in determining the initial cost of equipment, while both flow rate and pollutant loading are significant in determining the operating cost of the system. System design is also affected by wide variations in pollutant loading often found in electroplating waste streams.

Table 2 lists the variability in wastewater characteristics found by EPA in a survey of the electroplating industry. While no single waste stream is expected to experience such variability in all of its components, it is essential that the variations be understood, and that the waste treatment system be sized to handle variations that cannot be eliminated. In addition, complex plating systems or unusual wastewater characteristics often mean that laboratory- or pilot-scale tests must be conducted to determine whether a proposed system will bring a waste stream into compliance with regulations. (Guidance on the design and sizing of wastewater treatment systems is available in several publications listed in the Bibliography.)

NOTE: All costs in this report relate to 1984.

This chapter discusses the factors needed to estimate capital and operating costs of the conventional wastewater treatment systems. Capital costs are described for each of six components of a wastewater treatment system:

- Wastewater collection
- Chromium reduction
- Cyanide oxidation
- Neutralization/precipitation
- Clarification
- Sludge handling.

Operating costs are described for sludge disposal, municipal wastewater treatment, and wastewater treatment chemicals. The chapter concludes with an example that illustrates how total costs can be estimated for a specific facility.

Table 2.

Composition of Raw Waste Streams from Common Metals Plating

	Range (mg/l)
Copper	0.032 - 272.5
Nickel	0.019 - 2,954
Chromium:	
Total	0.088 - 525.9
Hexavalent	0.005 - 334.5
Zinc	0.112 - 252.0
Cyanide:	
Total	0.005 - 150.0
Amenable to chlorination	0.003 - 130.0
Fluoride	0.022 - 141.7
Cadmium	0.007 - 21.60
Lead	0.663 - 25.39
Iron	0.410 - 1.482
Tin	0.060 - 103.4
Phosphorus	0.020 - 144.0
Total suspended solids	0.100 - 9,970

SOURCE: U.S. Environmental Protection Agency, *Development Document for Proposed Existing Source Pretreatment Standards for the Electroplating Point Source Category*, EPA 440/1-78-085, Feb. 1978.

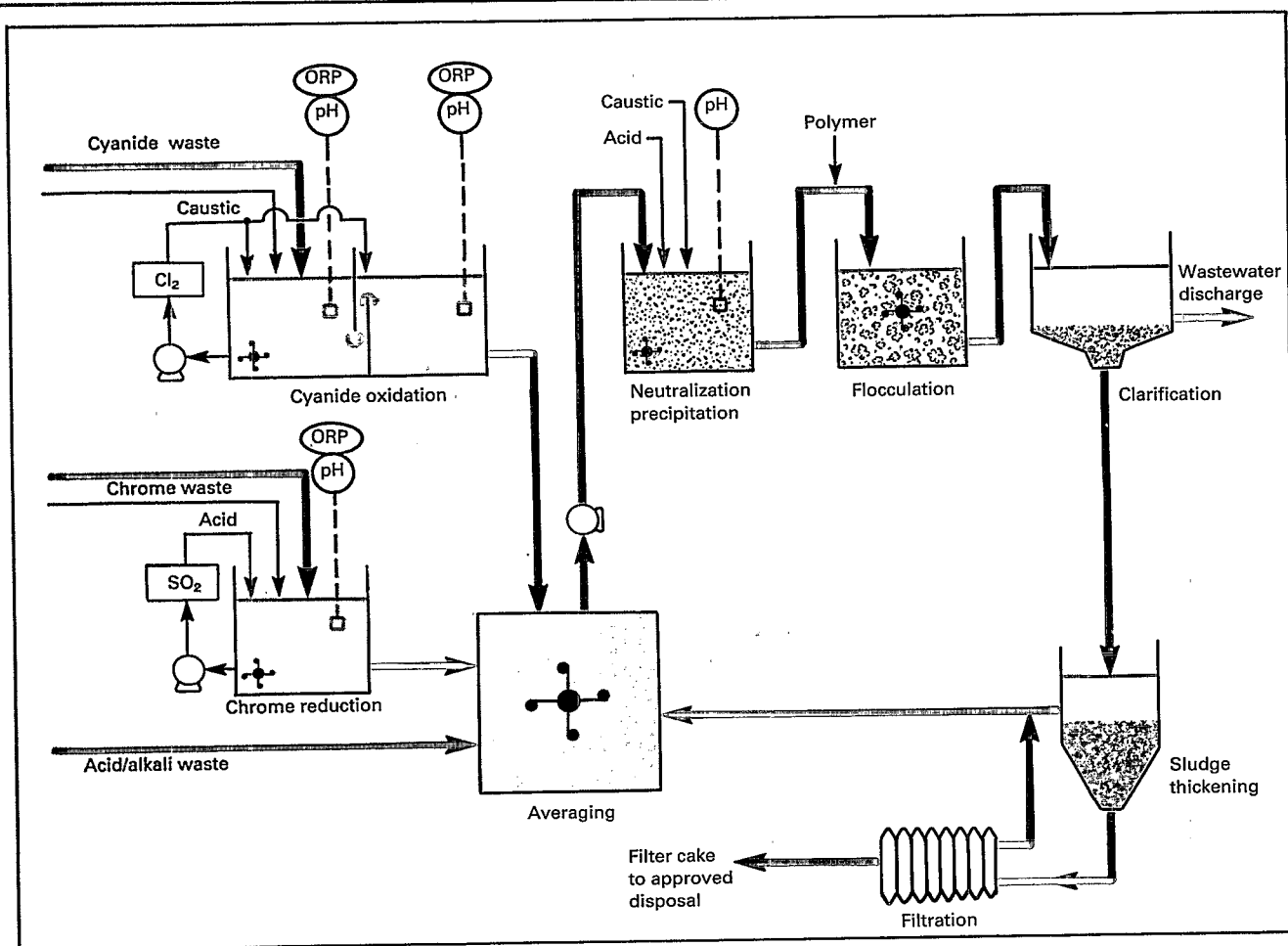


Figure 1. Conventional Wastewater Treatment System for Electroplating

Capital Costs

The unit processes shown in Figure 1 are used extensively in the electroplating industry, and as a result their design has become somewhat standard. The ability to standardize has reduced the high cost of site preparation and construction through the development of skidmounted package systems, complete with all hardware and auxiliaries. Installation costs for package systems usually range between 10 and 30 percent of the purchase price of the equipment compared with installation costs of 50 to 100 percent of purchase price for component systems.

The costs presented in the following sections assume the purchase of all components of the individual systems. Costs could be reduced by using existing pumps, tanks, and instrumentation, but these are rarely available. Installation costs will increase as a result of site-specific requirements related to wastewater collection systems, new building space, structural modifications, or the relocation of existing equipment.

Wastewater Collection. Wastewater from individual plating operations must be directed to the appropriate treatment components, which often entails separate wastewater collection systems for chromium, cyanide, and acid/alkali wastes. Typically, a collection pipe system is installed parallel to the plating line so that spent rinse waters can be drained to one or more treatment sumps. Equipment washdown and spilled materials are collected in a trench which also leads to the appropriate sump. The treatment sump is equipped with a level-

controlled pump that delivers the wastewater to the treatment unit. In addition, some provision must be made for collecting the strong cleaning and plating chemical solutions which are usually bled at a slow, constant rate into the treatment system by way of an averaging tank to avoid upsetting the process.

Wastewater collection costs are highly variable. A plant may have to provide as many as five separate collection/sump systems to handle three different kinds of wastewater plus spent cyanide and noncyanide chemical solutions. Cost components for wastewater collection are given in Table 3. Larger sumps are used for chromium and cyanide wastewater collection; in fact, the sump itself may be used as the treatment vessel (this practice is rare). It should be pointed out that custom-made concrete sumps are considerably more expensive to construct than the preformed sumps indicated in Table 3.

Table 3.

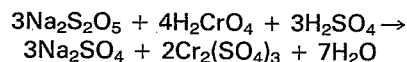
Costs for Wastewater Collection

Collection conduit: ^a	
Linear runs:	4" - \$1.50/ft 6" - \$2.30/ft
Each connection:	4" - \$200.00 6" - \$350.00
Each bend:	4" - \$200.00 6" - \$300.00
Collection sumps: ^b	
100 gallons ...	\$2,000
300 gallons ...	\$2,500
500 gallons ...	\$3,000
1,000 gallons ...	\$3,500

^aPVC pipe.

^bPreformed PVC inground tank with sump pump, level control and steel grating cover. 1984 data.

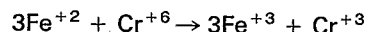
Chromium Reduction. Chromium is usually present in electroplating wastewater as trivalent chromium (Cr^{+3}) or as hexavalent chromium (Cr^{+6}). Although most heavy metals are precipitated readily as insoluble hydroxides in the neutralizer, hexavalent chromium must first be reduced to trivalent chromium. Reduction is usually achieved by reaction with gaseous sulfur dioxide (SO_2) or a bisulfite solution (NaHSO_3). Using sodium metabisulfite ($\text{Na}_2\text{S}_2\text{O}_5$), the net reaction involves chromic acid and sulfuric acid in the formation of sodium sulfate, chromium sulfate, and water:



Because the reaction proceeds rapidly at low pH, an acid is added to maintain the chromic acid wastewater between a pH of 2 and 3. To prevent the release of sulfur dioxide during the reaction, maintaining the pH near 3 is advised.

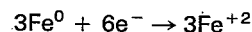
In addition to the use of sulfur dioxide or bisulfites to reduce hexavalent chromium, there are three commercially available methods that use the reducing potential of iron and/or iron salts.

Ferrous sulfate reduces hexavalent chromium according to the following equation:

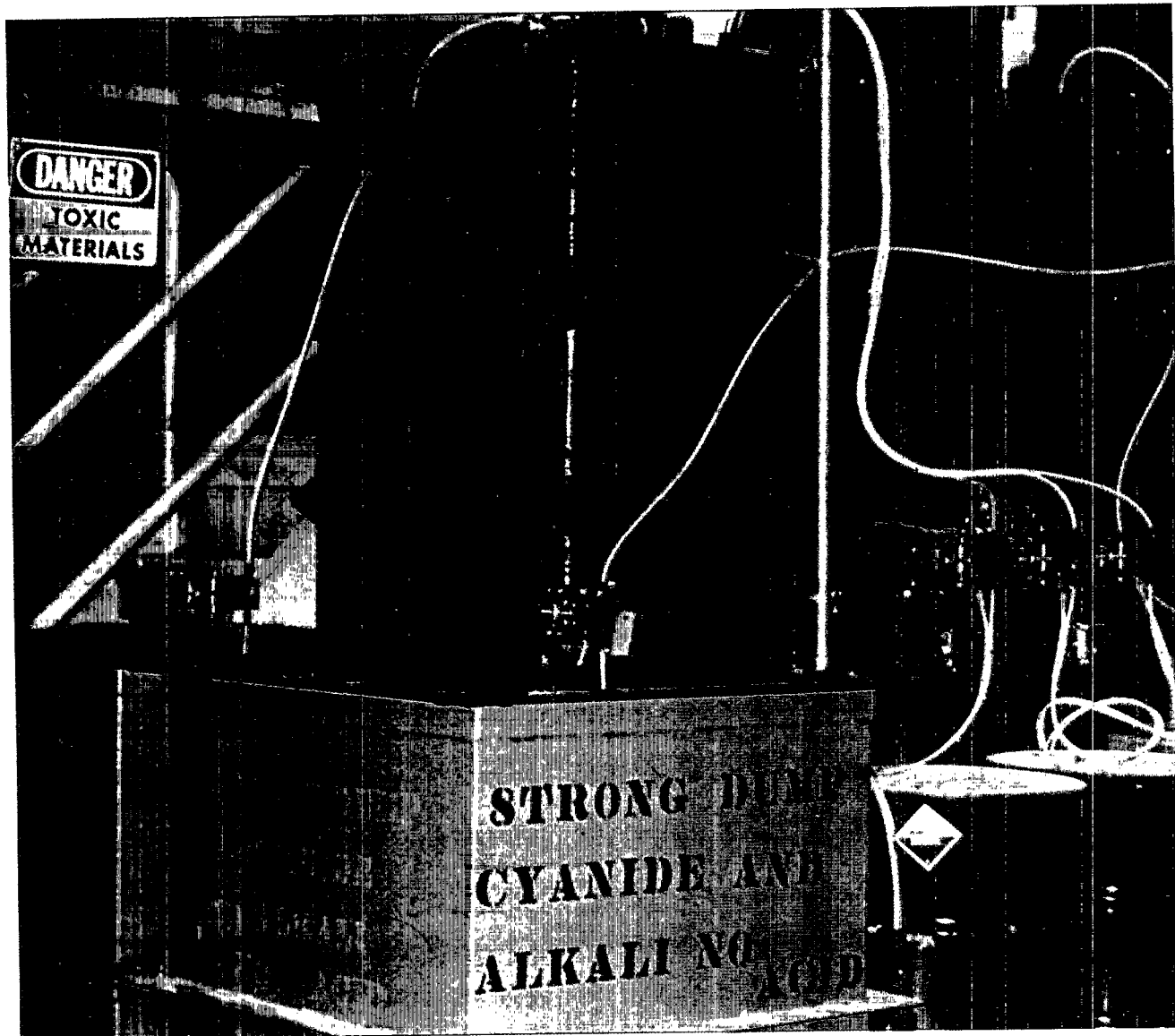


The use of iron in this form adds considerably to the volume of sludge produced by the treatment system. This method is rarely used except where an effluent of high chromium content is located close to a cheap and abundant source of ferrous salts, such as waste pickle liquor from a steel mill.

Ferrous salts may also be generated by a patented process employing an electrolytic cell having iron anodes and inert cathodes:



As the iron anodes are dissolved in the process, they must be replaced with new ones. While this process will also produce substantial quantities of excess sludge, it provides a convenient method of continuously treating low concentrations of hexavalent chromium, such as might be found in cooling tower blow-down waters or in rinses from chromate conversion coatings. With higher chromium concentrations it is difficult to economically produce the required ferrous ion concentrations on a flow-through basis.



Concentrated bath dump tank with oxidation and neutralization/flocculation units.

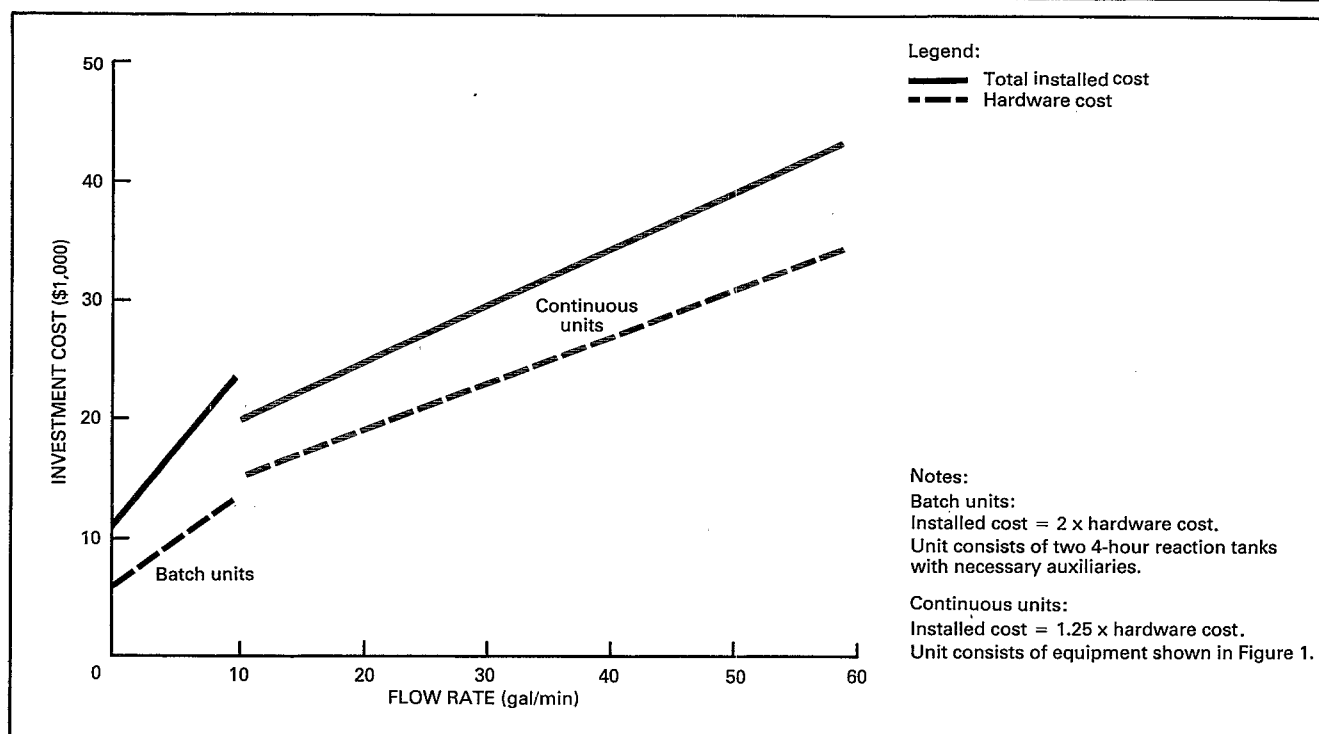
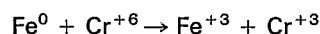


Figure 2. Investment Cost for Chromium Reduction Units

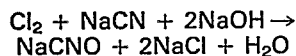
Iron, in the form of small pieces of scrap steel, may also be used to directly reduce hexavalent chromium. This patented process is controlled by adjusting the pH of the influent according to its hexavalent chromium content. A pH of 2.0 to 2.2 will usually accommodate chromium loadings as high as 200 to 300 ppm. Lower pH settings would be required to treat higher chromium concentrations. The following equation represents the reduction mechanism:



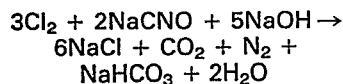
Because the iron and chromium react on a one-to-one basis, rather than a three-to-one ratio, the sludge produced by this process is significantly less than that generated in the other processes using iron. The low cost of scrap steel allows this process to compete favorably with the cost of sulfur dioxide treatment, while eliminating the potential of releasing hazardous gas.

Hardware and installation costs from packaged continuous and batch chromium reduction units are shown in Figure 2. Costs include storage and feed systems for the treatment reagents, as well as the costs for hardware, piping, instrumentation, and utility connections. The continuous unit costs are based on using sodium metabisulfite as the reducing agent and a wastewater retention volume of 30 min. For very small flows, simpler and less costly batch systems are feasible. The batch system costs include two reaction tanks each sized to hold 4 h of wastewater flow and equipped with high level alarms, portable pH and oxidation reduction potential (ORP) meters, a portable mixer, and storage tanks and feed pumps to add sodium metabisulfite and sulfuric acid to the reaction tanks.

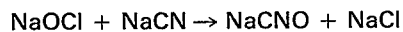
Cyanide Oxidation. Dilute cyanide rinse streams resulting from plating operations and cyanide dips must also be treated separately to oxidize the highly toxic cyanide, first to less toxic cyanate, then to harmless bicarbonates and nitrogen. The oxidizing reagent is usually chlorine gas (Cl_2) or sodium hypochlorite (NaOCl). Using chlorine, the typical reaction in the first stage involves sodium cyanide (NaCN) and sodium hydroxide (NaOH):



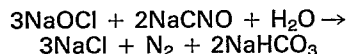
and in the second stage:



When sodium hypochlorite is used, the typical reaction in the first stage is:



and in the second stage,



In most continuous systems, it is preferable to conduct the operations in two series-connected reaction tanks rather than in stages in one tank. In the first stage, the pH is adjusted between 9 and 11 using an alkali such as caustic soda or lime. The pH in the second reaction chamber is controlled to approximately 8.5. Sodium hypochlorite is added continuously to both stages. Demand in each stage is determined by measuring ORP. The reaction time needed is approximately 30 to 60 min in each stage.

Figure 3 shows cost curves for continuous and manual batch cyanide oxidation units. The continuous unit cost is for a unit which uses sodium hypochlorite as the oxidizing agent. The cost includes storage and feed systems for the treatment reagents. The batch system cost is for a system with two 4 h reaction tanks and the auxiliaries required to add sodium hypochlorite and control pH.

Again, for very small flows, simpler and less costly batch systems are feasible. At wastewater flow rates below 20 gal/min (76 l/min) batch units appear to be more cost-effective than continuous units.

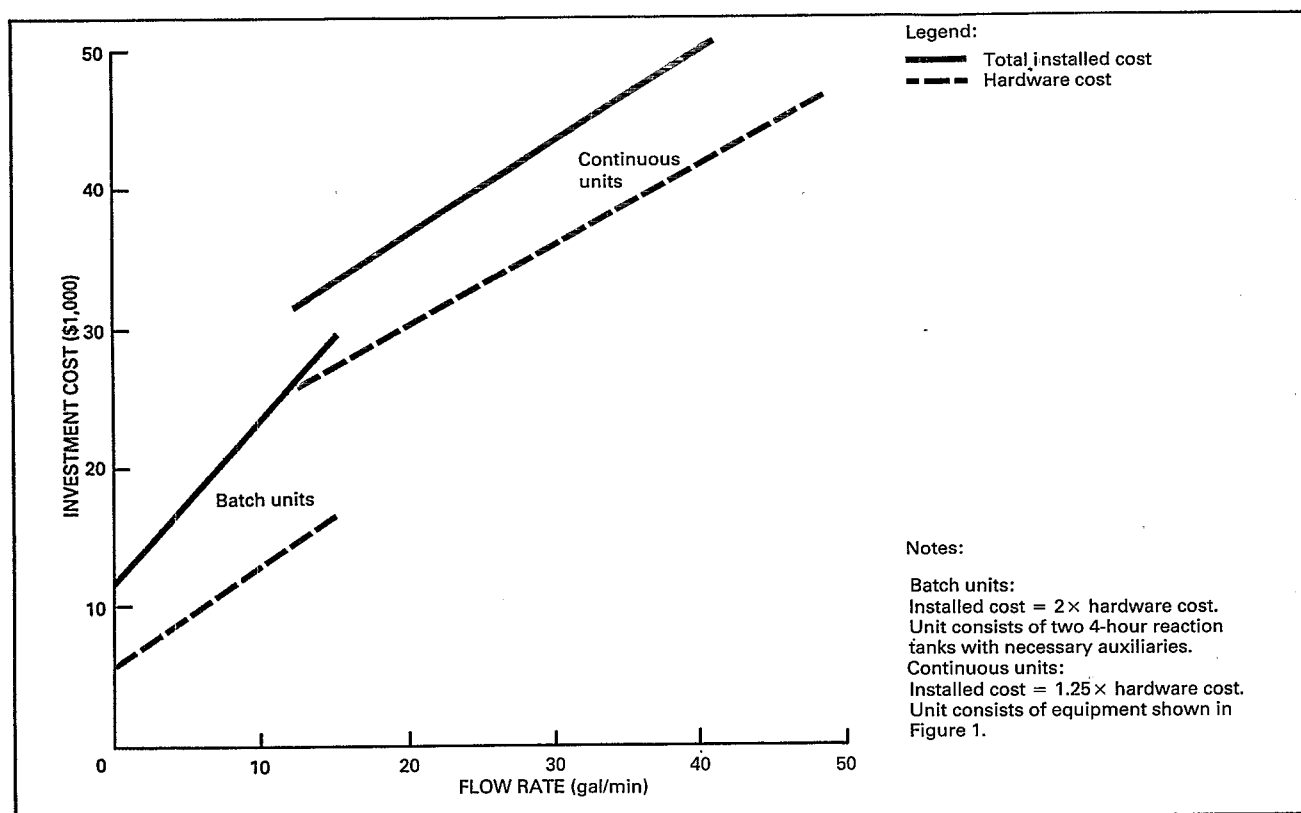


Figure 3. Investment Cost for Cyanide Oxidation Units

Neutralization/Precipitation. Waste streams from each of the various metal cleaning and plating operations are combined in the neutralizer with the effluent from the chromium reduction and cyanide oxidation units. Because the heavy metals are only soluble under acidic conditions, the pH is adjusted to a range of 9.0 to 9.5, at which the metals precipitate as hydroxides. After the pH is adjusted, a small amount (1 to 5 mg/l) of a coagulating agent such as aluminum sulfate, ferrous sulfate, or calcium chloride is added to the wastewater to facilitate precipitation.

Many types of neutralization systems can be designed with various degrees of automation and controls, depending on the magnitude and variability of the flow and its pH. Because a change of 1 pH unit represents a tenfold change in hydrogen ion concentration and a hundredfold change in solubility for some metals, maintaining the pH in the narrow range where maximum removal of pollutants is realized is critical, though difficult, especially when the neutralizer feed is subject to wide variation.

A single-stage, continuous neutralizer, in which all the alkali such as lime or caustic soda is fed into a single reaction vessel, is suitable for most electroplating applications. If the wastewater is subject to rapid changes in flow rate or pH, however, a multistage neutralizer is required. In the multistage units, most of the alkali is added in the first vessel to increase the pH to 6. Final pH adjustments of the wastewater are made in the remaining reaction vessels to promote precipitation and to enhance the settling characteristics of the metal hydroxides. To maintain adequate pH control, the retention time for typical neutralization is 15 to 30 min. The efficient use of lime requires a minimum of 30 min because lime reacts more slowly than sodium hydroxide.

Figure 4 shows hardware and total installed costs for a continuous neutralization unit typically used in the electroplating industry. This unit is single stage with pH-controlled addition of caustic soda and sulfuric acid.

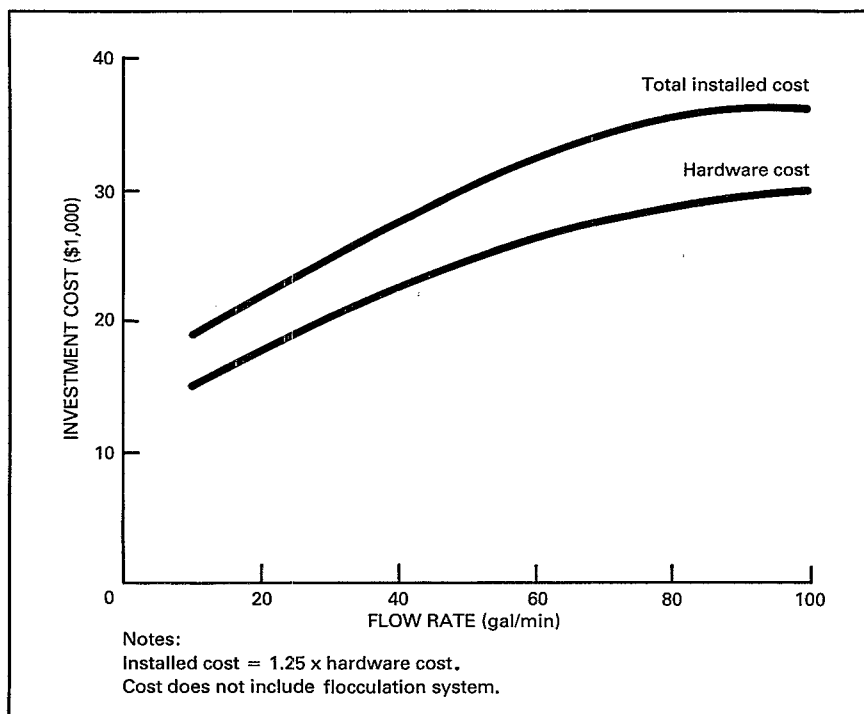


Figure 4. Investment Cost for Continuous Single-Stage Neutralization/Precipitation Unit

Clarification. Metal hydroxides and other insoluble pollutants are removed from the wastewater by gravity settling and/or filtration. The removal efficiency depends on the settling rate of the suspended solids in the wastewater feed. Typically, some of these solids settle very slowly because of their small size and their slight density difference compared with the water. Because an economical design of the clarifier limits the retention time in the settling chamber, some level of suspended solids will appear in the overflow.

To enhance the settling characteristics of the suspended solids, a flocculating agent, usually an organic polymer, is added in a flocculator where the wastewater is agitated gently to allow the solids to agglomerate. The wastewater then enters the clarifier where the solids settle out. The solids in the underflow can be discharged to a holding tank for subsequent dewatering.

The optimal doses of flocculating agents, hardware specifications, and solids removal efficiencies are usually estimated based on laboratory tests conducted by the equipment vendor. Figure 5 shows hardware and total installed costs for flocculation and clarification units typical of those used in the industry. These costs are presented as a function of volumetric flow rate; they also reflect the effect on unit costs of solids-settling rates and the level of solids allowed in the effluent. The units are assumed to have a separate flocculation tank, a polymer feed system, a "lamella" (or slant-tube separator), and a zone in which sludge collects before being discharged. These units are more widely used than simpler rectangular settling chambers.

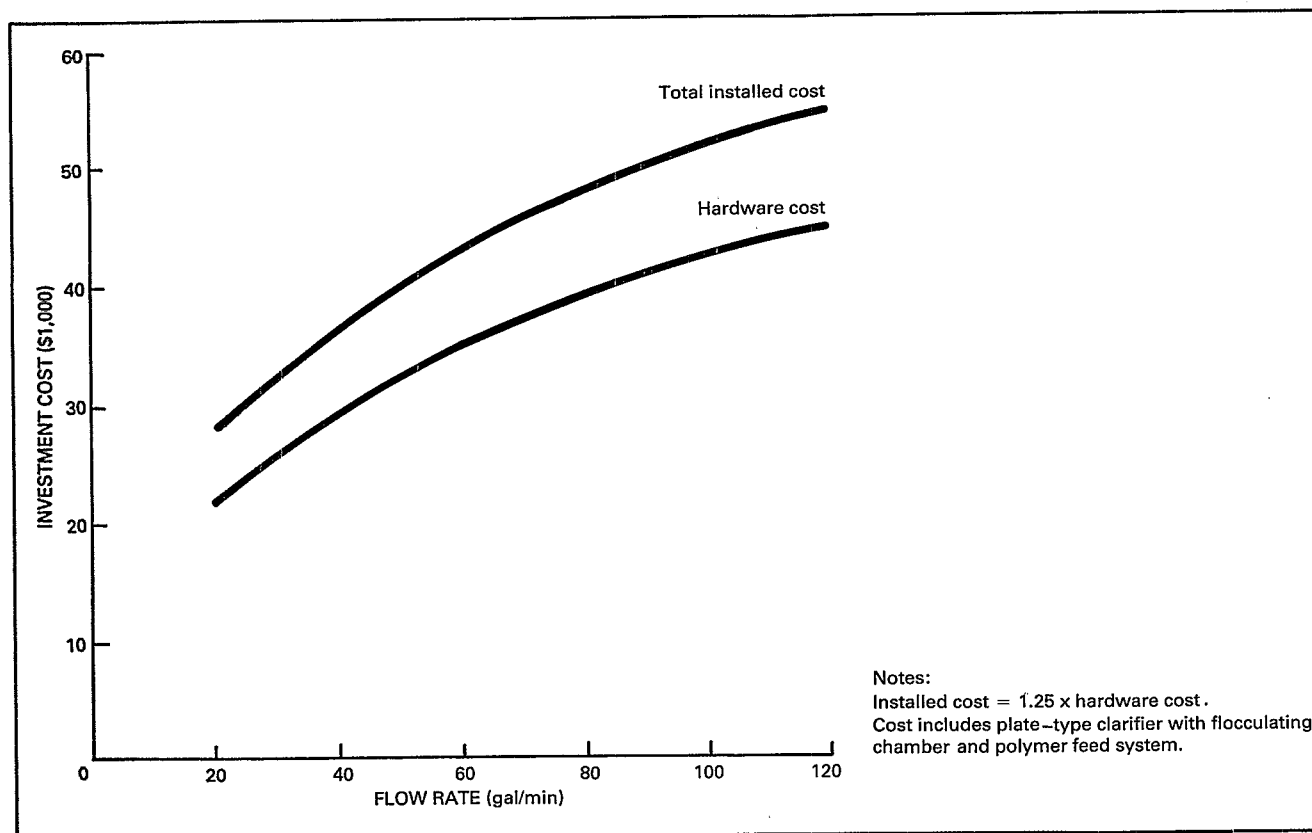
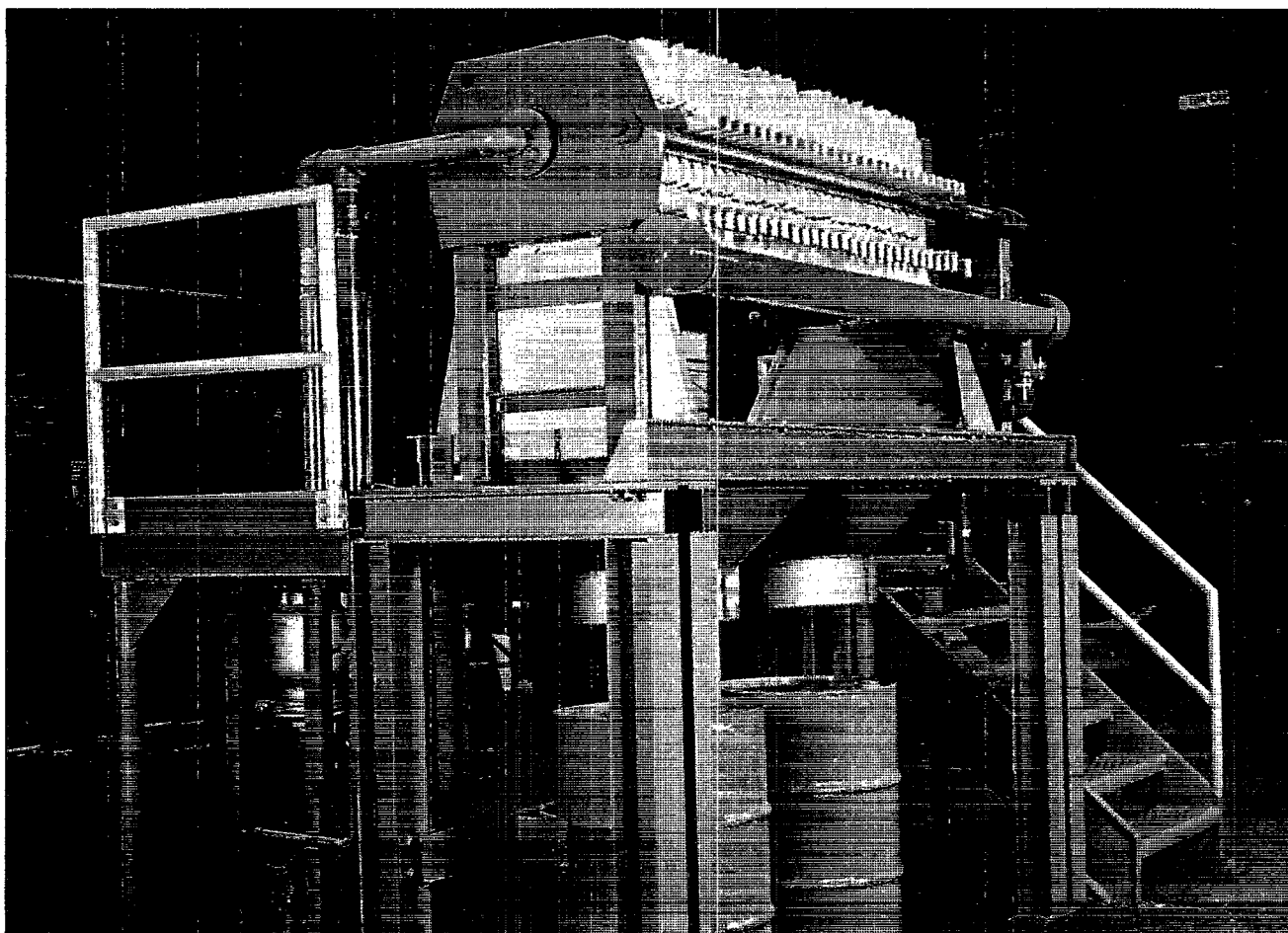


Figure 5. Investment Cost for Flocculation/Clarification Units



Manually shifted low-pressure recessed-plate filter press with capabilities for filling disposal drums directly.

Sludge Handling. The solids from clarifiers are typically discharged to sludge holding tanks at solids concentrations of 0.5 to 3%; overflow from the tank is recycled to the clarifier or to the flocculation tank. Usually, metal hydroxide solids will concentrate to approximately 3 to 5% solids in the sludge holding tanks if sufficient retention time is allowed. The tanks must also provide adequate volume to store the sludge before it is shipped to a disposal site or transferred to another dewatering stage. Figure 6 shows the investment for sludge holding tanks as a function of tank volume.

Further concentration of the thickened sludge requires the use of mechanical dewatering equipment. Centrifuges, rotary vacuum filters, belt filters, and filter presses have been used to dewater metal hydroxide sludge. The applicability of a particular dewatering device for a specific sludge, and the degree of cake dryness the device will achieve, can be determined by bench-scale tests conducted by the vendor using the intended feed material.

Figure 7 presents the unit costs for recessed plate filter presses as a function of the filter-cake volume of the unit. The feed volume capacity of the unit is also given, based on a feed solids concentration of 2%, a cake solids concentration of 20%, and a press cycle of 8 h. The costs shown in Figure 7 do not include installation or auxiliary equipment associated with the press, because these costs are highly variable and site-specific. Items that will contribute to the cost of installation include:

- High pressure feed pump(s) (often quoted and supplied with the filter)
- Sludge feed storage tanks
- Filtrate return lines (to clarifier)
- Cake solids handling equipment.

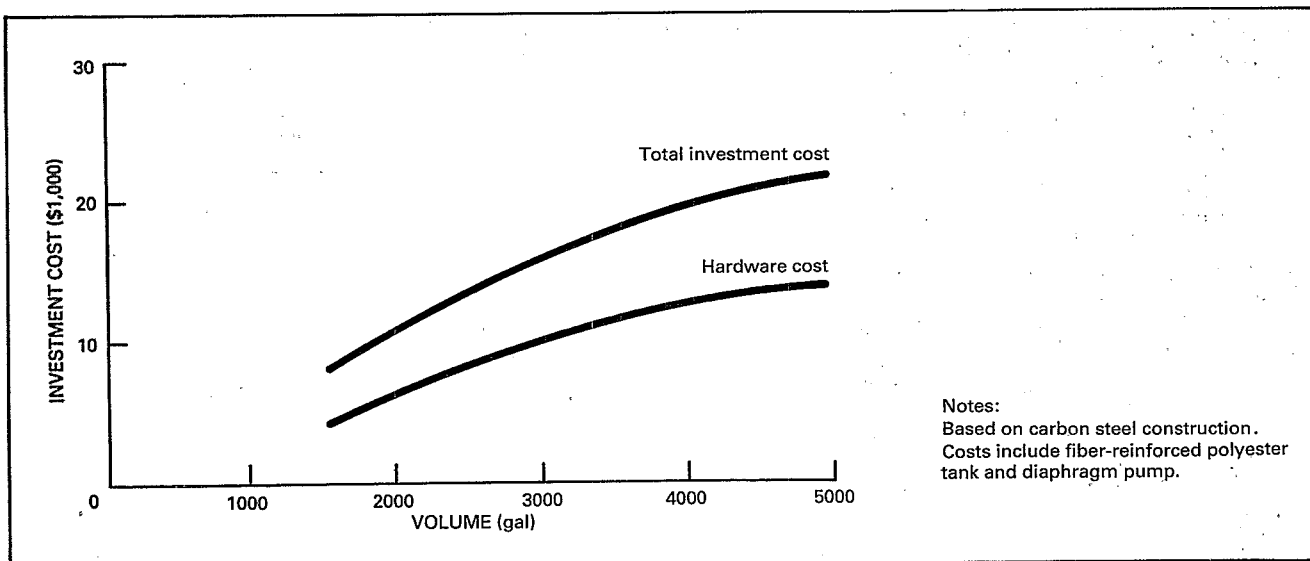


Figure 6. Investment Cost for Sludge Storage/Thickening Units

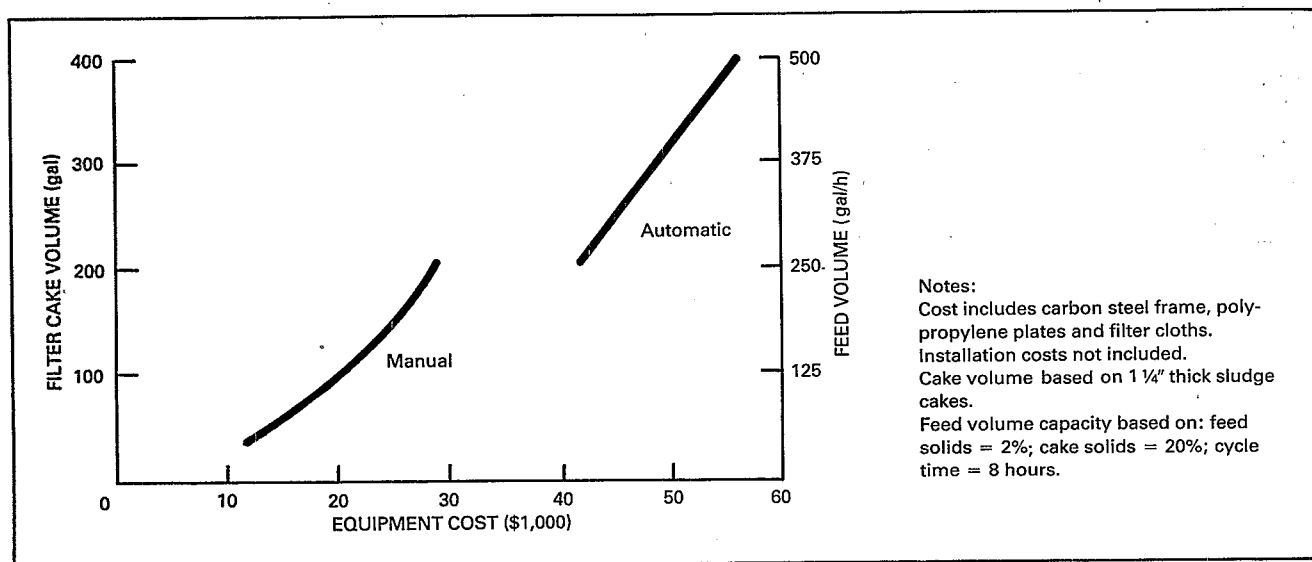


Figure 7. Hardware Cost for Recessed Plate Filter Presses

Operating Costs

Although the investment costs for conventional wastewater treatment systems depend principally on wastewater flow rate, the operating costs depend on the following factors:

- Sludge disposal and municipal sewer fees
- Wastewater treatment chemical costs
- Operating and maintenance labor costs
- Utility costs (primarily electricity to operate pumps and agitators)
- Overhead costs and depreciation.

Of these, sludge disposal and chemical costs offer the best opportunities for reducing operating costs since labor and utility requirements are usually fixed. The cost components of these and other operating costs are described below; Table 4 shows the assumptions for these evaluations.

Table 4.

Basis for Economic Evaluations^a

	Basis
Annual operating costs:	
Operating labor	\$9/h
Supervision	\$12/h
Maintenance	6% of investment
General plant overhead	0.58 (operating + supervisory + maintenance labor); maintenance labor = $0.37 \times (\text{maintenance cost})$
Depreciation (10-year straight line)	10% of investment
Taxes and insurance	1% of investment
Utility charges:	
Electricity	\$0.08/kWhr
Cooling water (cooling tower)	\$0.25/1,000 gal
Steam	\$6.00/1,000 lb
Net operating savings (\$/yr)	Operating cost reduction resulting from investment minus increase in fixed and operating cost for new system
Net savings after taxes (\$/yr)	Net operating savings \times 0.54 (assumes 46% tax rate)
Return on investment	Net savings after taxes \div total installed investment
Cash flow (\$/yr)	Net savings after taxes + depreciation
Payback period (yr)	Total installed investment \div cash flow

^aNo interest on capital is included in the economic analyses.

Sludge Disposal and Municipal Fees. Installation of wastewater treatment systems results in the discharge of two streams: overflow from the clarifier and sludge from the clarifier or dewatering equipment. The costs associated with these discharges are site-specific, and depend on the availability of local disposal sites to receive the sludge and on municipal sewer costs. These costs are expected to escalate in the future as new regulations are implemented.

For the overflow from the clarifier, typical sewer fees for a major city as presented in Table 1 are \$1.20/1,000 gal. Figure 8 shows the direct relationship of wastewater flow rates and these sewer fees.

The cost of hauling the sludge to a licensed hazardous waste landfill will depend on the volume of sludge, the distance hauled, and the sludge composition. In most areas, the costs of sludge disposal are currently in the range of \$.50 to \$1.00/gal, but there are cases where disposal costs run as high as \$2.00/gal. Figure 9 shows the annual disposal costs for each 100 lb. (45 kg) of solids generated daily over a range of sludge concentrations and disposal costs. For example, if 100 lb of solids are generated daily at a concentration of 6% and at a disposal cost of \$1.00/gal, the annual cost for disposal will be \$60,000 (point A on Figure 9).

The disposal cost savings achievable by thickening can also be estimated by using Figure 9 to calculate the difference between the disposal costs at the present concentration and at the projected final concentration after thickening. For example, a plant now disposes of 100 lb/d (45.4 kg/d) of dry solids as a sludge with a concentration of 6% solids. From Figure 9, the disposal cost of 100 lb/d at \$1.00/gal would be \$60,000 per year. If a filter press performance test predicts a sludge concentration of 25% solids, sludge disposal costs would decrease to \$14,000. The disposal costs saved as a result of greater dewatering of the sludge (from 6 to 25% solids concentration) would then be \$46,000 (\$60,000 minus \$14,000).

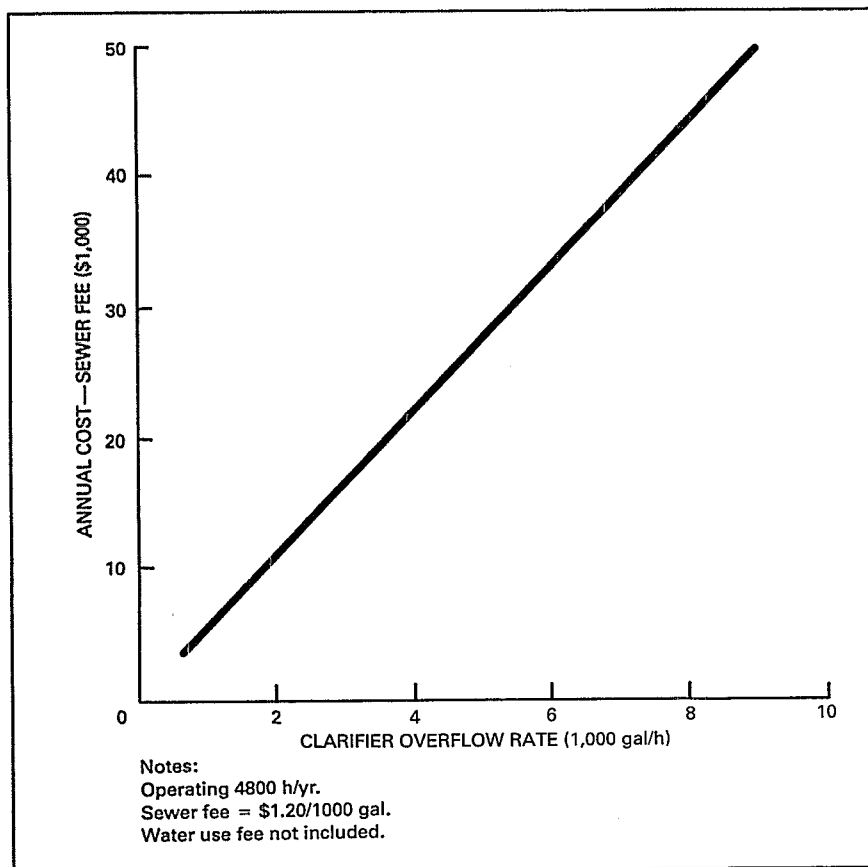


Figure 8. Annual Sewer Fee as a Function of Clarifier Overflow Rate

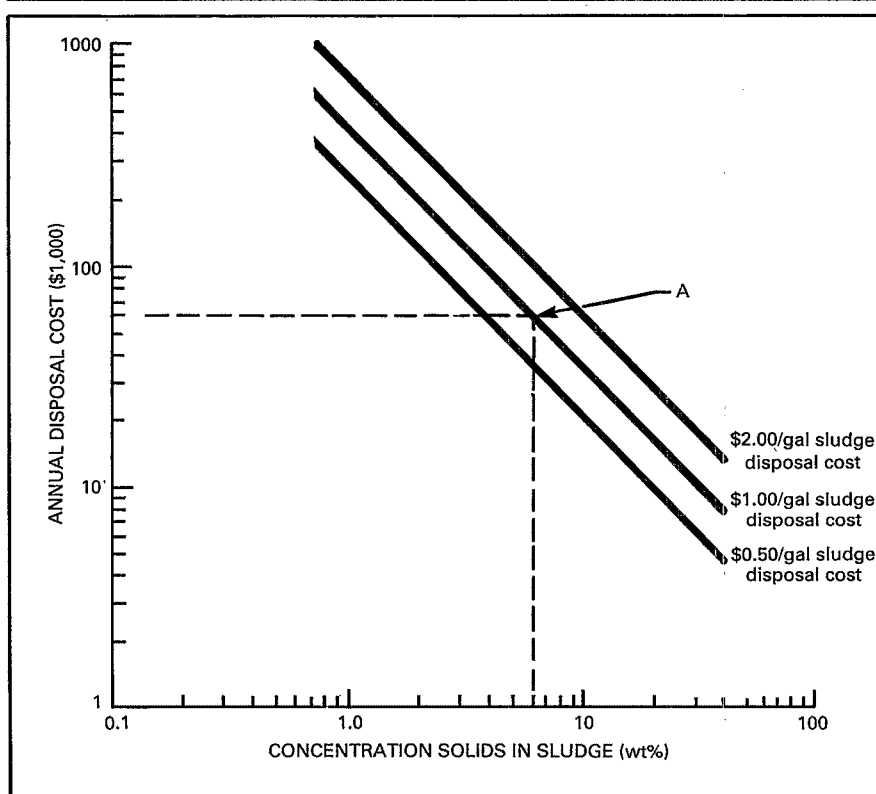


Figure 9. Annual Cost for Disposal of Industrial Sludge (per 100 lb dry solids generated per day)



Filter press hopper with 40% solids filter cake.

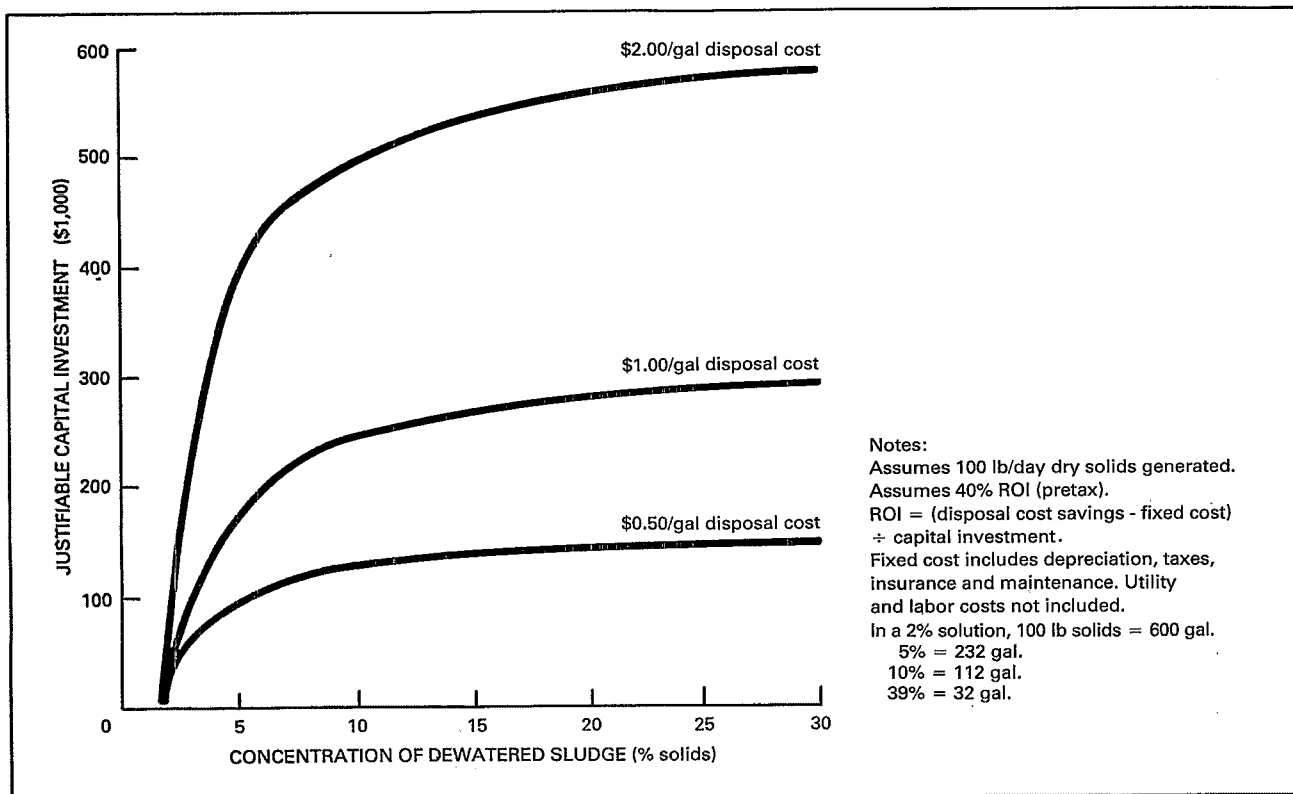


Figure 10. Capital Cost Justification for Sludge Dewatering Equipment

Mechanical dewatering of metal hydroxide sludges can achieve solids concentrations in the range of 15 to 50%. Figure 10 shows the capital investment that could be justified for dewatering equipment to concentrate a sludge containing only 2% solids. The cost reduction used to calculate this return on investment (ROI) did not take into account variable operating costs, such as utility costs and operating labor, which for mechanical dewatering devices could be significant. These additional costs would result in a lower ROI.

To illustrate the investment justification for the mechanical dewatering of a dilute sludge, assume a plant generating 100 lb/d of dry solids was able to concentrate its sludge from 2 to 20% solids with a filter press. The annual disposal cost would be reduced from \$180,000 to \$16,000 based on a disposal cost of \$1.00/gal (Figure 9). The maximum capital investment justified to achieve this saving, using Figure 10, is \$270,000. Figure 7 shows the cost of filter presses as a function of cake volume and feed-processing capacity. Using the feed-processing capacity, the press feed rate would equal:

$$\frac{100 \text{ lb solids}}{d} \times \frac{\text{lb sludge}}{0.02 \text{ lb solids}} \times \frac{\text{gal sludge}}{8.34 \text{ lb}} \times \frac{d}{8h} = \frac{75 \text{ gal sludge}}{h}$$

From Figure 7, the minimum size commercial unit could dewater this feed rate and would cost \$16,000. Assuming the total installed cost of the system is twice the press cost, or \$32,000, the ROI would be well in excess of the 40% used as a basis for Figure 10.

Wastewater Treatment Chemical Costs. These chemical costs are dependent on the concentration of pollutants, the volumetric flow rate of the waste stream, and the types of chemicals chosen for wastewater treatment. Because the addition of chemicals involves a chemical reaction with the pollutants, the types of treatment chemicals selected will produce different use rates, volumes of sludge for disposal, and removal efficiencies. Each of these factors affect the operating costs and must be considered by the plater.

Common treatment reagents used in the electroplating industry are sodium bisulfite (NaHSO_3) for chromium reduction, sodium hypochlorite (NaOCl) for cyanide oxidation, and caustic soda for neutralization/precipitation. These chemicals, and the chemical costs given in Table 1, were used to provide the cost model for conventional wastewater treatment shown in Figure 11. This model enables the user to calculate the consumption of treatment chemicals (consumption factor) and the associated cost (cost factor), based on the volumetric flow rate of the wastewater being treated and mass flow of each pollutant. The sludge disposal cost model (shown in Figure 12) can be used to determine the dry solids generated by

precipitation of the heavy metals contained in the waste stream. Figure 12 also indicates the disposal cost for each pound of metal precipitated, assuming the resulting sludge is 20% solids and the disposal cost is \$1.00/gal.

Many treatment systems use hydrated lime for neutralization instead of sodium hydroxide; some systems use soda ash. The selection of neutralization chemicals is usually based on convenience and price. The choice of neutralization reagent will affect the volume of sludge generated, the costs of sludge disposal, and the investment cost for storage and handling for each chemical. A cost comparison of the various neutralizing reagents is shown in Table 5. Lime requirements

and chemical and sludge disposal costs can be estimated using the model in Figure 13, which is similar to the models developed in Figures 11 and 12 for caustic neutralization. The cost models in Figures 11, 12, and 13 can be used to determine the incremental costs associated with removing each pound of pollutant and treating each gallon of wastewater entering the system. These models can also be used to predict the impact of water use reduction programs, chemical recycle systems, and other modifications on the costs of operating a wastewater treatment system.

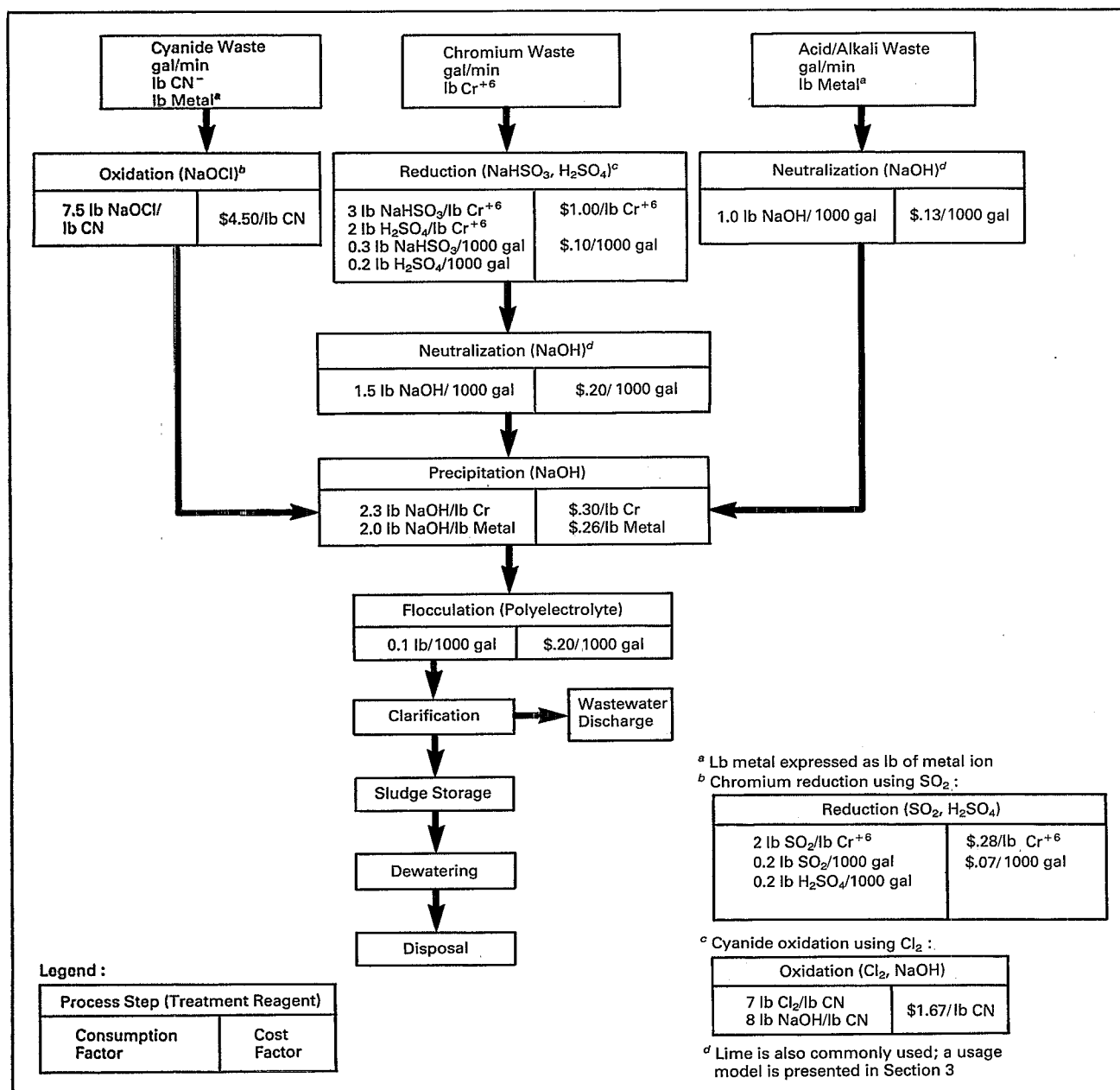


Figure 11. Consumption and Cost Factors for Wastewater Treatment Chemicals

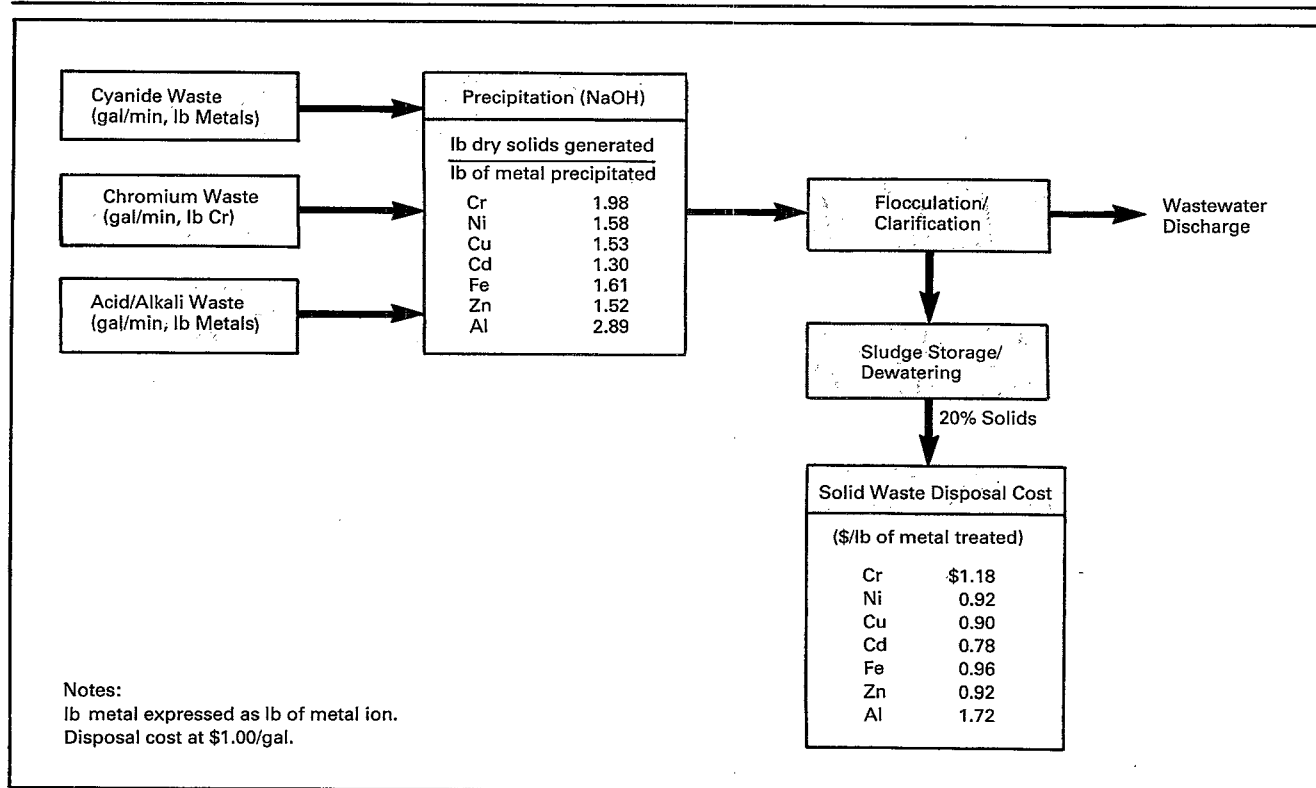


Figure 12. Generation and Cost Factors for Sludge Disposal

Table 5.

Cost Comparison of Common Alkaline Reagents

Agent	Chemical formula	Price (\$/ton)	lb/lb H_2SO_4 neutralized	Relative cost	Process equipment required and sequence for use of reagents
Sodium hydroxide (caustic)	NaOH	\$260 (98% NaOH equivalent)	0.826	9.0	Caustic storage → Neutralizer
Sodium carbonate (soda ash)	Na_2CO_3	\$120 (58% Na_2CO_3)	1.90	9.58	Soda ash storage → Slurry tank → Neutralizer
High calcium hydrate (hydrated lime)	$Ca(OH)_2$	\$50 (96% pure)	0.788	1.65	Hydrated lime storage → Slurry tank → Neutralizer
High calcium lime (quicklime)	CaO	\$40 (96% pure)	0.595	1.00	Quicklime storage → Slaking → Slurry Tank → Neutralizer

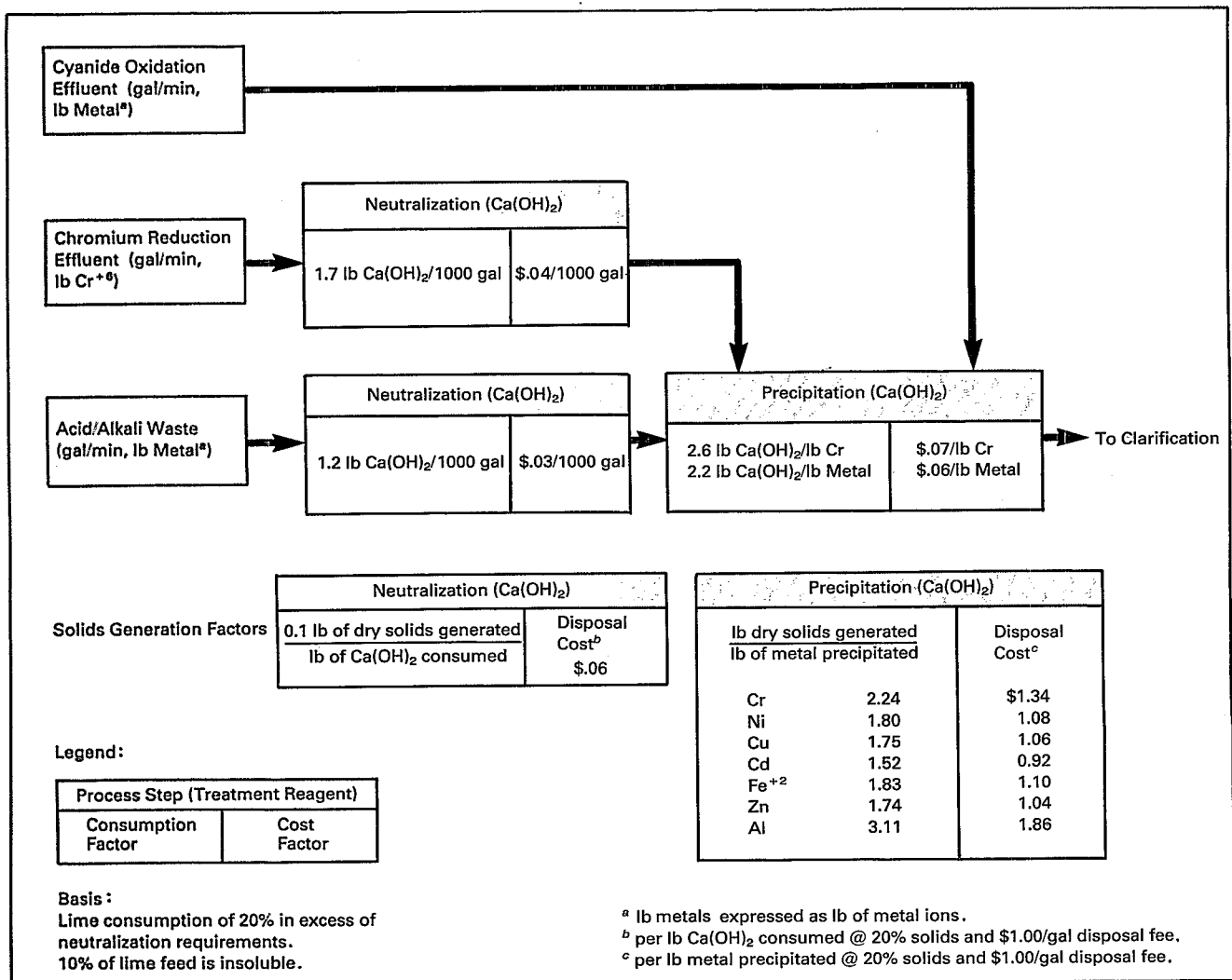


Figure 13. Consumption and Cost Factors Using Lime for Wastewater Treatment

Total Facility Costs—An Example Calculation

This section provides a sample calculation of the investment and operating costs of a hypothetical conventional wastewater treatment system. (A similar analysis can be made at any plant once the wastewater characteristics and flow rates are known.) Figure 14 illustrates the hypothetical treatment system, including the characteristics and flow rates of the wastewater and the required treatment steps. The plant has chromium, cyanide, and a mixed acid/heavy metals wastewater entering the treatment system.

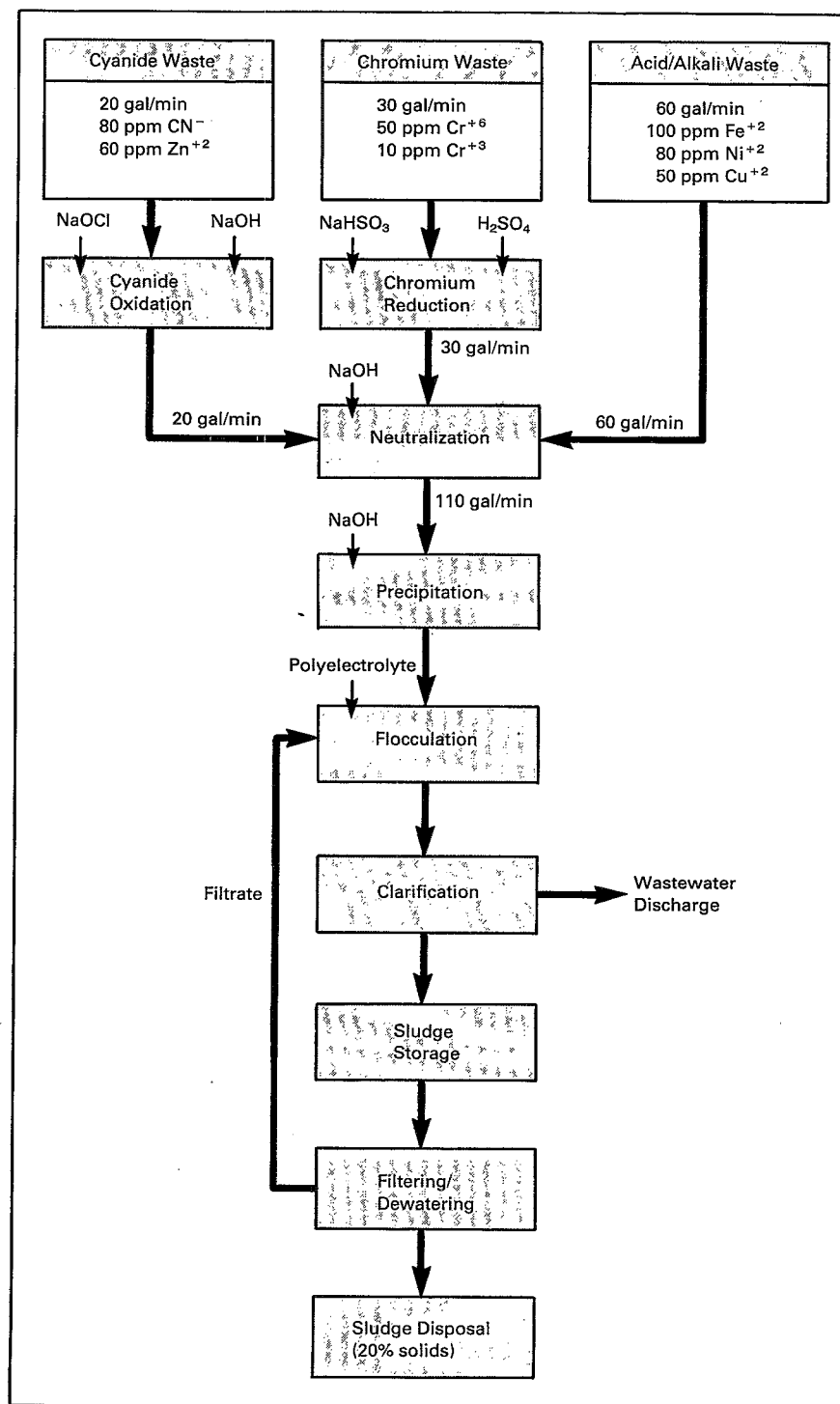


Figure 14. Wastewater Treatment Flow Chart: Example System

Table 6.

Wastewater Treatment Chemical and Sludge Disposal Costs: Example System^a

Treatment step	Waste streams	Treatment chemicals		Sludge disposal		Total annual costs ^d (\$)
		Rates (lb/h)	Cost (\$/h)	Dry solids generated (lb/h)	Disposal cost (\$/h)	
Chromium reduction	30 gal/min = 1,800 gal/hr 0.75 lb/h Cr ⁺⁶ 0.15 lb/h Cr ⁺³	2.79 NaHSO ₃ 1.86 H ₂ SO ₄	0.93			
Cyanide oxidation	20 gal/min = 1,299 gal/h 0.80 lb/h Cn ⁻ 0.60 lb/h Zn ⁺²	6.016 NaOCl	3.60			
Neutralization						
Chrome effluent	1,800 gal/h	2.7 NaOH	0.35			
Cyanide effluent	1,200 gal/h					
Acid/alkali waste	60 gal/min = 3,600 gal/h	3.6 NaOH	0.47			
Precipitation	0.90 lb/h Cr ⁺³ 0.60 lb/h Zn ⁺² 3.01 lb/h Fe ⁺² 2.41 lb/h Ni ⁺² 1.51 lb/h Cu ⁺²	2.1 NaOH 1.2 NaOH 6.0 NaOH 4.8 NaOH 3.0 NaOH	0.27 0.16 0.78 0.62 0.39	1.80 0.91 4.83 3.80 2.30	1.06 0.55 2.89 2.22 1.36	
Subtotal, precipitation		17.1 NaOH	2.22	13.64	8.08	
Flocculation	110 gal/min = 6,600 gal/h	0.66 polyelectrolyte	1.32			
TOTALS						
Treatment chemicals		2.8 NaHSO ₃ 1.9 H ₂ SO ₄ 6.0 NaOCl 29.8 NaOH 0.66 polyelectrolyte	8.89			42,700
Sludge disposal				13.64 ^c	8.08	38,800

^aSystem shown in Figure 14.^bpH adjustment not required.^cSludge volume at 20% solids = 8 gal/h; filter feed at 2% solids = 80 gal/h.^d4,800 h/yr.

Table 6 presents the chemical treatment and sludge disposal costs using the factors presented in Figures 11 and 12. Based on 4,800 h/yr operation (16 h/d, 300 d/yr), costs are \$42,700/yr for chemical treatment and \$38,800/yr for sludge disposal. The required investment for treatment hardware is calculated in Table 7, using the equipment cost data presented in this section. A cost for engineering (10%) and

contingency (10%) were also added to the estimate. The cost for wastewater collection is based on three wastewater conduits, each 60 ft long with two bends and six rinse connections. Each feeds a separate collection sump. Table 8 expresses these results as an annual cost

(using the basis defined in Table 4). Including projected manpower requirements and utility consumption, the total annual cost is \$205,700.

Table 7.**Investment Costs for Wastewater Treatment: Example System^a**

	Cost (\$)
Wastewater collection (3 systems: each 60' of 6" dia. pipe with 2 bends and 6 connections, from Table 3)	8,500
3 sumps @ 1,000 gals (from Table 3)	10,500
Chromium reduction unit (continuous system rated at 30 gal/min, from Figure 2)	28,000
Cyanide oxidation unit (continuous system rated at 20 gal/min, from Figure 3)	36,000
Neutralizer (single stage continuous system rated at 110 gal/min, from Figure 4)	38,000
Flocculation/clarification unit (system rated at 110 gal/min, from Figure 5)	55,000
Sludge storage tank (5,000 gal tank, from Figure 6)	22,000
Filter press (75 gal/hr, from Figure 7; installed cost is 2 × unit cost)	36,000
Total equipment and installation cost	234,000
Contingency ^b	23,000
Engineering ^b	23,000
Total installed cost	280,000

^aSystem shown in Figure 14.^b10% of total equipment and installation cost.**Table 8.****Total Annual Cost for Wastewater Treatment:^a Example System^b**

	Cost (\$/yr)
Operating labor (based on 4 hr per shift)	21,600
Supervision	°
Maintenance	16,800
General plant overhead	16,100
Depreciation	28,000
Taxes and insurance	2,800
Chemical cost (Table 6)	42,700
Sludge disposal cost (Table 6)	38,800
Sewer fee (Figure 8)	31,700
Utilities (electricity; approximately 25 hp required)	7,200
Total annual cost ^d	205,700

^aBased on Table 4.^bSystem shown in Figure 14.^cNone required.^dExcluding interest.

3. Process Modifications to Reduce Costs

The capital and operating costs for wastewater treatment systems were shown in the preceding chapter to depend primarily on the concentration of pollutants in the wastewater and on its volumetric flow rate. Modifications to the design and operation of plating baths and rinse systems can significantly reduce wastewater flow rates and pollutant loading, thereby reducing the amount of new plating chemicals that must be added and reducing the need for additional pollution control.

The two primary areas where such cost reduction efforts are undertaken involve reducing rinse water rates and reducing drag-out losses. Rinse water is the major contributor to the total volumetric flow rate of most electroplaters' wastewater treatment systems. Drag-out (electroplating chemicals inadvertently carried out of the plating bath on a workpiece) is the major contributor to the pollutant loading in most electroplaters' wastewater treatment systems, as well as a significant contributor to the need for replacement chemicals. A number of ways to reduce rinse water use and drag-out losses are presented below, as is an example of a cost/benefit calculation which ordinarily precedes the decision to implement such controls efforts.

Two additional areas where wastewater treatment costs can be readily reduced with little investment in capital or operating funds are:

- Implementing a housekeeping program
- Using spent reagents in wastewater treatment.

Implementing a successful housekeeping program, as a rule, requires little or no capital investment, yet can result in significant savings, especially when the loss of concentrated solutions of plating chemicals is reduced or prevented. The primary activities involved in a housekeeping program aimed at reducing costs are:

- Repairing leaks around processing equipment (tanks, pipes, valves, pump seals, heating coils). Losses of 2 gal/h (7.6 l/h) can easily occur through leaking pump seals alone.
- Installing antisiphon devices equipped with self-closing valves on inlet water lines where required.
- Inspecting tanks and tank liners periodically to avoid failures that may overload the waste treatment system.
- Inspecting plating racks frequently for loose insulation that would cause excessive drag-out of plating solutions.
- Making provisions to ensure that cyanide solutions do not mix with compounds (iron, nickel) that would form difficult-to-treat wastes.
- Using dry cleanup, where possible, instead of flooding with water.
- Installing drip trays and splash guards where required.

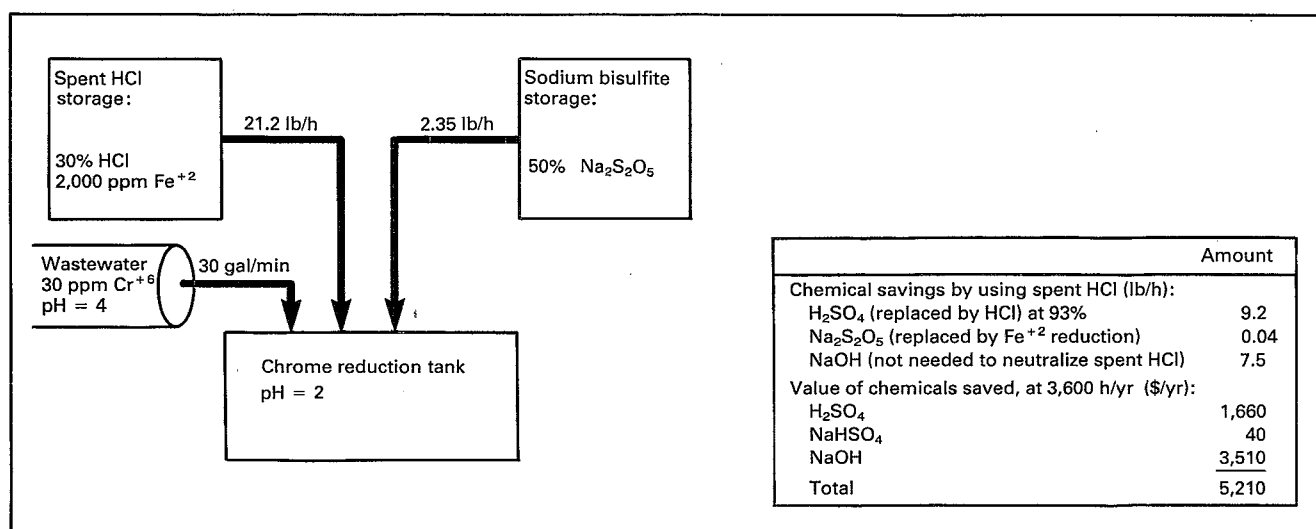


Figure 15. Savings Resulting from Use of Spent HCl in Chrome Reduction Treatment Process

Substantial savings can be realized through a housekeeping program, but can be easily lost if the program is allowed to lapse.

Using spent baths as treatment reagents is a second possible cost-reduction area. In many cases, the processing solutions used in alkaline or acid cleaning can easily be used as pH adjustment reagents in the waste treatment system. Typically, cleaning solutions are either dumped when the contaminant level exceeds some acceptable concentration, or bled off to waste treatment and replaced with fresh reagents. In either case, the solution could instead be transferred to a holding tank and subsequently used to treat the wastewater.

Spent caustic soda solutions can be used to adjust pH in the precipitation tank. Spent sulphuric and hydrochloric acid solutions can also be used here as needed (though the quantity used would be minimal because waste streams are usually acidic).

Waste acid solutions can also be used for pH adjustment in chrome reduction (Figure 15). A minor added benefit in this case would be a decreased demand for reducing agents caused by the presence in the acid of Fe⁺² which reduces Cr⁺⁶.

Decreasing the demand for a reducing agent can be further achieved by dissolving scrap iron in the spent acid, thus raising the concentration of the Fe⁺². However, as suggested in Chapter 2, substituting iron for sodium bisulphite as the reducing agent may drastically increase the amount of hydroxide sludge generated. Depending on the cost of sludge disposal it may, in fact, cost more to dispose of the additional solid waste than will be saved by reducing chemical consumption.

The impact of any additives in the spent solutions on the waste treatment process should be considered before they are used as treatment reagents.

Reducing Rinse Water Use

The greatest potential for reducing water use is in the rinse tanks that follow many of the plating process steps; these account for up to 90 percent of a plant's water demand. Furthermore, if a plant is able to reduce its process water discharge to below 10,000 gal/d (37,850 l/d), it will be classified as a "small plater" and may be regulated by different pretreatment standards. (This size criterion only applies to plants discharging to a POTW.)

Rinsing is used to dilute the concentration of contaminants adhering to the surface of a workpiece to an acceptable level before the workpiece passes on to the next step in the plating operation. The amount of water required to dilute the rinse solution depends on the quantity of chemical drag-in from the upstream rinse or plating tank, the concentration of chemicals in the rinse water, and the contacting efficiency between the workpiece and the water.

Various techniques are used in the electroplating industry to reduce the volume of water needed to achieve the required dilution including:

- Installing multiple rinse tanks (including counterflow rinse tanks) after a processing bath
- Using solenoids controlled by conductivity cells or timers or using flow regulators to control water addition to rinse tanks and avoid excessive dilution of the rinse water
- Reusing contaminated rinse water where feasible
- Subjecting the workpiece to a spray rinse as it emerges from the process tank
- Using air agitation or workpiece agitation to improve plating efficiency.

If multiple rinse tanks are installed so that the rinse flows in a direction counter to that of the parts movement (Figure 16), the amount of chemicals entering the final rinse will be significantly less than the amount that enters a single-tank rinse system. The volume of rinse water required for dilution will be reduced accordingly. The volume for each rinse step can be predicted by using a model that assumes a complete rinsing of the workpiece. The ratio, r , of rinse water volume to drag-out volume is approximated by:

$$r = (C_p/C_n)^{1/n}$$

where

C_p = concentration in process solution

C_n = required concentration in last rinse tank

n = number of rinse tanks.

This model does not predict required rinse rates accurately when the value of r falls below 10. Also, complete rinsing requires sufficient residence time and agitation in the rinse tank.

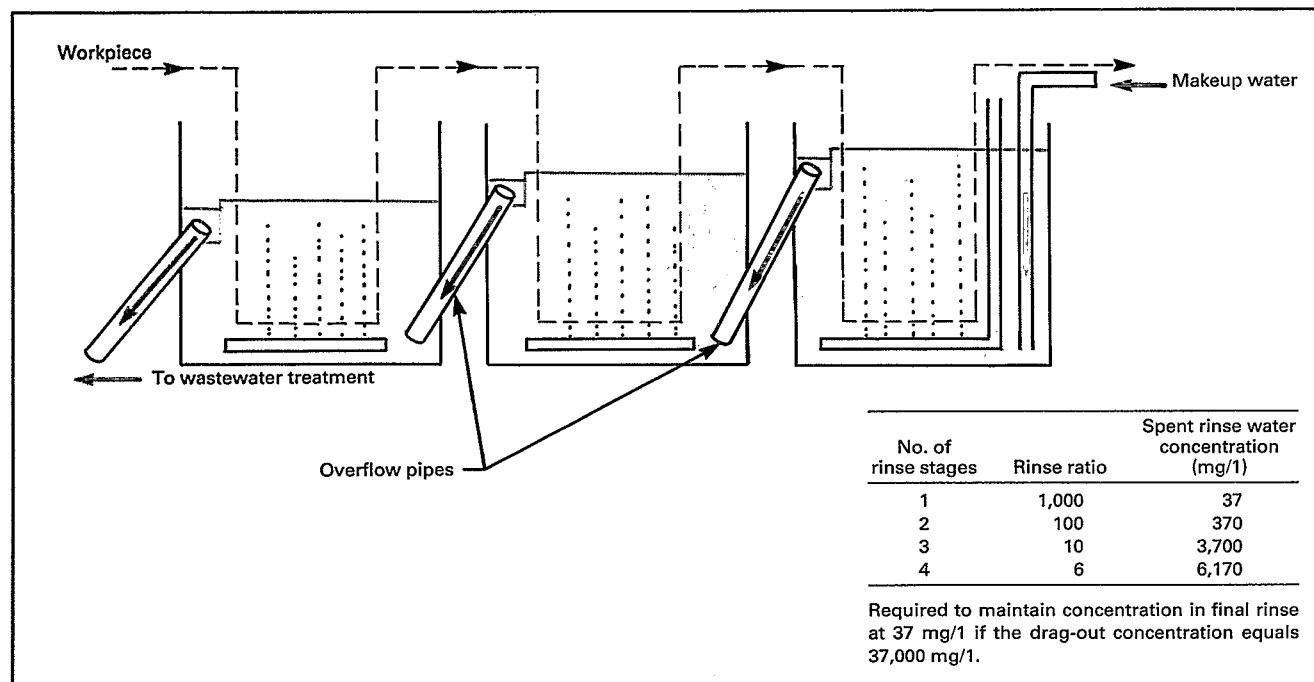


Figure 16. Countercurrent Rinse System

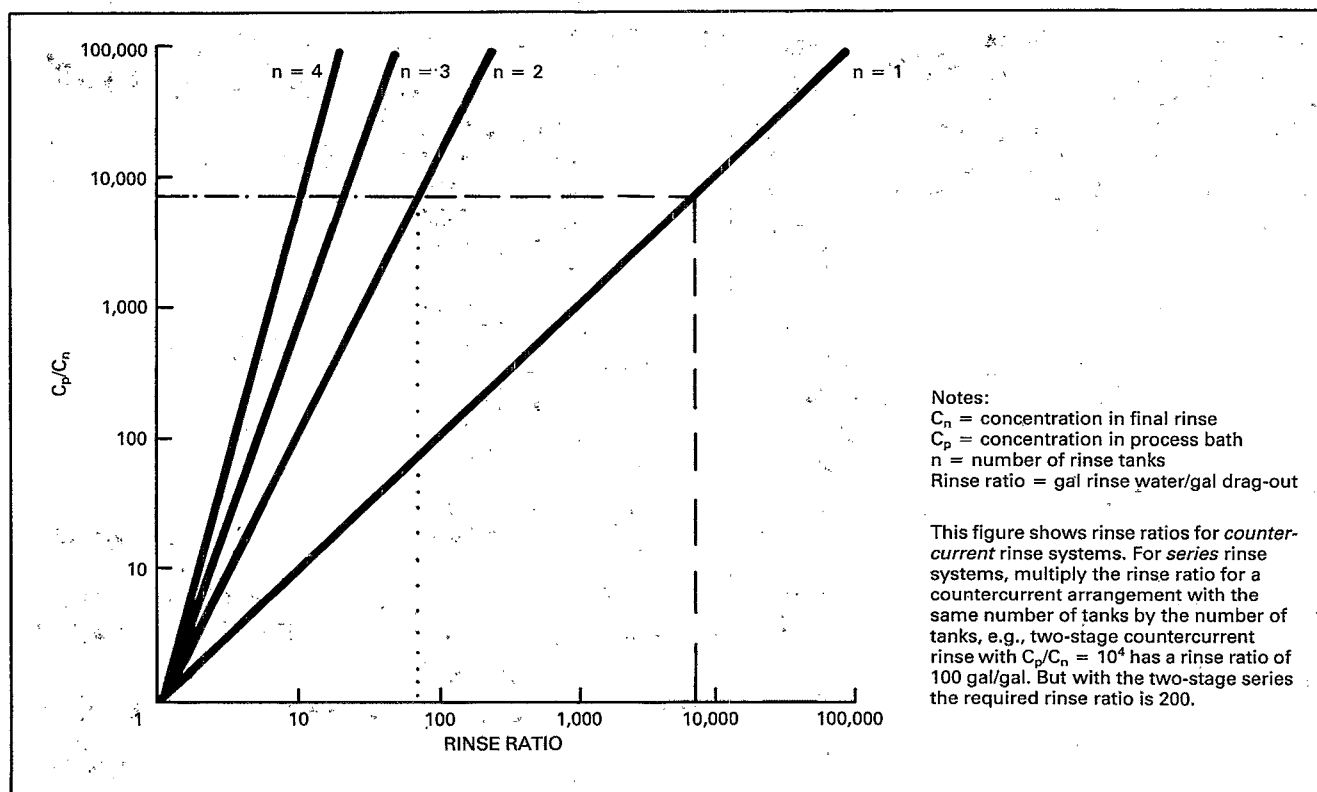
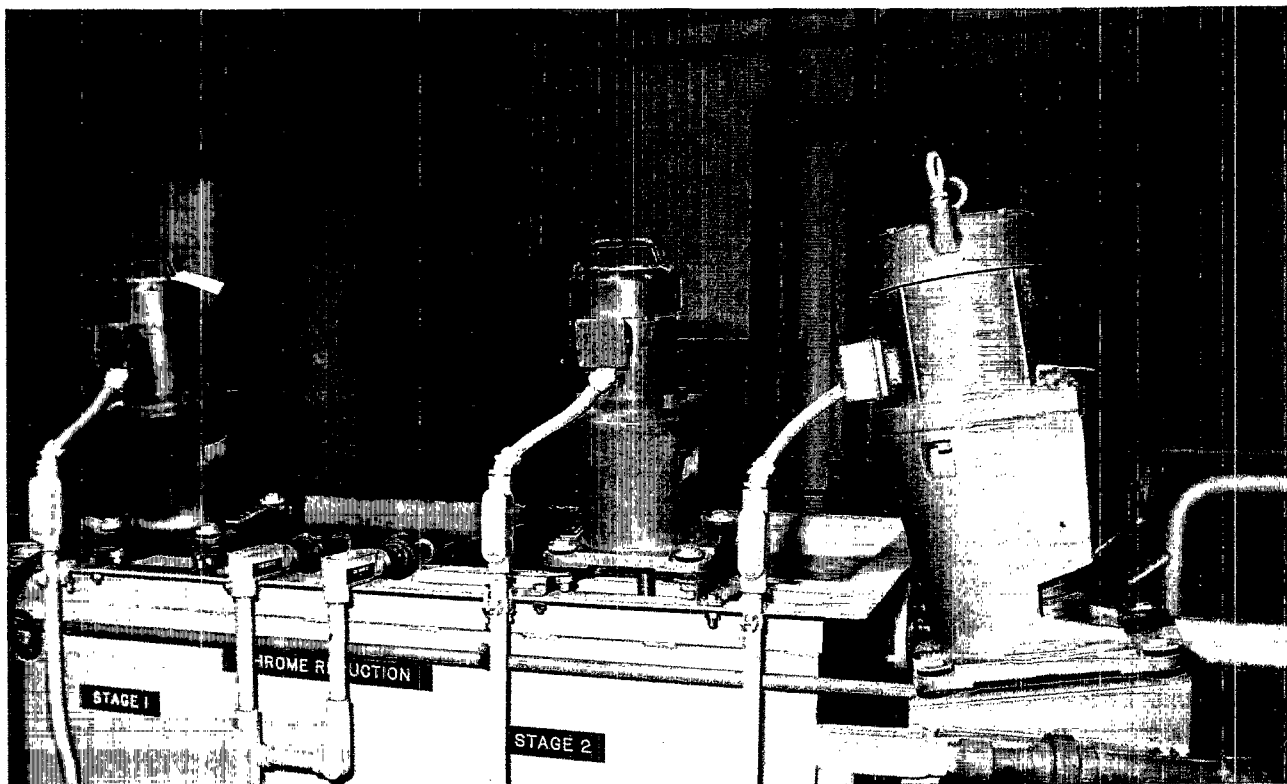


Figure 17. Estimating Rinse Ratios for Multiple-Tank Rinse Systems

Figure 17 shows the volume of rinse water required as a function of initial concentration in the plating bath, required concentration in the final rinse tank, and number of rinse tanks. For example, a typical Watts-type nickel plating solution contains 270,000 mg/l of total dissolved solids (C_p), and the final rinse must contain no more than 37 mg/l of dissolved solids (C_n). The ratio of C_p/C_n is 7,300; hence, 7,300 gal of rinse water would be required for each gallon (3.8 l) of process solution drag-in, assuming a single-tank rinse system. By installing a two-stage rinse system, the same degree of dilution is achieved with only 86 gal (326 l) of water per gallon of process solution drag-in, a reduction in rinse water consumption of almost 99%. The mass flow of pollutants leaving the rinse system remains constant.

A three-stage countercurrent rinse arrangement would further reduce water consumption to 20 gal/h (76 l/h). The resulting cost savings by going from a one-stage to a three-stage rinse system would include reducing water use and sewer fees by \$14.56/h (based on \$2.00/1,000 gal combined water use and sewer fees as shown in Table 1) and reducing the size of the required waste treatment systems. The investment cost to add two additional rinse tanks is highly site-specific. For manual plating operations, the major factor affecting cost is the availability of space in the process area. For automatic plating machines, the cost of modifying the unit to add additional stations may be as high as \$20,000 per station. Rubber-lined steel open-top tanks with appropriate weir plates and nozzles cost anywhere from \$1,000 to \$3,000, depending on the cross-sectional area required for the workpiece.

A series rinse arrangement can also be used with multiple rinse tanks. In this case, each rinse tank receives a fresh water feed and discharges the overflow to waste treatment. The rinse ratio required for a series rinse arrangement is defined by $r = n (C_p/C_n)^{1/n}$. If the rate given in Figure 17 for a countercurrent rinse system with the same number of rinse tanks is multiplied by the number of rinse tanks, the series rinse water rate can be estimated. Rinse water rates are significantly higher for series rinsing than for countercurrent.



Two-stage chrome reduction tank.

A conductivity probe is another effective water-saving device used in rinse systems. Except on highly automated plating machines, the frequency of rinse dips can vary considerably. This means that because the fresh rinse water usually is fed continuously, there are periods where considerably more water is used than is needed for dilution. A conductivity cell can measure the level of dissolved solids in a rinse water. When the level reaches a preset minimum, the cell sends a current to close a solenoid valve on the fresh water feed. When the concentration of dissolved solids reaches the maximum desired level, the solenoid valve is opened. These units are generally reasonably priced. A complete set, including a probe, controller, and automatic valve can be purchased and installed for \$300 to \$1,000. A similar system consisting of a solenoid operated by a timer may be even less expensive.

A further water conservation step uses flow regulators as a means of controlling the fresh water feed rate within a narrow range despite variations in water pressure. These devices, which cost from \$10 to \$40, also eliminate the need to reset the flow rate each time the valve is opened. Some flow regulators are designed to act as siphon breakers and aerators (by the venturi effect).

Reusing rinse water is another means of reducing water use. In critical or final rinse operations, the amount of contaminants remaining on the workpiece must be extremely small; for some intermediate rinse steps, however, the level of contaminants can be higher. Water consumption can be reduced by reusing the contaminated overflow from the critical rinse in a less critical rinse. Rinse water may also be reused when the contaminants in it do not affect the rinse water quality required in a subsequent rinse. For example, the overflow from an acid dip rinse may be reused as the feed to a rinse after an alkaline dip. Interconnections between rinsing systems might make operations more complicated, but the cost advantage they represent frequently justifies the extra attention required.

Reducing Drag-Out Loss

As a workpiece emerges from a plating bath, it carries some of the plating solution into the rinse. As Table 9 shows, this carryover, known as drag-out, can result in a significant economic penalty for each pound of plating chemicals lost to the waste stream. Plant modifications to minimize drag-out are frequently cost-effective, due to the high cost of replacing the raw materials and treating and disposing of the waste.

Drag-out losses can be minimized either by reducing the amount of plating solution which leaves the plating bath, or by recycling plating chemicals in the rinse water to the plating bath. The cost-effectiveness of each of these methods is discussed below.

Assessing the potential economy of drag-out recovery requires that the quantity of plating chemicals lost to the rinse system be determined. A first approximation of this quantity can be derived by multiplying the quantity of plating chemicals added to the bath by an assumed loss factor. In chrome plating operations, about 0.9 lb of chrome is typically lost as drag-out for every pound of chrome added to the plating tank. The loss factors for other plating operations are between 50 and 90%.

If, as a result of a preliminary assessment, chemical loss due to drag-out represents a significant cost, a more precise determination can be undertaken to substantiate the benefits of investing in drag-out recovery modifications. The following five steps constitute the recommended analysis:

1. Fill the rinse station after the process bath with a known volume of water.
2. Using normal production procedures, plate and rinse a representative production unit.
3. Stir the rinse tank and collect a sample of rinse water.
4. Plate and rinse several additional production units and collect samples of rinse water after each.
5. Find the concentration of plating chemicals in the rinse water by laboratory analysis.

Table 9.

Economic Consequences of Plating Chemicals Losses

Chemical	Cost (\$/lb)			Total
	Replacement	Treatment ^a	Disposal ^b	
Nickel:				
As NiSO ₄	1.19	0.20	0.35	1.74
As NiCl ₂	1.05	0.30	0.41	1.76
Zinc cyanide as Zn(CN) ₂ :				
Using Cl ₂ for cyanide oxidation	2.00	1.03	0.57	3.60
Using NaOCl for cyanide oxidation	2.00	2.02	0.57	4.59
Chromic acid as H ₂ CrO ₄ :				
Using SO ₂ for chromium reduction	1.18	0.51	0.52	2.21
Using NaHSO ₃ for chromium reduction	1.18	0.84	0.52	2.54
Copper cyanide as Cu(CN) ₂ :				
Using Cl ₂ for cyanide oxidation	2.62	1.02	0.50	4.14
Using NaOCl for cyanide oxidation	2.62	2.30	0.50	5.42
Copper sulfate as CuSO ₄	0.88	0.20	0.36	1.44

^aBased on treatment model presented in Figure 12 at a concentration of 100 mg/l in wastewater.

^bBased on Figure 14.

Multiplying the volume of the rinse tank by the concentration of each chemical will determine the quantity of chemical drag-out per production unit. The volume of drag-out per hour can be determined if the production rate and the chemical concentrations of the plating solution are known.

Recovering Drag-Out from Rinse Tanks. The drag-out lost from the plating bath can be reduced significantly by usually low-cost modifications after other modifications to reduce rinse water use have been completed. One of these new modifications is a recycling system, including the countercurrent rinse system, which returns concentrated solutions of dragged-out plating chemicals to the plating bath to make up for water lost by surface evaporation. The amount of chemicals actually recovered depends on the amount of chemicals lost from the plating tank, the number of rinse tanks used, the concentration of chemicals permitted in the final rinse tank, and the rate at which rinse water can be recycled to the plating tank.

Of these, the rate at which rinse water can be recycled to the plating tank is usually the most critical; it is primarily dependent on the amount of surface evaporation from the plating tank.

High temperatures increase the surface evaporation from the plating baths. However, when the heat is increased to get higher evaporation, the increased temperature may destroy heat-sensitive additives in the bath. New additives that are not as readily degraded have recently been developed for many plating applications, making these operations possible at higher temperatures, and thus facilitating recycling techniques. Usually, the increased energy cost to operate the bath at a higher temperature is justified by the value of the recovered plating chemicals.

Surface evaporation can also be increased by injecting air bubbles into the bath (air agitation). The rate of surface evaporation for plating tanks with air agitators is shown in Figure 18; the rate for those without air agitators is shown in Figure 19. If air agitators significantly increase the evaporation rate, they will also significantly increase the heat loss from a plating tank and the energy cost to maintain the bath temperature. Figure 20 shows the heat input required to compensate for heat loss resulting from the use of air agitators. Heat loss caused by surface evaporation in a plating bath without air agitators can be calculated from: $\text{heat load (Btu/h)} = \text{surface evaporation (gal/h)} \times 8,300 \text{ (Btu/gal)}$.

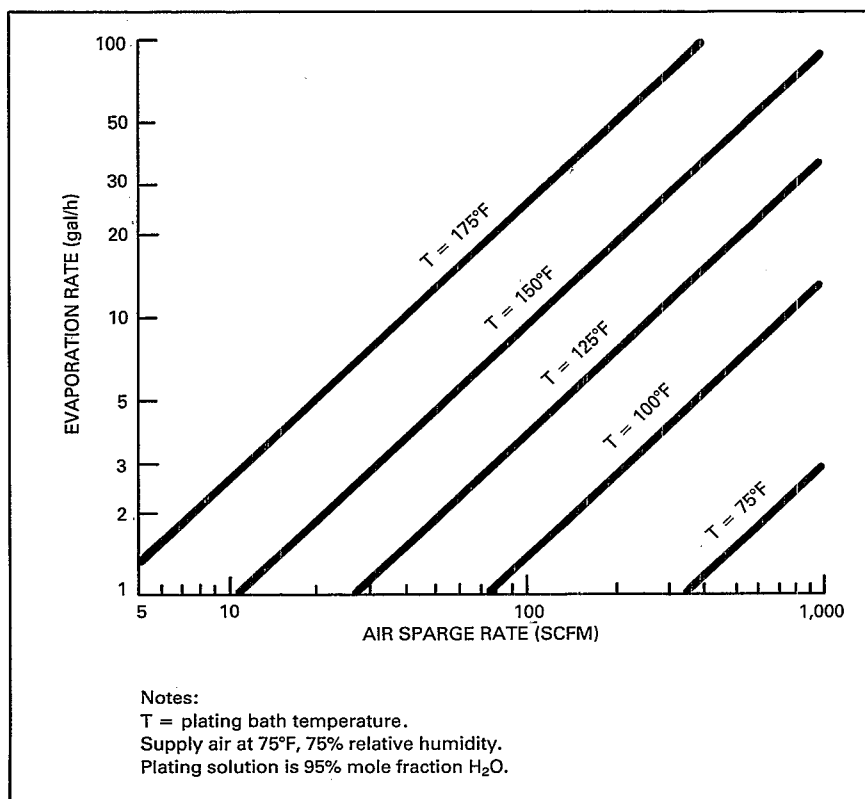


Figure 18. Surface Evaporation Rate from Plating Baths with Aeration

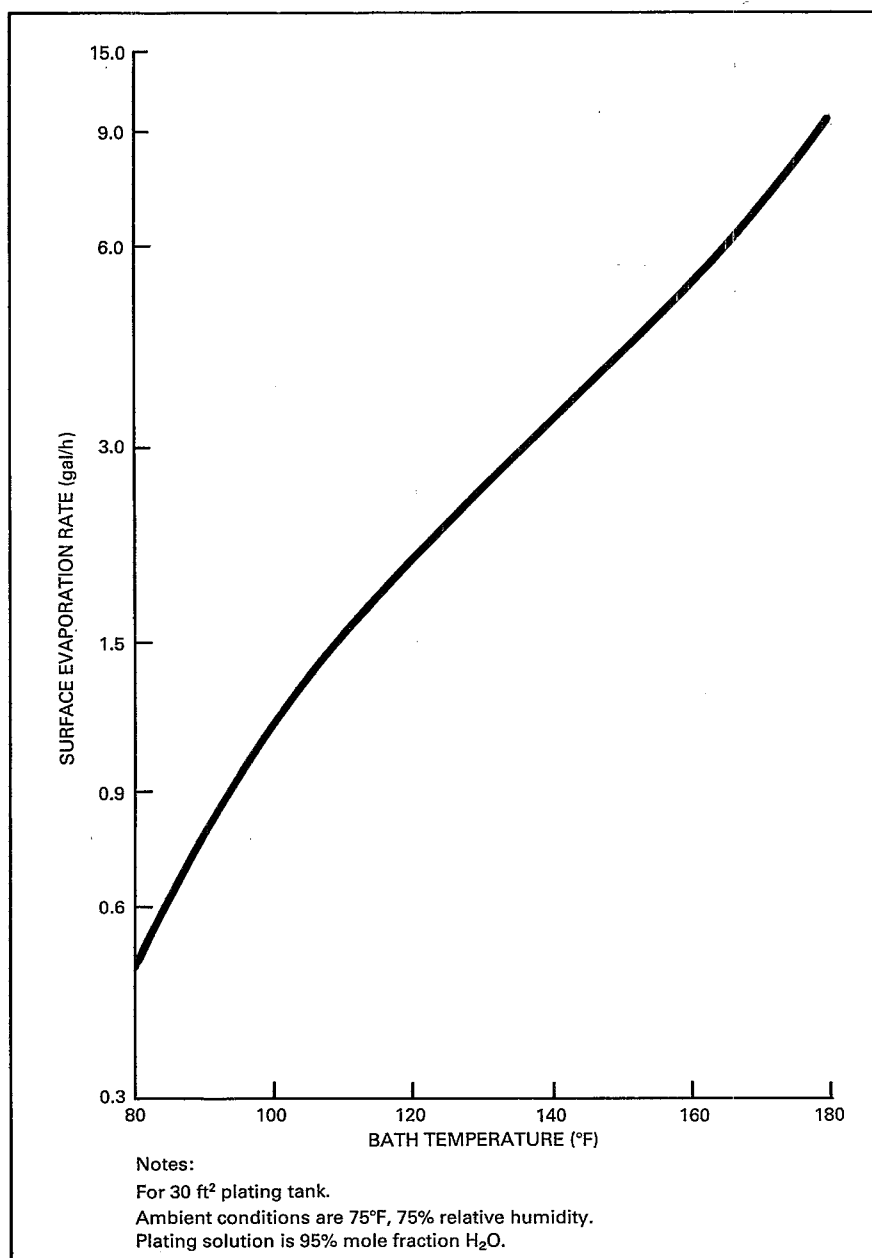


Figure 19. Surface Evaporation Rate from Plating Baths without Aeration

For example, two plating tanks, each with a 30 ft² (2.8 m²) surface area, are operated at 150°F (66°C). One uses 100 scfm (2.8 normal m³/min) air for agitation; the second operates without air agitation. The respective surface evaporation rates are 9.8 gal/h (37.1 l/h) (Figure 18) and 4 gal/h (15.9 l/h) (Figure 19). The respective heat inputs required are 107,500 Btu/hr (Figure 20) and 34,860 Btu/h using the formula above. Using indirect steam heating to compensate for the heat loss would cost \$0.64/h for the agitated bath compared to \$0.20/h for the bath without air agitation, based on an energy cost of \$6/MMBtu.

If the evaporation rate can be matched to the required rinse water rate, the entire volume of rinse water could be returned to the plating bath. This set-up is referred to as a closed-loop recovery system. In this case, the reduction of drag-out loss can be estimated by the following formula:

Percent recovery of drag-out =

$$1 - \frac{C_n}{C_p} \times 100$$

where

C_p = concentration in plating bath

C_n = concentration in final rinse tank.

In a closed-loop rinse water system the only chemical loss is from the drag-out after the last rinse tank, which has a dilute concentration of plating chemicals.

A closed-loop system may be impractical when:

- A very low final rinse concentration is required and only achievable through a larger number of rinse stages.
- Excessive drag-out is unavoidable.
- Plating tank surface evaporation is minimal.

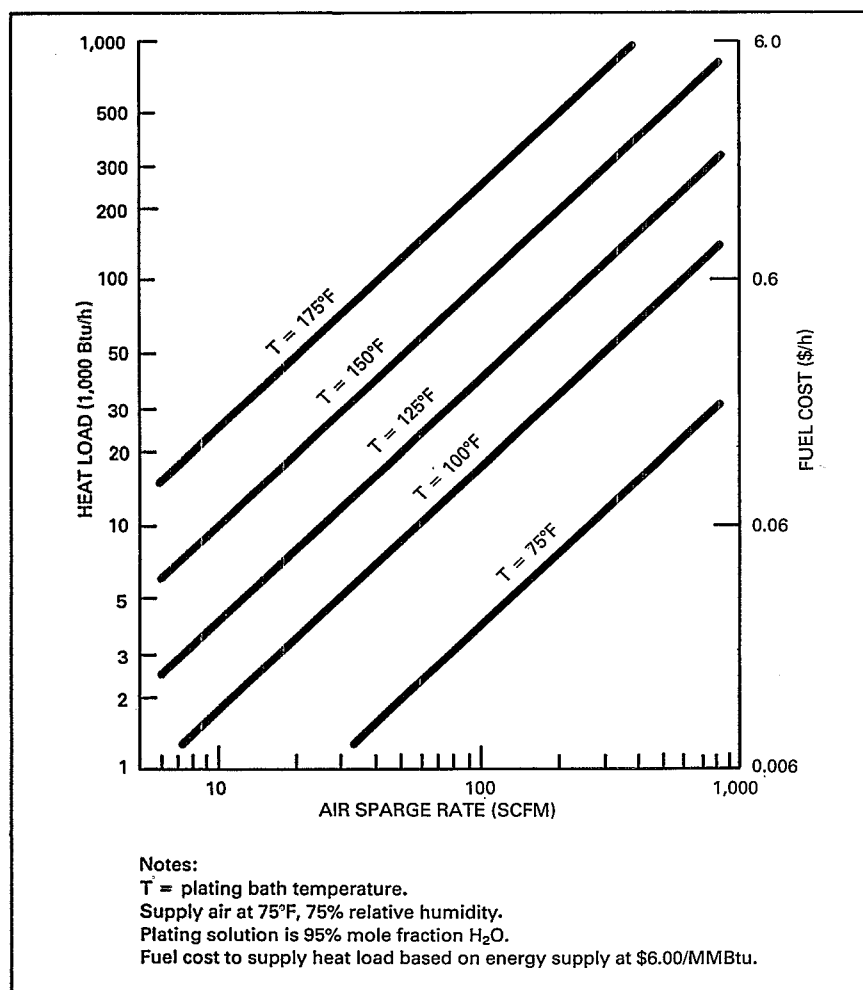


Figure 20. Input Required with Heat Aeration of Plating Baths

The low final concentration problem can be overcome in many cases by operating the final rinse as a free (running) rinse, and using the upstream tanks as a countercurrent rinse-and-recycle system. Using this approach, significant drag-out recovery can be realized while providing a final rinse with a low level of contaminants. Figure 21 shows an automatic rinse-and-recycle system with a running rinse. Level-control devices in the plating and rinse tanks control the flow of rinse water through the system. Figure 22 shows the percent recovery of drag-out for such a

system as a function of the recycle ratio, defined as the volume of recycled rinse divided by the volume of drag-out in a given time. In Figure 22, the recycle rate is assumed to be equal to the evaporation rate.

For example, a nickel plating operation has these operating characteristics: a drag-out rate of 0.5 gal/h (1.9 l/h), a surface evaporation rate of 5 gal/h (19 l/h), and a final

rinse concentration of 40 mg/l. Therefore, the recycle rinse ratio could be set at 10. As seen in Figure 22, a one-stage recovery rinse and recycle system would reclaim 91% of the drag-out (point A). At this recovery rate, the concentration ratio is $0.09 (100\% - 91\%) \div 100\%$. Assuming an initial plating tank concentration of 270,000 mg/l, the concentration entering the final rinse is $0.09 \times 270,000$, or 24,300 mg/l. The makeup water requirements in the final rinse would be reduced by the same level as drag-out losses, when compared with the required rinse rates for a single-tank rinse system. The rinse water required in the final rinse tank to achieve 40 mg/l is 300 gal/h (1,150 l/h).

The data presented in Figure 22 is based on the assumption that make-up water is added continuously to the rinse tanks as surface evaporation occurs. Typically, recycled water is added in increments rather than continuously, in which case one would see cyclical movement along the same curves shown in Figure 22. The longer the time interval between additions, the greater the variation in the recovery of the drag-out realized. Particularly at low recycle rates where the recovery potential is very sensitive to changes in the recycle ratio, minimizing the time between additions will significantly increase the amount of drag-out recovered. A level-control device will approach the potential of continuous water addition and is recommended if the recycle ratio is in the range of 3 or less. These control devices cost from \$700 to \$1,500.

In any of these recycling systems, deionized water is specified for any rinse stream that is recirculated to the plating bath in order to avoid the progressive buildup of contaminants in the bath.

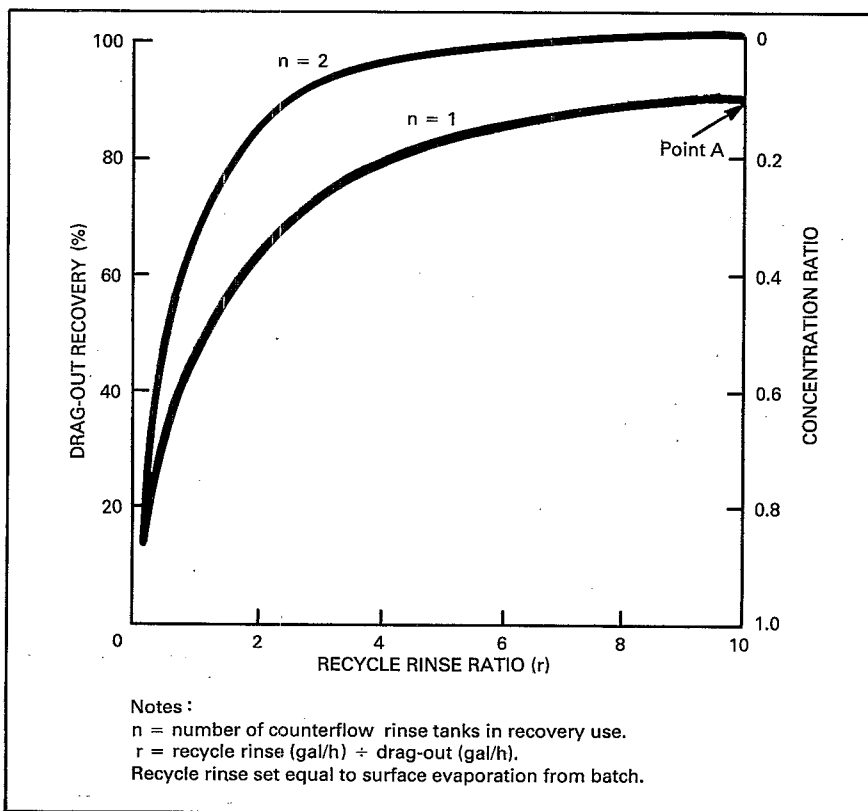


Figure 22. Drag-Out Recovery Rate for Rinse-and-Recycle Systems

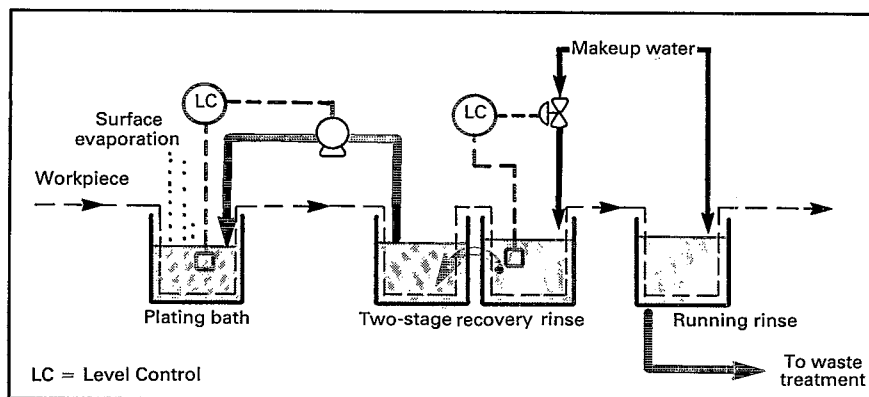
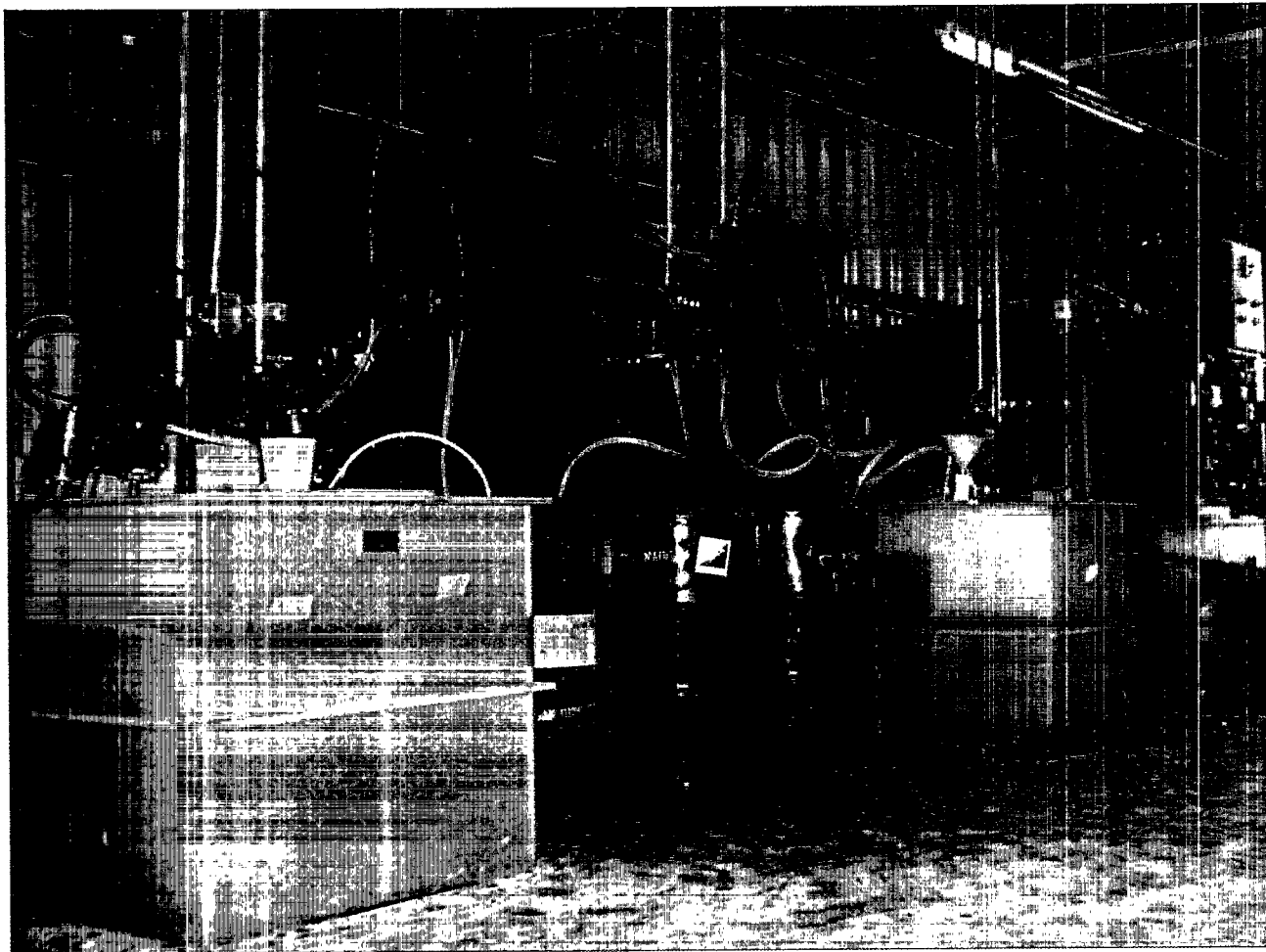


Figure 21. Automated Rinse-and-Recycle System



Four-stage chelated treatment unit (red) with clarifier (blue).

Reducing Drag-Out from Plating Tanks. There are three effective methods of reducing the concentration or amount of plating solution lost from plating tanks: spray rinses, air knives, and minimizing plating bath concentrations.

Spray rinses are ideal for reducing drag-out from plating tanks on automated lines. As the workpiece is mechanically withdrawn from the plating solution, a spray of water automatically washes the workpiece, returning as much as 75% of the drag-out chemicals back to the plating tank. Spray rinsing is best suited for flat parts, but will reduce drag-out effectively on any plated part. The volume of spray rinse cannot exceed the volume of surface evaporation from the plating tank.

The savings are calculated in terms of the concentration change in the drag-out. For example, if the concentration of the drag-out was 100,000 mg/l and a spray rinse reduced the concentration to 50,000 mg/l, the chemical losses would be reduced by 50%.

An air knife reduces drag-out in much the same way as a spray rinse, and is particularly useful when the surface evaporation rate in the plating bath is low. Air knives reduce the volume of drag-out adhering to the workpiece by subjecting the workpiece to a high-velocity stream of air. The drag-out is returned to the plating bath without changing its concentration.

A third means of reducing chemical losses from drag-out is by reducing the concentration of the plating bath, since the amount of loss is determined both by the volume of the drag-out and its concentration. Typically, the target operating concentration is the midpoint of a range of acceptable operating concentrations. This practice is sound unless the savings in chemical replacement costs are exceeded by the costs of controls needed to maintain the target operating concentrations.

An example of how to calculate the savings from not having to replace the additional chemicals involves a standard nickel plating solution having the concentration limits shown in Table 10. The plating shop operates at an average of 12 h/d, 250 d/yr, and processes 600 ft²/h (56 m²/h). Drag-out losses are estimated at 2,700 gal/yr (10,220 l/yr), based on a drag-out rate of 1.5 gal/ 1,000 ft² (61 l/1,000 m²). Modifying the operating conditions to the minimum values indicated in Table 10 would save this shop 390 lb (177 kg) of nickel sulfate and 150 lb (68 kg) of nickel chloride annually. The money saved in replacement chemicals, treatment costs, and sludge disposal (as shown in Table 9) would amount to \$940/yr.

Example of Cost/Benefit Analysis.

The following example illustrates how to calculate the potential savings in water and chemicals that result from the application of the modifications discussed in this chapter to a typical nickel-chromium plating operation. The worksheet provided in Appendix A can be used to develop a similar analysis for most plating shops.

The shop plates approximately 600 ft²/h (56 m²/h) in its nickel-chromium operation, operating an average of 10 h/d, 300 d/yr. Figure 23 shows the original processing sequence and water use rates for the operation. This processing sequence used two-stage countercurrent rinse systems after the nickel and chromium plate tanks. As shown in Figure 24, in-plant modifications were made at six locations (Station 2 and Stations 4 through 8) to reduce raw material losses and waste treatment costs.

For the alkaline rinse (Station 2) and pickling rinse (Station 4), testing indicated that with air agitation the rinse rate for each station could be reduced from 360 to 180 gal/h (1,363 to 681 l/h) with adequate rinsing efficiency. This reduction was accomplished by installing a venturi-style water flow regulator that also provided air agitation. In addition, the overflow from the acid rinse was fed to a suction pump and was used as the feed to the alkaline rinse. These modifications reduced process water demand at these two stations from 720 to 180 gal/h (2,725 to 680 l/h). Costs for the modifications came to \$2,500:

- Pump and foundation, \$1,600
- Flow regulators, piping, valves, and electrical connections, \$300
- Labor, \$600.

In the nickel plating and rinse stations (5 and 6), the plating bath operates at 150°F (66°C) and has the following chemical composition:

- NiSO₄·6H₂O @ 45 oz/gal (337 g/l) or NiSO₄ @ 1.65 lb/gal (0.2 kg/l)
- NiCl₂·6H₂O @ 10 oz/gal (75 g/l), or NiCl₂ @ 0.34 lb/gal (0.04 kg/l)
- H₃BO₃ @ 6 oz/gal (45 g/l), or 0.38 lb/gal (0.045 kg/l)
- Specific Gravity = 1.25

The plating tank has a surface area of 30 ft² (2.8 m²) and drag-out is determined by testing to equal 1.5 gal/1,000 ft² (61 l/1,000 m²) of work plated, or 0.9 gal/h (3.41 l/h) drag-out at 600 ft²/h (56 m²/h) plated. The tank is aerated at a rate of 60 scfm (1.7 normal m³/min). From Figure 18, the evaporative rate in the plating tank is 5.85 gal/h (22.14 l/h).

Table 10.

Standard Nickel Solution Concentration Limits

Chemical	Concentration range (oz/gal)	Operating condition (oz/gal)	Modified operating condition (oz/gal)
Nickel sulfate: NiSO ₄ ·6H ₂ O As NiSO ₄	40-50	45.0 26.5	41.0 24.2
Nickel chloride: NiCl ₂ ·6H ₂ O As NiCl ₂	8-12	10.0 5.5	8.5 4.6
Boric acid (H ₃ BO ₃)	6-6.5	6.25	6.1

The plant decided to reduce drag-out losses with a rinse-and-recycle system similar to that in Figure 21. Using the existing two-stage, countercurrent rinse as a single-stage recovery (recycled) rinse followed by a running rinse, drag-out losses could be reduced by 85 percent. This can be seen in Figure 17 based on a recovery rinse ratio of $(5.85 \text{ gal/h}) / (0.9 \text{ gal/h}) = 6.5$. If a third rinse tank was installed, allowing a two-stage recovery rinse before the single-stage final rinse, the recovery system would reclaim 98 percent of the current drag-out losses. The economy of adding the third rinse tank was analyzed.

Table 11 summarizes the results. Case 1 represents the current operating practice. Case 2 represents a single-stage recovery rinse followed by a running rinse using the two existing rinse tanks. Case 3 represents adding a third rinse tank and operating a two-stage recovery rinse followed by a running rinse. The additional \$3,000 investment for a third rinse tank further reduced operating costs by \$4,400 per year (Case 3). Because of this excellent return on investment, a third rinse tank was installed.

In the chrome plating and rinse stations (7 and 8), the plating tank has a surface area of 30 ft^2 (2.8 m^2) and drag-out averages $1.5 \text{ gal}/1,000 \text{ ft}^2$ ($61 \text{ l}/1,000 \text{ m}^2$) of work plated, or 0.9 gal/h (3.4 l/h) drag-out at $600 \text{ ft}^2/\text{h}$ ($56 \text{ m}^2/\text{h}$) plated. This tank is also aerated at a rate of 60 scfm ($1.7 \text{ m}^3/\text{min}$). The plating solution contains 50 oz/gal (375 g/l) chromic acid (H_2CrO_4), has a specific gravity of 1.25, and is maintained at 125°F (52°C). From Figure 18, the surface evaporation rate is 2.4 gal/h (9.1 l/h).

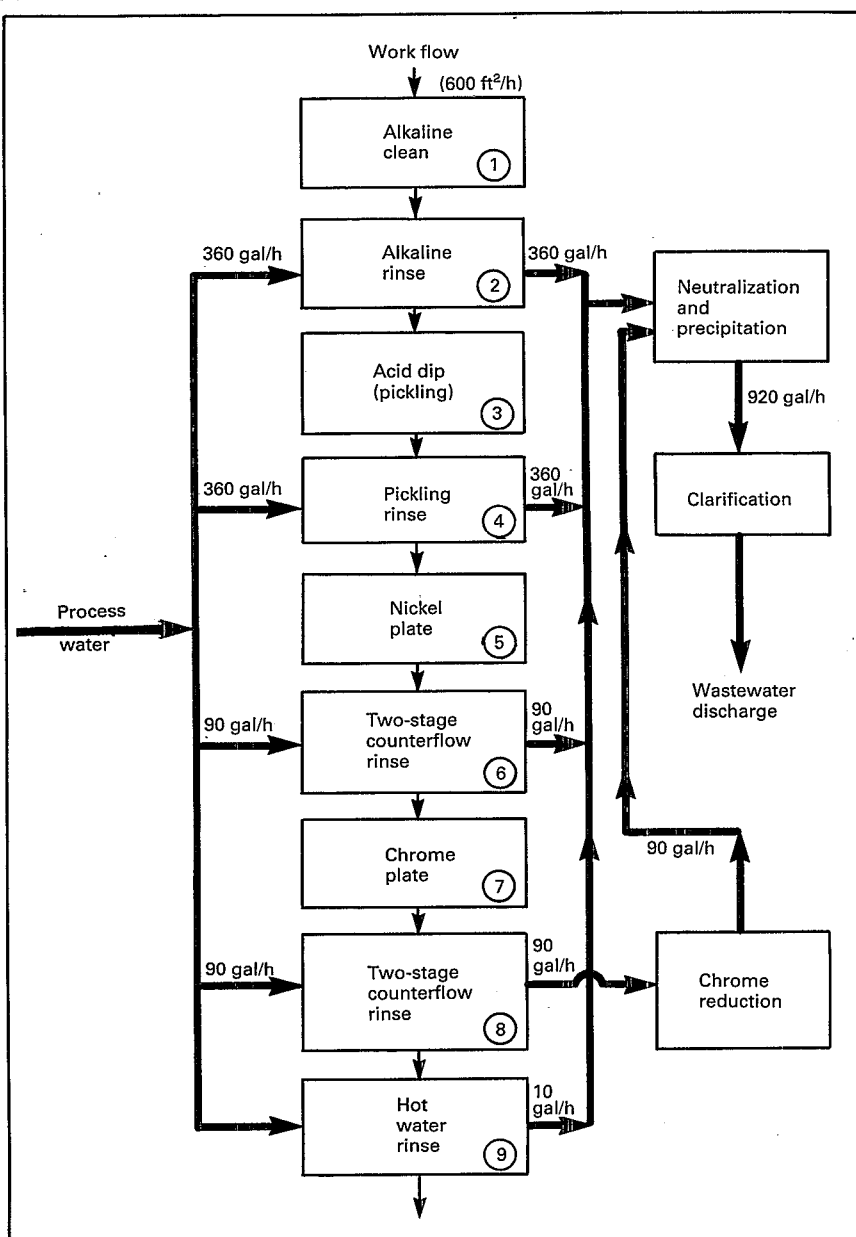


Figure 23. Nickel-Chrome Wastewater Flow Rates: Original System

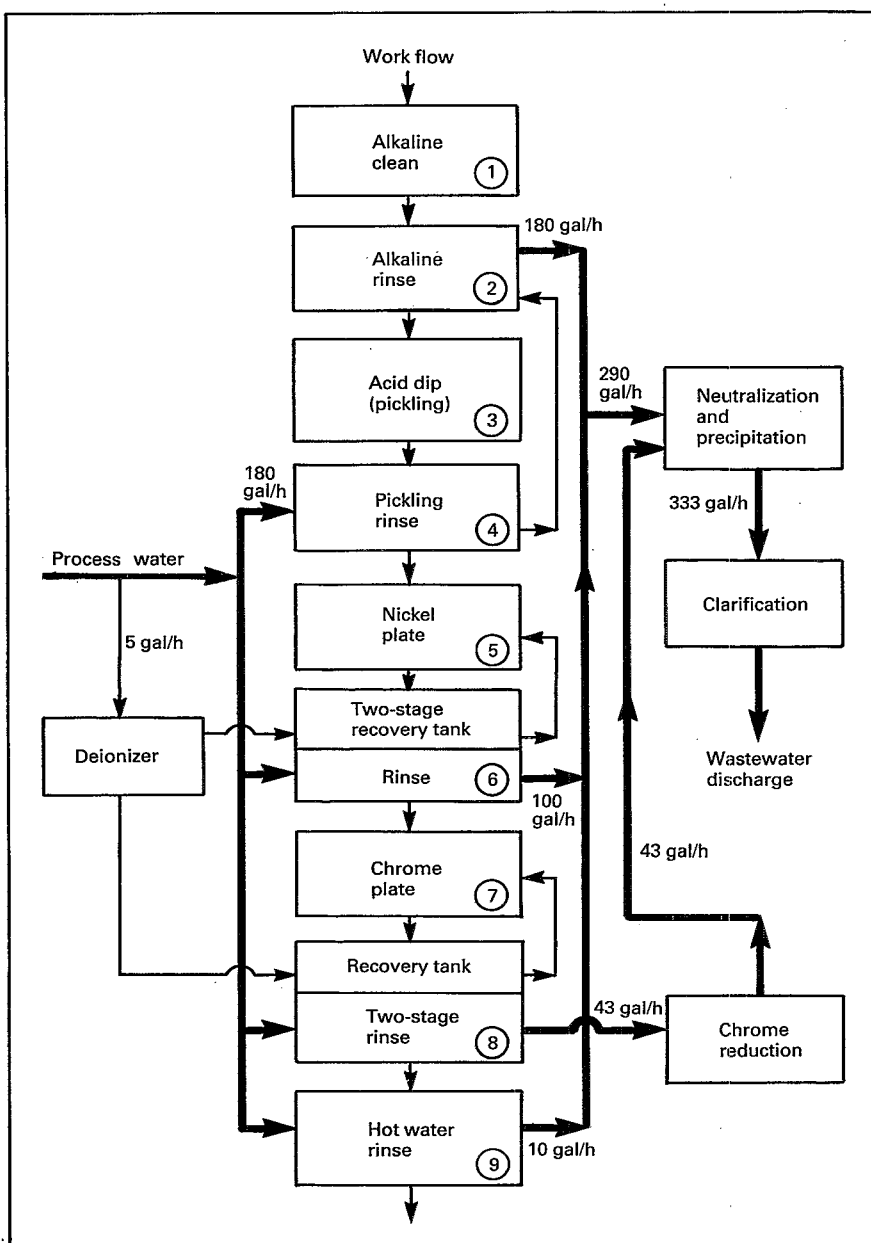


Figure 24. Nickel-Chrome Wastewater Flow Rates: Modified System

A rinse-and-recycle system, as shown in Figure 21, would operate at a rinse ratio of $(2.4 \text{ gal/h}) / (0.9 \text{ gal/h}) = 2.66$. If an additional rinse tank was installed and a two-stage recovery rinse was operated, 90 percent of the drag-out would be recycled (Figure 22). A one-stage recovery rinse could recover 72 percent of the current losses. The plant decided to add a third rinse tank and an analysis was performed to determine whether it would be more advantageous to operate the three rinse tanks as a two-stage recovery rinse followed by a single-stage final rinse or as a single-stage recovery rinse followed by a two-stage final rinse. Table 12 summarizes the results. Case 1 represents the current operating practice. Case 2 represents the two-stage final rinse. Case 3 represents the two-stage recovery rinse.

The two investment options (Cases 2 and 3) entail equal capital costs and reduce operating costs by almost equal amounts. However, Case 3 would result in an almost tenfold increase in wastewater flow to the chromium reduction waste treatment system, from 81 gal/h to 730 gal/h (307 to 2,763 l/h). This volume increase would exceed the capacity of the unit and would reduce the efficiency of downstream waste treatment equipment. When the additional criteria were considered, Case 2 represented the most attractive option and these modifications were incorporated into the operation.

The total cost of the modifications selected was \$15,000. The cost assumes that the plating baths already have purification systems that would control any contaminant buildup resulting from drag-out recycling. The benefits from the modifications include:

- An operating cost reduction of \$8,500/yr for the nickel plating bath.
- An operating cost reduction of \$13,000/yr for the chrome plating bath.
- A reduction in the baseline flow to waste treatment from 920 gal/h to 333 gal/h (3,445 to 1,250 l/h) (see Figures 23 and 24). The total shop discharge rate may now be below 10,000 gal/d (37,854 l/d), putting the plater in an industry category with less stringent treatment regulations under the proposed pretreatment standards.
- A reduced quantity of pollutants discharged in the wastewater effluent in direct proportion to the reduced effluent volume.

Materials Recovery Processes

The high cost of replacing and treating the plating chemicals lost to the waste stream has resulted in the development of chemical separation processes to reclaim these materials for reuse. What all of these processes have in common is that they separate the spent rinse water into a purified stream which is returned to the rinse system, and a stream of plating chemicals concentrated to the point that the solution can be returned to the plating bath.

Table 11.

Evaluation of Rinsing Options for Nickel Plating Operation

	Case 1 (present 2-stage countercurrent rinse)	Case 2 (proposed 1-stage recovery rinse)	Case 3 (proposed 2-stage recovery rinse)
Required modifications	—	Level control rinse feed, repiping	Level control rinse feed, repiping, additional rinse tank
Cost of modifications	—	\$2,000	\$5,000
Drag-out recovery	0	85%	98%
Drag-out losses ^a	\$9,360/yr	\$1,400/yr	\$190/yr
Rinse water required ^b	70 gal/h	735 gal/h	100 gal/h
Water use cost (at \$1.80/1,000 gal)	\$380/yr	\$4,000/yr	\$540/yr
Annual operating cost ^c	\$9,740	\$5,600	\$1,230
Annual savings	—	\$4,140	\$8,510

^aFrom Table 9.

^bFrom Figure 16, Cn = 37 mg/l.

^cDepreciation at 10-year straight line.

Table 12.

Evaluation of Rinsing Options for Chrome Plating Operation

	Case 1 (present 2-stage countercurrent rinse)	Case 2 (proposed 1-stage recovery rinse, 2-stage final rinse)	Case 3 (proposed 2-stage recovery rinse, 1-stage final rinse)
Required modifications	—	Level control rinse feed, conductivity controller, repiping, additional rinse tank	Level control rinse feed, conductivity controller, repiping, additional rinse tank
Cost of modifications	—	\$7,000	\$7,000
Drag-out recovery	0	72%	90%
Drag-out losses ^a	\$18,800/yr	\$5,300/yr	\$1,800/yr
Rinse water required ^b	81 gal/h	43 gal/h	730 gal/h
Water use cost (at \$1.80/1,000 gal)	\$440/yr	\$230/yr	\$3,940/yr
Annual operating cost ^c	\$19,240	\$6,230	\$6,530
Annual savings	—	\$13,010	\$12,720

^aFrom Table 9.

^bFrom Figure 17, Cn = 37 mg/l.

^cDepreciation at 10-year straight line.

Table 13.**Summary of Recovery Technology Applications**

	Plating Bath									
	Decorative Chromium	Hard Chromium	Nickel	Electroless Nickel	Cadmium (CN)	Zinc (CN)	Zinc (Cl)	Copper (CN)	Tin (BF ₄)	Silver
Evaporation	•		•		•	•	•	•	•	•
Electrodialysis			•		•	•	•	•	•	•
Reverse osmosis			•			•		•		
Ion exchange	•		•	•						•
Electrolytic					•	•	•	•	•	•

Recovery processes, which include evaporation, reverse osmosis, ion exchange, electrodialysis and electrolytic processes, should be considered when low-cost plating line modifications or rinse-and-recycle modifications are not available. However, the type of purification system needed in a given situation depends, among other things, on the type of plating chemicals being recovered. Table 13 presents the range of applications for each of the recovery systems.

Each of the recovery processes can be used either as a closed-loop system or as an open-loop system. A closed-loop system is one in which the purified effluent from the recovery unit provides the feed for the final rinse tank (Figure 25a). In this case very little rinse water is sent to wastewater treatment, although purge streams from the recovery unit are almost always necessary.

Where rinse waters require separate waste treatment systems (chromium and cyanide are examples), closing the loop around the plating operation with a recovery system can avert the need for the separate wastewater treatment system. In cases where the high quality of the final rinse is more important than closing the loop, the quality of the final rinse can be ensured by using an open-loop system (Figure 25b). With this approach, the final rinse is set up as a running rinse; its influent is fresh water and its effluent is sent directly to wastewater treatment. The purified effluent from the recovery unit is sent to the next-to-final rinse.

Using a recovery unit requires first reducing the rate of rinse water use to a level that can be processed economically. The use of a multistage countercurrent rinse system is, therefore, critical. Some means of bath purification is also needed to control the buildup of contaminants in the closed-loop system resulting from return of the drag-out to the process bath. In an open-loop system, the drag-out acts as a bleed stream and serves to control the buildup of contaminants.

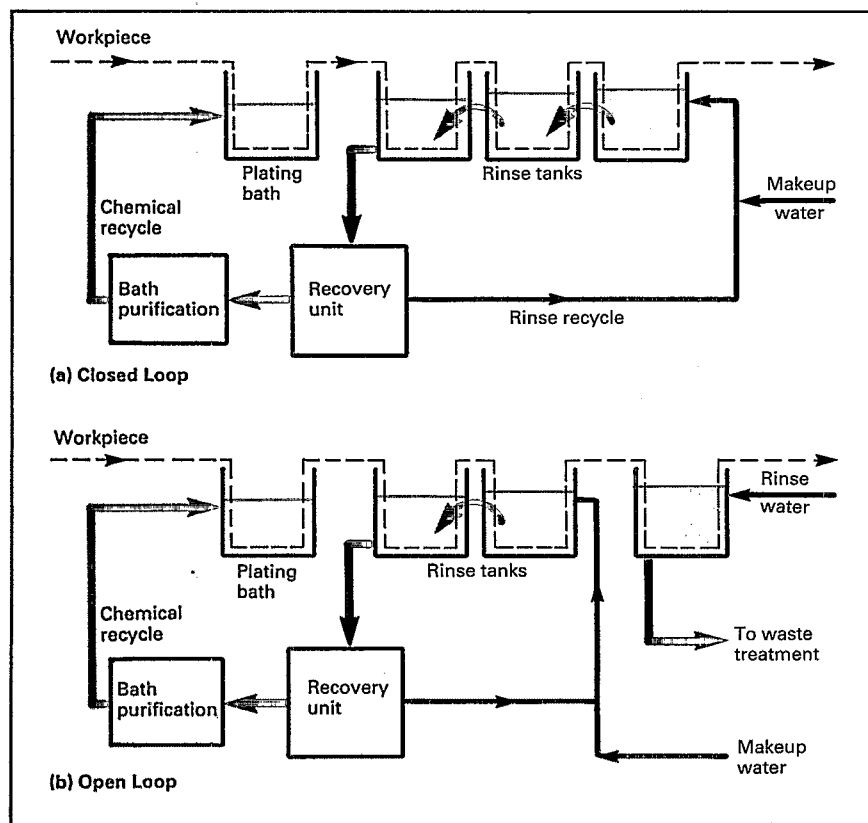


Figure 25. Recovery Systems: (a) Closed Loop; (b) Open Loop

The potential of a recovery system to recover chemicals is shown in Figure 26 as a function of rinse ratio. The curve is the same as that developed for the recovery potential of a two-stage rinse-and-recycle system in Figure 22. In this case, however, the recovery rinse rate is determined by the capacity of the recovery unit, not by the surface evaporation rate of the plating bath.

This chapter examines the operating parameters and costs of the different recovery systems used in the electroplating industry. This information will enable the electroplater, after assessing specific loss factors, to determine the site-specific economics of installing one or more recovery units.

Evaporation. Evaporation was the first separation process used to recover plating chemicals lost to rinse streams. The process has been demonstrated successfully on virtually all types of plating baths, and several hundred units are currently being operated.

Recovery is accomplished by boiling off sufficient water from the collected rinse stream to allow the concentrate to be returned to the plating bath. The condensed vapor is recycled for use as rinse water in the rinse tanks. The boil-off rate, or evaporative duty, is set to maintain the water balance of the plating bath. The evaporation usually is performed under vacuum to prevent thermal degradation of additives in the plating solution.

Figure 27 illustrates a closed-loop evaporative recovery system used on a chromium plating bath with a three-stage, countercurrent rinse system. This system includes a cation exchange column which is necessary in order to prevent the buildup of metallic impurities (primarily dissolved metals from the processed work plus excess trivalent chromium). The losses of plating chemicals from the plating bath have been minimized. Only the amount of chromium plate on the finished product must be replaced in the plating tank. Water consumption is reduced to the water lost to surface evaporation.

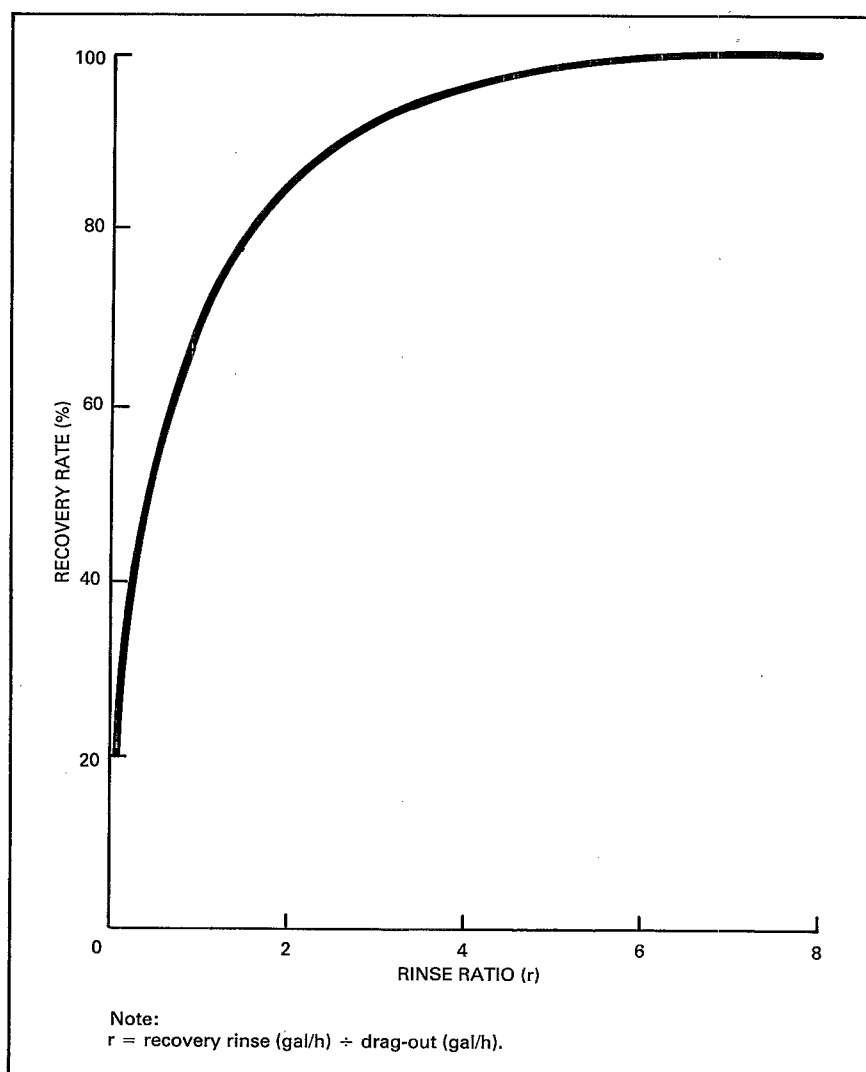


Figure 26. Recovery Potential for a Two-Stage Countercurrent Rinse System

Total installed investment and operating costs for a 20 gal/h (76 l/h) evaporator such as the one shown in Figure 27 are given in Table 14. Of the total annual cost of operation (\$20,100), steam and depreciation are the major portions. The annual savings from the recovery of plating chemicals and the reduction in wastewater treatment costs total \$38,000. The before-tax annual savings for this system is approximately \$18,000 and yields a short payback period of 3.3 years.

The cost of the steam needed and the installed cost for the evaporator depend on the evaporative capacity needed (Figure 28); for rinse recovery systems, this is equal to the required rinse water flow rate. To minimize the rinse rate, such methods as countercurrent rinse systems are usually cost-effective, as discussed in Chapter 3. A 50% savings in the amount of steam needed can be achieved with double-effect evaporators; however, the capital costs are considerably higher and their operation is more complicated. As a rule, at evaporation rates below 150 gal/h, (568 l/h), the additional investment for double-effect evaporators is not justified.

Because of the high initial investment of an evaporative recovery system, the decision to acquire it depends largely on the quantity of plating chemicals available for recovery in the rinse water. For example, if the 20 gal/h (76 l/h) evaporator discussed in Table 14 were fed a stream with 50% less plating chemicals, the annual savings for treatment and recovery would decrease by 50% to approximately \$19,000 ($38,200 \times 0.5$), or slightly less than the annual operating cost of the unit.

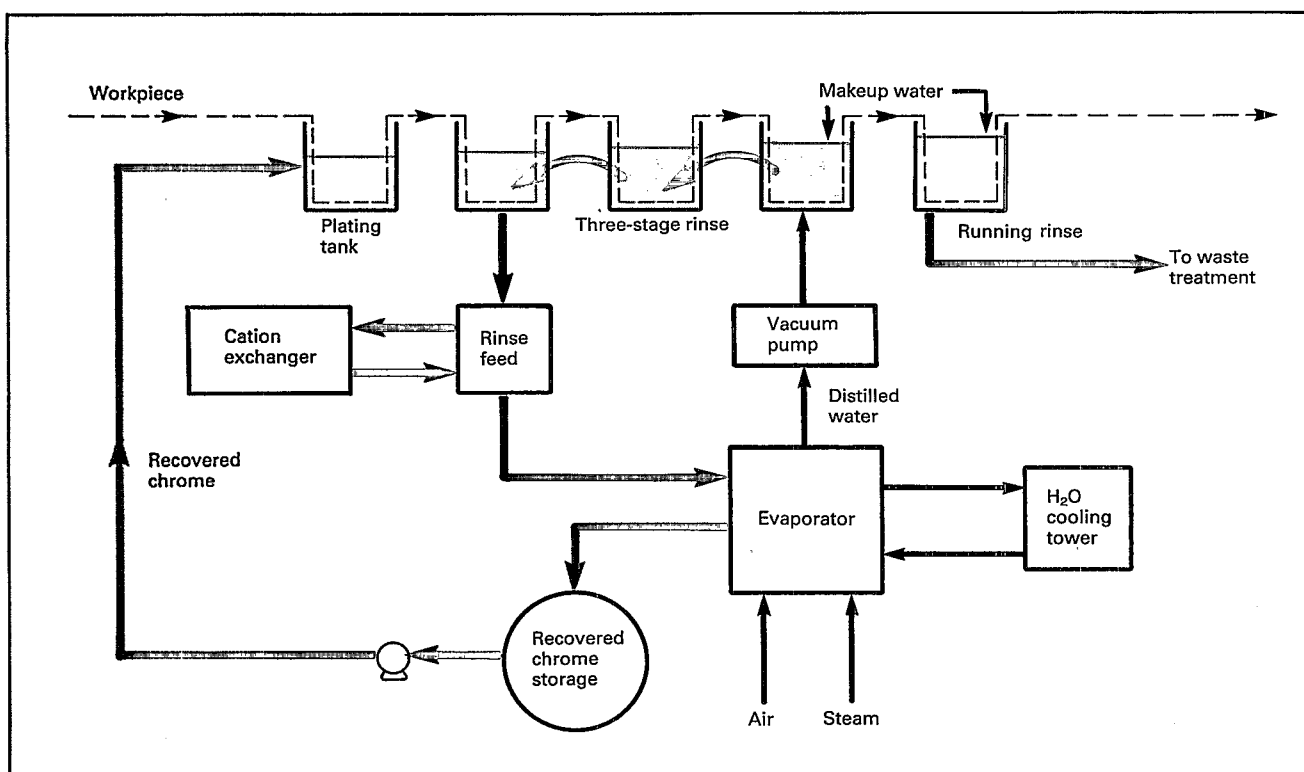


Figure 27. Closed-Loop Evaporative Recovery System

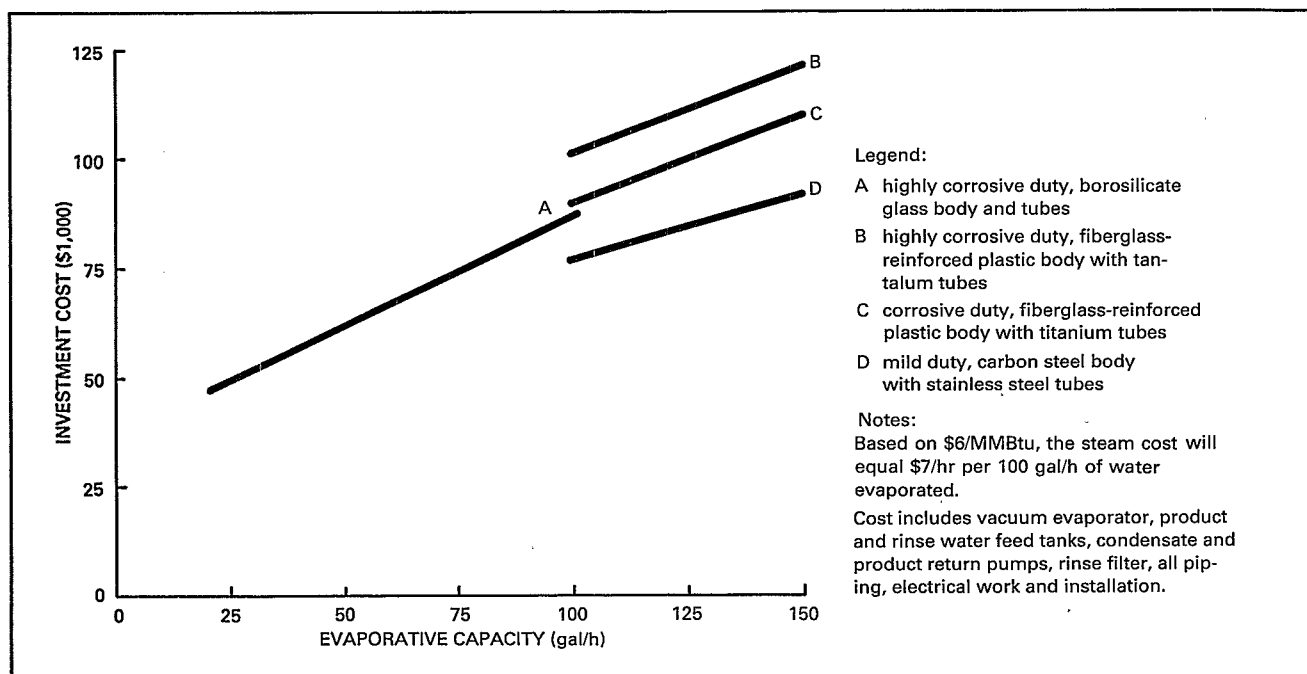


Figure 28. Investment and Energy Costs for Evaporative Recovery Units

Table 14.

Economics of Evaporator System for Chromic Acid Recovery^a

	Cost
INSTALLED COST, 20-gal/h evaporator (\$):	
Equipment:	
Evaporator	24,000
Tanks	2,000
Pumps	1,000
Cation exchanger	9,000
Piping	3,500
Miscellaneous	500
Subtotal	40,000
Installation, labor and materials (\$):	
Site preparation	500
Plumbing	4,000
Electrical	1,500
Equipment erection	500
Miscellaneous	500
Subtotal	7,000
Total installed cost	47,000
ANNUAL COSTS (\$/yr):	
Operating:	
Labor, 100 h/yr at \$10/h	1,000
Supervision	^b
Maintenance, 6% of investment	2,800
General plant overhead	1,200
Raw materials, cation exchanger:	
H ₂ SO ₄ , 3,500 lb/hr at \$0.05/lb	200
NaOH, 3,000 lb/yr at \$0.13/lb	400
Utilities:	
Electricity (\$0.08/kWhr)	2,500
Cooling water, 1,000 gal/h x \$0.10/1,000 gal	500
Steam, \$6.00/MMBtu	6,300
Total annual operating cost	14,900
Fixed:	
Depreciation, 10% of investment	4,700
Taxes and insurance, 1% of investment	500
Total annual fixed cost	5,200
Total annual cost of operation (\$/yr)	20,100
ANNUAL SAVINGS (\$/yr):^c	
Recovered plating chemicals, (3.0 lb/h H ₂ CrO ₄)	17,700
Water treatment chemicals	12,600
Sludge disposal	7,800
Water use, 13 gal/h at 1.80/1,000 gal	100
Total annual savings	38,200
NET SAVINGS = annual savings - (annual operating cost + annual fixed cost) (\$/yr)	18,100
NET SAVINGS AFTER TAXES, 48% tax rate (\$/yr) ^d	9,400
AVERAGE ROI = (net savings after taxes/ total investment) × 100 (%)	20
CASH FLOW FROM INVESTMENT = net savings after taxes + depreciation (\$/yr)	14,100
PAYBACK PERIOD = total investment/ cash flow (yr)	3.3

^aOperating 5,000 h/yr. No interest on capital is included.^bNone required.^cFrom Table 9.^dThe analysis beginning with this line is based on tax law as it existed at the time of the installation.

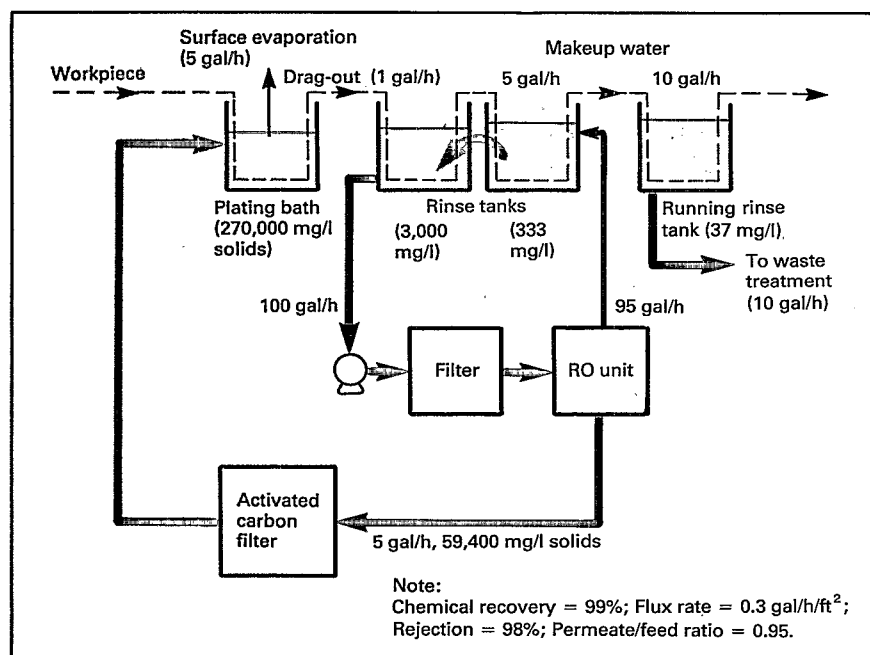


Figure 29. Reverse Osmosis System for Nickel Plating Drag-Out Recovery

Reverse Osmosis. Reverse osmosis (RO) is a pressure-driven membrane separation process. The feed is separated under pressure (400 to 800 psig) through the microscopic pores of a semipermeable membrane into a purified "permeate" stream and a concentrate stream. Commercial RO units have proven successful in concentrating and recycling rinse streams in metal plating operations for a number of years. The primary area of application is in the concentration of rinse waters from mild pH nickel plating baths.

Figure 29 illustrates a typical RO installation for recovering drag-out from an acid nickel plating operation. The system uses a 50 μ m filter to prevent blinding of the membrane by solid particles. The preassembled RO unit consists of a high pressure centrifugal pump and six membrane modules, and installation requires only piping and electrical connections. An activated carbon filter is used to avoid organic contaminant buildup in the plating bath.

The performance of RO units is defined by flux (the rate of passage of purified rinse water through the membrane per unit of surface area) and the percent rejection of a dissolved constituent in the rinse water, which relates to the membrane's ability to restrict that constituent from entering the permeate stream. Percent rejection is defined by:

$$\text{Percent rejection} = \frac{C_f - C_p}{C_f} \times 100$$

where

C_f = concentration in feed stream

C_p = concentration in permeate stream.

The major limitation of commercial RO systems has been their inability to maintain membrane performance, though in recent years two-year warranties from vendors have become common. Fouling and gradual deterioration of membranes can reduce the processing capacity of the unit and necessitate frequent membrane replacements. Currently, feed solutions must be in a pH range between 2.5 and 11 to ensure maximum life of the membrane.

Another limitation of RO is that the membranes are not suitable for treating solutions having a high oxidation potential, such as chromic acid. Furthermore, the membrane does not completely reject certain species, such as nonionized organic wetting agents.

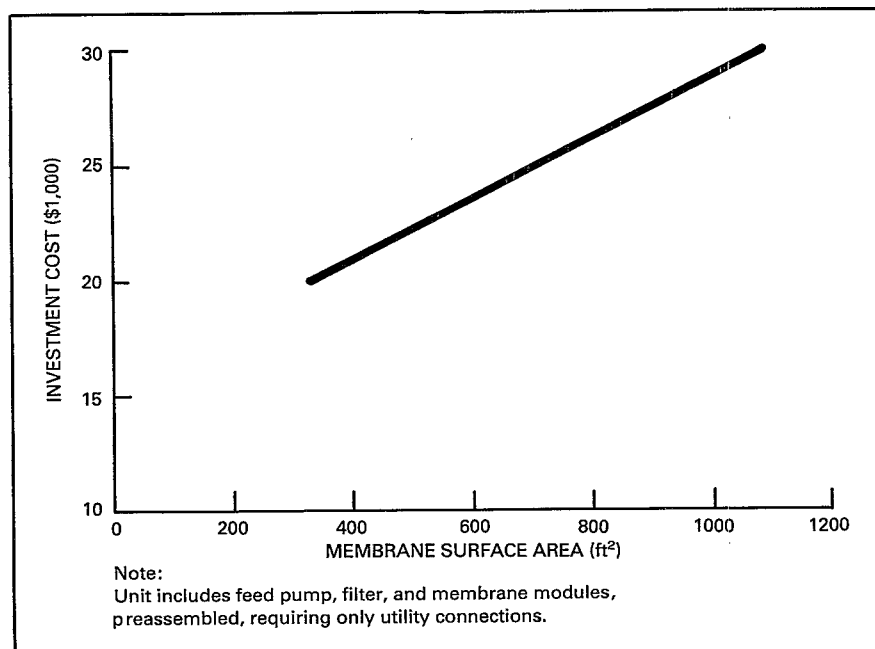
Table 15 lists typical maximum concentrations reached by RO units in commercial applications. Because of these limits, further concentration of the stream by a small evaporator may be required for ambient temperature baths where there is minimal surface evaporation. Therefore, acid nickel plating baths, which experience considerable surface evaporation, are the primary applications for RO.

Table 15.**Reverse Osmosis Operating Parameters**

	Maximum concentration of concentrate stream (%)	Percent rejection
Ni ⁺²	10-20	98-99
Cu ⁺²	10-20	98-99
Cd ⁺²	10-20	96-98
CrO ₄ ⁻²	10-12	90-98 ^a
CN ⁻	4-12	90-95 ^a
Zn ⁺²	10-20	98-99
Low molecular weight organics	<i>b</i>	<i>b</i>

^aPerformance depends greatly on pH of solution.

^bThese compounds are concentrated in permeate stream because of selective passage through membrane.

**Figure 30. Investment Cost for Reverse Osmosis Unit**

These limitations must be weighed when RO applications are considered. Where well chosen, however, RO is an inexpensive, automated process for recovery of plating chemicals.

Figure 30 presents the investment cost of RO units as a function of membrane surface area. A determination of the flux rate for a specific application, and thus of the necessary membrane surface area, usually requires testing by the vendor. As an approximate tool, the flux rate of 0.3 gal/h/ft² (12.2 l/h/m²) (indicated in Figure 29 for a nickel plating bath plating bath with a feed concentration of 3,000 mg/l and a permeate/feed ratio of 0.95) can be used to determine membrane surface area requirements. The flux rate will decrease as the feed concentration increases. Higher permeate/feed ratios also will decrease the flux rate. Experience has shown that doubling the feed concentration or reducing the concentrate volume by 50% (increasing the permeate/feed ratio from 0.85 to 0.975) will decrease the flux rate by 25%. For example, if the RO system shown in Figure 29 was to concentrate the drag-out into 2.5 gal/h (9.5 l/h) of concentrate instead of the 5 gal/h (19 l/h) shown, the membrane surface area requirements would increase from 317 to 422 ft² (29.5 to 39.2 m²).

The cost of the system, shown in Figure 29 and itemized in Table 16, is \$27,500. Theoretically, the system would recover 99% of the plating chemicals lost to the rinse tank. However, Table 16 presents the operating cost reduction that would be achieved if the unit operated 90% of the time. The system has a payback period of 4.6 years.

Table 16.

Economics of Reverse Osmosis System for Nickel Salt Recovery^a

	Cost	
INSTALLED COST, 330 ft² unit (\$):		
Equipment:		
RO module including 50 µm filter, pump and 6 membrane modules at 55 ft ² per module	20,000	
Activated carbon filter	3,000	
Piping	1,000	
Miscellaneous	1,000	
Subtotal		25,000
Installation, labor and material:		
Site preparation	400	
Plumbing	500	
Electrical	700	
Miscellaneous	900	
Subtotal		2,500
Total installed cost		<u>27,500</u>
ANNUAL COSTS (\$/yr):		
Operating:		
Labor, 100 h/yr at \$10/h	1,000	
Supervision ^b		
Maintenance	1,700	
General plant overhead	900	
Raw materials:		
Module replacement, 2-yr life (6 × \$500)(module × 0.5/yr)	1,500	
Resin for carbon filter	700	
Utilities, electricity (\$0.08/kWhr)	900	
Total annual operating cost		6,700
Fixed:		
Depreciation, 10% of investment	2,800	
Taxes and insurance, 1% of investment	300	
Total annual fixed cost		3,100
Total annual cost of operation (\$/yr)		<u>9,800</u>
ANNUAL SAVINGS (\$/yr):^c		
Plating chemicals:		
1.65 lb/h NiSO ₄	8,500	
0.34 lb/h NiCl ₂	1,600	
Water treatment chemicals	2,200	
Sludge disposal cost	3,600	
Water use (no saving)	-	
Total annual savings		<u>15,900</u>
NET SAVINGS = annual savings - (annual operating cost + annual fixed cost) (\$/yr)		
	6,100	
NET SAVINGS AFTER TAXES, 48% tax rate, 6,100 × 0.52 (\$/yr)^d		
	3,200	
AVERAGE ROI = (net savings after taxes/ total investment) × 100 (%)		
	12	
CASH FLOW FROM INVESTMENT = net savings after taxes ; depreciation (\$/yr)		
	6,000	
PAYBACK PERIOD = total investment/ cash flow (yr)		
	4.6	

^aOperating 5,000 h/yr. No interest on capital is included.^bNone required.^cFrom Table 9, based on a 90% operating factor.^dThe analysis beginning with this line is based on tax law as it existed at the time of the installation.

Ion Exchange. In ion exchange, a chemical solution is passed through a resin bed which selectively removes charged particles (ions). Either the positively charged cations (e.g., Cu^{+2} , Fe^{+2}) or the negatively charged anions (e.g., SO_4^{-2} , CN^{-}) are removed from the solution by the exchange of an ion on the surface of a resin particle for a similarly charged ion in the solution. Ion exchange is used in the electroplating industry to remove trace pollutants from wastewater after a conventional treatment process, or to recover plating solution drag-out from rinse water and to return the purified water for reuse.

Unlike other recovery systems whose operating costs are inversely proportional to the chemical concentration, ion exchange is ideally suited for dilute solutions. The treated water is of high purity. As with most recovery systems, however, the capital cost is proportional to flow rate.

A major drawback of ion exchange is that the resin must be regenerated after it has exhausted its exchange capacity. This problem complicates the operation of the system considerably and results in significant volumes of regenerant and wash solutions, which add to the wastewater treatment loading.

Figure 31 shows a fixed-bed ion exchange system used to recover chromic acid from rinse waters. Initially, water passes in series through a cation column and two anion columns. The cation column is

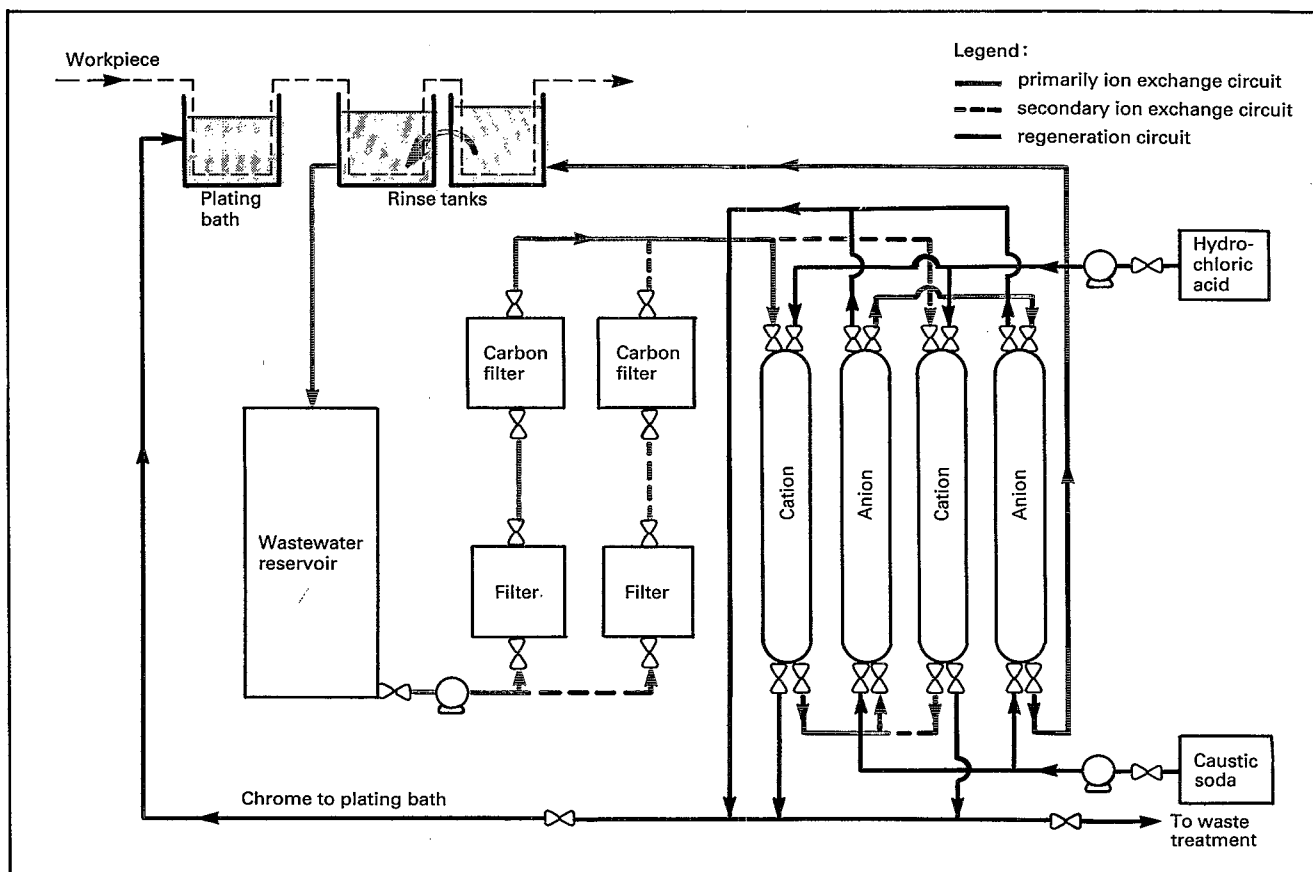


Figure 31. Ion Exchange System for Chromic Acid Recovery

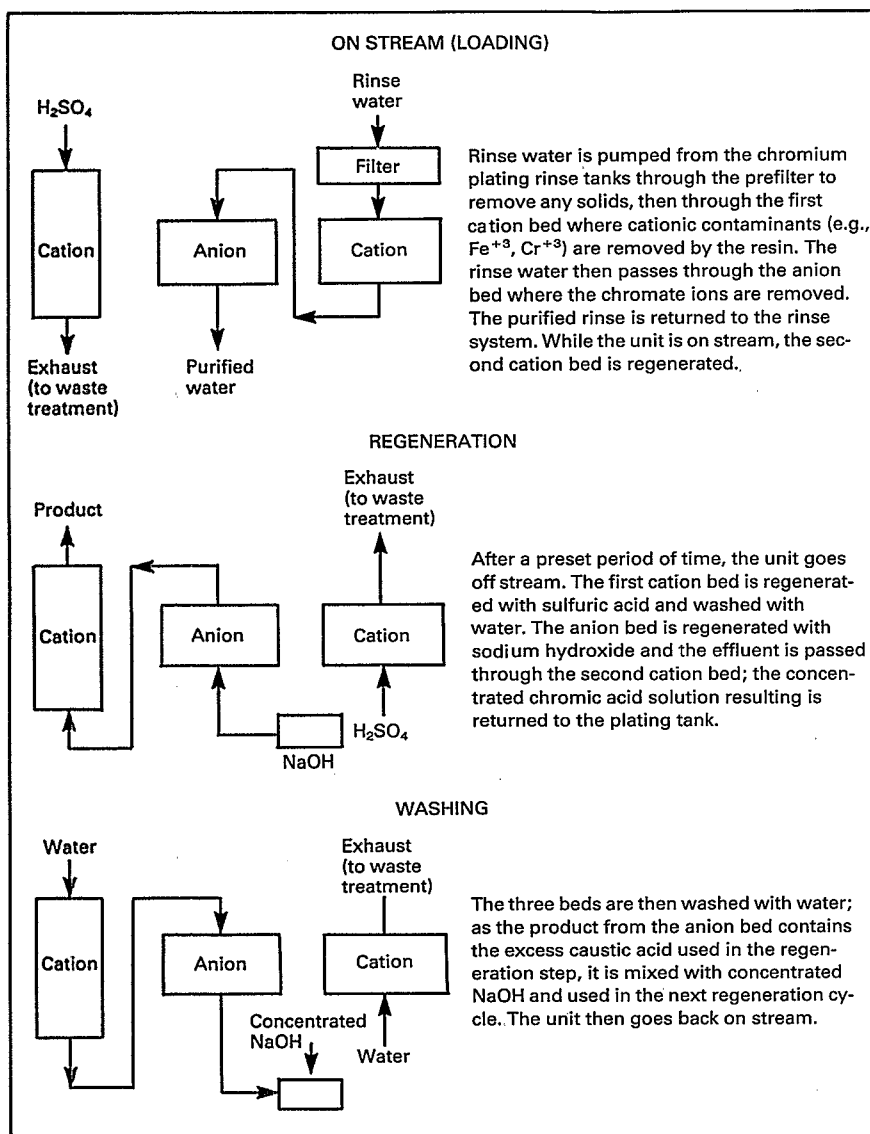


Figure 32. Operating Cycle for Reciprocating Flow Ion Exchange Unit

used to remove any heavy metal contaminants. The anion columns remove the hexavalent chromium from the rinse water. When the upstream anion column has exhausted its exchange capacity, it is taken off stream and regenerated with a caustic solution. This column is then returned to service as the downstream anion column. The product of the anion column regeneration is sodium chromate and any excess caustic used; it is passed directly to a second cation column where the sodium ions are exchanged for hydrogen ions, yielding chromic acid and water. This solution can be returned to the plating bath. The cation columns are regenerated with an acid solution when saturated. The spent regeneration solutions are ordinarily treated by the wastewater treatment plant.

The reciprocating flow ion exchanger (Figure 32) was developed to simplify the operation of units processing large-volume solutions such as the rinse overflow from an electroplater's rinse tank. The unit operates on the principle that during the short period of time the unit goes off stream for regeneration, the buildup of contaminants in the rinse system is negligible. Several reciprocating units with various column configurations are being used to reclaim plating chemicals from rinse streams and to remove metal impurities from various acid baths.

Table 17 presents the total investment and operating costs for and the annual savings realized by installing a reciprocating flow ion exchanger to recover drag-out from a chromium plating operation. The payback period for the initial investment is estimated at 3.6 years.

Table 17.**Economics of Reciprocating Flow Ion Exchange System for Chromic Acid Recovery^a**

	Cost
INSTALLED COST (\$):	
Equipment:	
Reciprocating flow ion exchanger, including cartridge filter and three ion exchange beds	34,000
Piping	1,000
Miscellaneous	1,000
Subtotal	36,000
Installation, labor and material:	
Site preparation	300
Plumbing	500
Electrical	700
Miscellaneous	500
Subtotal	2,000
Total installed cost	38,000
ANNUAL COSTS (\$/yr):	
Operating:	
Labor, 100 h/yr at \$10/h	1,000
Supervision ^b	
Maintenance	2,300
General plant overhead	1,100
Raw materials:	
Replacement resin	1,000
Regeneration chemicals:	
NaOH	1,100
H ₂ SO ₄	1,500
Utilities, compressed air	300
Total annual operating cost	8,300
Fixed:	
Depreciation, 10% of investment	3,800
Taxes and insurance, 1% of investment	400
Total annual fixed cost	4,200
Total annual cost of operation	12,500
ANNUAL SAVINGS (\$/yr):^c	
Plating chemicals, 2 lb/hr H ₂ CrO ₄	11,800
Water treatment chemicals	8,400
Sludge disposal	5,200
Water use, 18 gal/h at \$1.80/1,000 gal	160
Total annual savings	25,600
NET SAVINGS = annual savings - (annual operating cost + annual fixed cost) (\$/yr)	13,100
NET SAVINGS AFTER TAXES, 48% tax rate (\$/yr)^d	6,800
AVERAGE ROI = (net savings after taxes/ total investment) × 100 (%)	18
CASH FLOW FROM INVESTMENT = net savings after taxes + depreciation (\$/yr)	10,600
PAYBACK PERIOD = total investment/ cash flow (yr)	3.6

^aOperating 5,000 h/yr. No interest on capital is included.^bNone required.^cFrom Table 9.^dThe analysis beginning with this line is based on tax law as it existed at the time of the installation.

Electrodialysis. Electrodialysis concentrates ionic species contained in a water solution. The process is well established for purifying brackish water and recently has been demonstrated for recovering metal salts from plating rinse. Compact units suitable for this application have been recovering metal values successfully from rinse streams for a number of years. In addition, a recent EPA demonstration project confirmed the applicability of electrodialysis for plating solutions.

In electrodialysis, a water solution is passed through alternately placed cation-permeable and anion-permeable membranes stacked parallel to the direction of flow (see Figure 33). An electric potential is applied across the stack to induce the ions to migrate across the membranes. The selectivity of the membranes results in the concentration of ions in alternating channels of the stack. Alternating channels are hydraulically linked into two primary hydraulic circuits — one ion depleted, the other ion concentrated. The water flow rate through each of these circuits can be set to achieve the high level of concentration required for returning the plating chemicals to the plating bath. The degree of purification achieved in the ion depleted circuit is set by the electrical potential passed across the membrane. Because ion migration decreases as the diluting circuit increases in purity, a cost optimization is necessary to arrive at the best design.

An EPA demonstration project tested the applicability of an electrodialysis unit to recover nickel from rinse waters for reuse in a plating bath. The system diagrammed in Figure 34 was tested for nine months with no significant operating problems.

Electrodialysis units patterned after the model described are being marketed to reclaim metal values from rinse streams. The units are skid-mounted and require only piping and electrical connections. They cost in the range of \$25,000 to \$30,000.

Table 18 itemizes the cost of the demonstration unit and the operating advantages attributable to the unit. The payback period for the unit, as shown in Table 18, is estimated at 2.2 years.

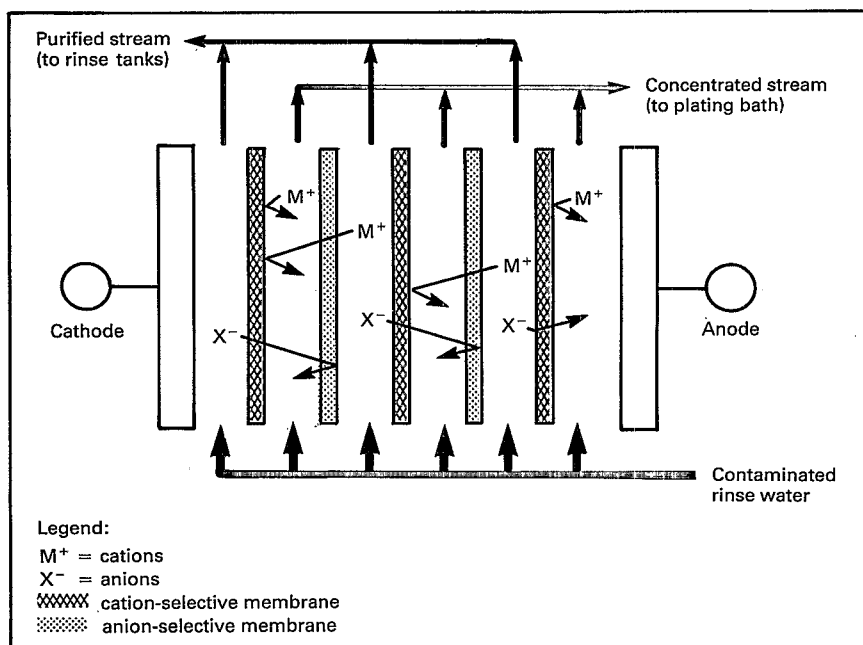


Figure 33. Electrodialysis Flow Schematic

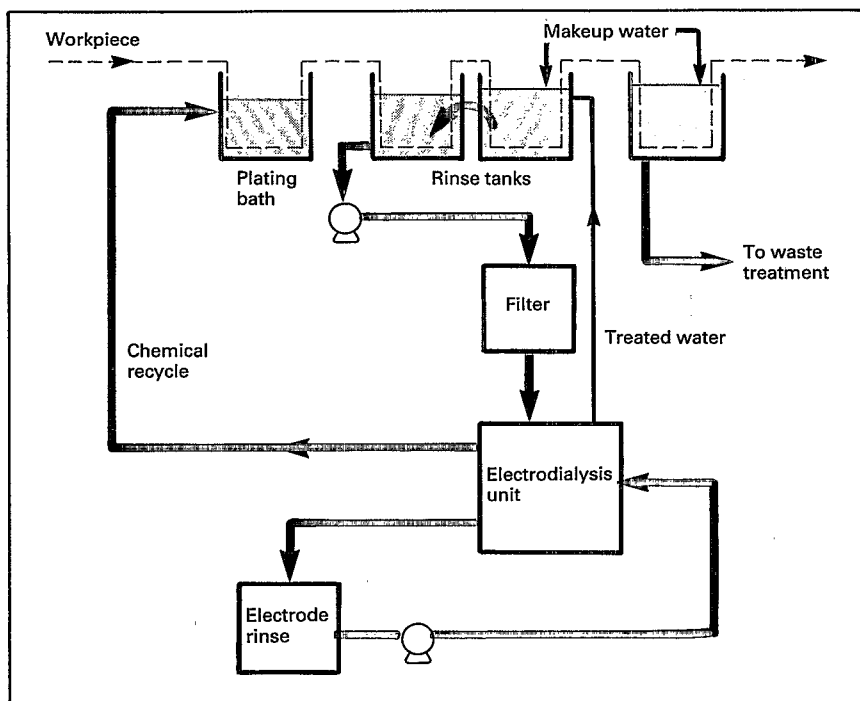


Figure 34. Electrodialysis System for Nickel Plating Drag-Out Recovery

Table 18.

Economics of Electrodialysis System for Nickel Plating Solution Recovery^a

	Cost	
INSTALLED COST (\$):		
Equipment:		
Electrodialysis unit, complete with cartridge filter strainer and electrode rinse system	30,000	
Piping	1,000	
Miscellaneous	1,000	
Subtotal		32,000
Installation, labor and material:		
Site preparation	300	
Plumbing	500	
Electrical	700	
Miscellaneous	500	
Subtotal		2,000
Total installed cost		<u>34,000</u>
ANNUAL COSTS (\$/yr):		
Operating:		
Labor, 100 h/yr at \$10/h	1,000	
Supervisor	^b	
Maintenance	2,000	
General plant overhead	1,000	
Raw materials:		
Filter cartridges	1,000	
Replacement membranes	500	
Utilities, electricity (\$0.08/kWhr)	400	
Total annual operating cost		5,900
Fixed cost:		
Depreciation, 10% of investment	3,400	
Taxes and insurance, 1% of investment	300	
Total annual fixed costs		<u>3,700</u>
Total annual cost of operation (\$/yr)		<u>9,600</u>
ANNUAL SAVINGS (\$/yr):^c		
Plating chemicals, 3.8 lb/h NiSO ₄	22,600	
Water treatment chemicals	3,800	
Sludge disposal cost	6,600	
Water use (no saving)	-	
Total annual savings		<u>33,000</u>
NET SAVINGS = annual savings -		
(annual operating cost + annual fixed		
cost)(\$/yr)	23,400	
NET SAVINGS AFTER TAXES, 48% tax rate (\$/yr)^d	12,200	
AVERAGE ROI = (net savings after taxes/		
total investment) × 100 (%)		
	36	
CASH FLOW FROM INVESTMENT = net savings		
after taxes + depreciation (\$/yr)		
	15,600	
PAYBACK PERIOD = total investment/		
cash flow (yr)		
	2.2	

^aOperating 5,000 h/yr. No interest on capital is included.^bNone required.^cFrom Table 9.^dThe analysis beginning with this line is based on tax law as it existed at the time of the installation.

Electrolytic Processes. The electrolytic cell is the basic device used in electroplating operations. With the advent of innovations in pollution control and interest in material conservation, the electrolytic cell has found other uses in the plating shop:

- Recovery of metals from a plating solution onto a conventional metal cathode (electrowinning)
- Treatment of cyanide plating rinse with simultaneous plating of metals onto a cathode and oxidation of cyanide at the anode
- Oxidation of cyanide contained in waste process solutions (electrolysis).

All electrolytic recovery/treatment systems are predicated upon principles well known to the electroplater. First and foremost are those devices based on an expanded cathode surface area. Many electroplaters have constructed their own units by placing parallel rows of closely spaced (1 to 3 in) anodes and cathodes in a plating tank. The close spacing shortens the path required for metal ions to migrate before meeting a cathodic surface, thus increasing the rate at which they can be removed from the rinses by electrodeposition. Recirculation of the rinse increases deposition rates and improves deposit quality. A modest heat buildup in the recirculating waters, caused by the passage of current through the dilute solution will aid rinsing as well as deposition quality.

New applications of this concept use sponge-like cathodes fabricated from carbon fibers and rigid polymeric foams. While this type of expanded cathode device may be quite efficient, it does not lend itself to direct reuse of the deposited metal. In contrast to the flat cathode units from which the electrodeposited metal can be physically stripped from the starter cathode, the metal deposited on foam cathodes is often sold to a refiner or scrap dealer "as-is" and the foam replaced. Alternatively, chemical and/or reverse current stripping of foam, carbon fiber, and cylindrical cathode can be used to recover the electrodeposited metal for reuse in the plating bath.



Electrolytic recovery unit.

Another method of electrodeposition enhancement involves rapid recirculation of the rinse or rapid movement of the cathode areas. In cylindrical or stacked-disk devices, rotation of the cathodes by means of a motor drive accelerates the exchange of metal ions at the cathode surface. Violently agitated glass beads are also being used for this purpose. Just as in conventional electroplating, this "work agitation" allows the application of higher current densities while preserving deposit integrity. Similarly, high-speed pumped recirculation or vertical-blade agitators in baffled electrowinning tanks can accelerate the deposition rates from dilute rinses by improving the metal ion availability at the cathode surface.

In addition to metal recovery applications, electrolytic processes are also useful for the destruction of cyanides in aqueous solutions. This process is particularly well-suited to the treatment of strong solutions (such as spent plating baths) because their conductivity maximizes the efficient use of electrical current both to electrodeposit metal and to oxidize cyanide at the anode. While conventional electroplating tanks can be used for this purpose, modifying the cells to provide more anode and cathode areas will expedite treatment (reaction times in the range of from 1 to 3 d are not uncommon). The addition of sodium chloride has been reported to accelerate this process, and heating the solution to near-boiling has also been reported helpful.

Standard electrowinning units are available in packages with from 1 ft² to 100 ft² of cathode surface area. Larger custom units are also available. The packaged units usually consist of a reactor tank, copper bussing, cathodes, anodes, recirculation pump, current controller, and rectifier. The installed cost for standard flat plate units as a function of cathode surface area is shown in Figure 35.

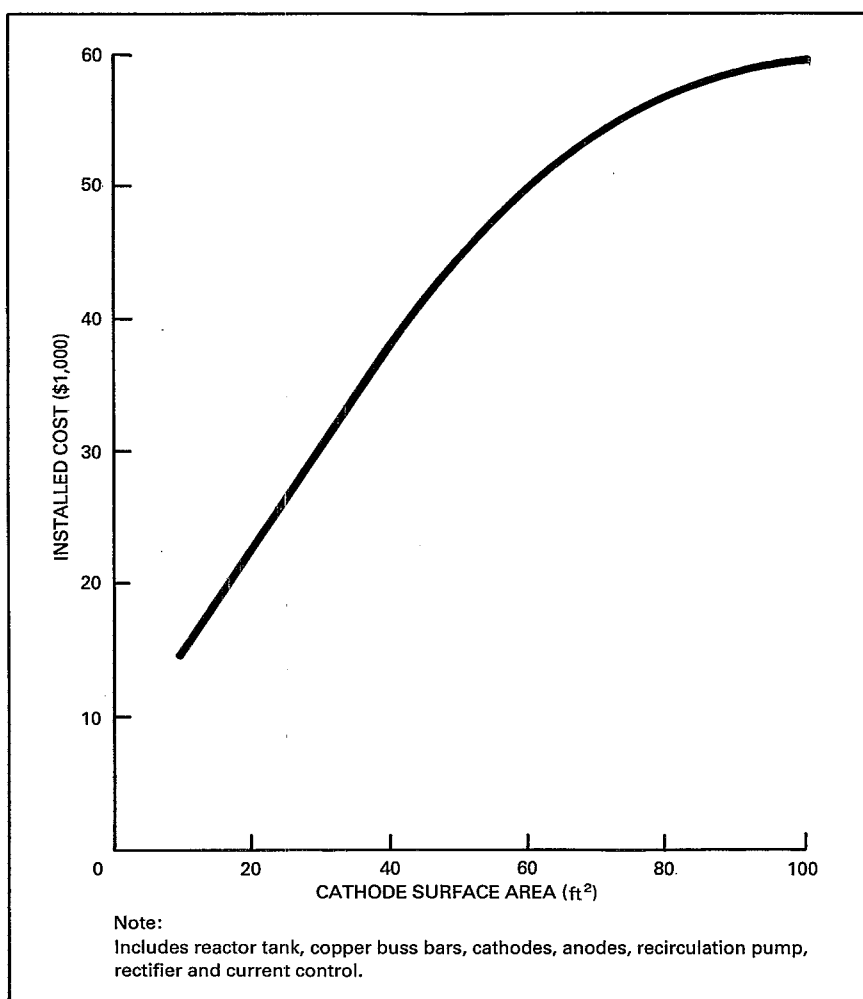


Figure 35. Investment Cost for Electrolytic Recovery Units

The capacity requirement depends on the amount of metal to be recovered. For example, at 10 g/l copper and 140°F temperature, the deposition rate is 0.06 lb/h/ft² at an allowable current density of 25 amp/ft². Drag-out recovery from a plating bath at 90 g/l copper would equal 88% if the recovery rinse were maintained at 10 g/l. If the drag-out rate is 1 lb/h of copper, the required cathode surface area would be (0.88 lb/h)/(0.06 lb/h/ft²), or equal to 15 ft².

Table 19 presents an analysis of an electrolytic copper recovery system in terms of installation and operating costs, economic benefits, and return on investment for the foregoing example. The payback period for this unit is estimated at 3.1 years. The significant amount of copper available for recovery makes the investment in the unit justified. The high surface area electrode systems also come in modular units with from one to four electrode modules per unit. Total installed costs for these units range from \$49,000 for one module to \$90,000 for a four-module unit.

Table 19.

Economics of Electrolytic Copper Recovery System^a

	Cost
INSTALLED COST, 330 ft² unit (\$):	
Equipment:	
Electrolytic unit including heater and recirculation pump	21,000
Piping	1,000
Miscellaneous	1,000
Subtotal	23,000
Installation, labor and material:	
Site preparation	300
Plumbing	500
Electrical	700
Miscellaneous	500
Subtotal	2,000
Total installed cost	25,000
ANNUAL COSTS (\$/yr):	
Operating:	
Labor, 100 h/yr at \$10/h	1,000
Supervisor ^b	
Maintenance	1,500
General plant overhead	900
Utilities, electricity (\$0.08/kWhr)	1,000
Total annual operating cost	4,400
Fixed cost:	
Depreciation, 10% of investment	2,500
Taxes and insurance, 1% of investment	300
Total annual fixed costs	2,800
Total annual cost of operation (\$/yr)	7,200
ANNUAL SAVINGS (\$/yr):^c	
Plating chemicals (1.0 lb/h Cu or 2.5 lb/h CuSO ₄)	11,000
Water treatment chemicals	2,500
Sludge disposal cost	4,500
Water use (no saving)	—
Total annual savings	18,000
NET SAVINGS = annual savings - (annual operating cost + annual fixed cost) (\$/yr)	
	10,800
NET SAVINGS AFTER TAXES, 48% tax rate, 10,800 × 0.52 (\$/yr)^d	
	5,600
AVERAGE ROI = (net savings after taxes/total investment) × 100 (%)	
	22
CASH FLOW FROM INVESTMENT = net savings after taxes + depreciation (\$/yr)	
	8,100
PAYBACK PERIOD = total investment/cash flow (yr)	
	3.1

^aOperating 5,000 h/yr. No interest on capital is included.^bNone required.^cFrom Table 9.^dThe analysis beginning with this line is based on tax law as it existed at the time of the installation.

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Appendix A. Drag-Out Recovery Cost Reduction Worksheet

This worksheet is intended to lead the user through the analysis required to determine the potential cost reduction achievable by recovering plating solution drag-out.

Figure A-1 illustrates the procedure; Figure A-2 is a blank worksheet. The annual operating cost (C.14) represents only the cost of raw material losses, bath heating, pollution control, and waste disposal for a plating line. A comparison of the operating cost associated with several different investment options will indicate the relative economy of the different options.

Other items to consider in a complete cost comparison include the labor and investment associated with the modifications. Table 6 of this report shows a procedure for determining the total annual cost of a waste treatment system. The same procedure can be used to determine total annual cost for drag-out recovery options.

Figure A-1.

Worksheet Completion Procedure

A. Plant conditions to be measured

1. Drag-out rate (gal/h)	1.5 gal/h
2. Plating tank temperature (°F)	150°F
3. Plating tank air agitation rate (stdft ³ /min)	0
4. Plating tank surface area (ft ²)	36 ft ²
5. Number of rinse tanks	4
6. Dissolved solids concentration in plating tank (mg/l)	260,000 mg/l
7. Dissolved solids concentration in final rinse (mg/l)	50 mg/l
8. Plating solution composition	$\text{NiSO}_4 = 1.66 \text{ lb/gal}$ $\text{NiCl}_2 = 0.34 \text{ lb/gal}$ $\text{Boric acid} = 0.40 \text{ lb/gal}$

B. Cost factors for each plant

1. Plating solution value (\$/gal)	$\text{NiSO}_4 (1.66 \text{ lb/gal} \times \$1.74/\text{lb})^a = \$2.89$ $\text{NiCl}_2 (0.34 \text{ lb/gal} \times \$1.76/\text{lb})^a = \$0.60$ $\text{Boric acid} (1.40 \text{ lb/gal} \times \$0.30/\text{lb})^b = \$0.42/\text{gal}$ $\text{Total} = \$3.91/\text{gal}$
2. Water use and sewer charges	\$1.80/1000 gal
3. Energy cost for plating tank heaters	\$6.00/10 ⁶ Btu
4. Annual operating hours	3,600

Figure A-1.**Worksheet Completion Procedure—Concluded****C. Operating cost estimation for rinse and recycle system (Figure 21)**

1. Surface evaporation rate (gal/h) ^c (0.14 gal/h-ft ²) × 36 ft ²	5.04
2. Recycle rinse ratio (surface evaporation rate/drag-out rate) or (C.1/A.1)	3.36
3. Number of countercurrent rinse tanks in recovery rinse	2
4. Percent recovery of drag-out (from Figure 22)	94
5. Concentration of drag-out from recovery rinse (mg/l) $C_r = 0.06 C_p$ (from Figure 22)	15,600
6. Number of countercurrent rinse tanks in final rinse	2
7. Dilution ratio in final rinse $\frac{C_r}{C_i} = \frac{15,600}{50}$	312
8. Rinse ratio in final rinse (from Figure 17)	18
9. Rinse water requirements in final rinse; gal/h (rinse ratio × drag-out rate) or (C.8 × A.1)	27
10. Drag-out chemical losses (\$/h) (drag-out rate) × (plating solution value) × (100 - percent recovery/100) or (A.1 × B.1) × [100 - (C.4/100)]	\$0.34/h
11. Rinse water use cost (\$/h) (water use rate × cost factor) or (C.9 × B.2)	\$0.049/h
12. Bath heating load due to surface evaporation, Btu/h; (5.04 gal/h) × (8,300 Btu/gal) ^d	41,800 Btu/h
13. Heating load cost (\$/h) (C.12 × B.3)	\$0.25/h
14. Annual operating cost (\$/yr) (C.10 + C.11 + C.13) × B.4	\$2,300/yr

^aFrom Table 9.^bFrom Table 1.^cFrom Figure 19. For aerated baths use Figure 18.^dFrom page 33. For aerated baths use Figure 20.

Figure A-2.

Cost Reduction Worksheet

A. Plant conditions to be measured

1. Drag-out rate (gal/h)	
2. Plating tank temperature (°F)	
3. Plating tank air agitation rate (stdft ³ /min)	
4. Plating tank surface area (ft ²)	
5. Number of rinse tanks	
6. Dissolved solids concentration in plating tank (mg/l)	
7. Dissolved solids concentration in final rinse (mg/l)	
8. Plating solution composition	

B. Cost factors for each plant

1. Plating solution value (\$/gal)	
2. Water use and sewer charges	
3. Energy cost for plating tank heaters	
4. Annual operating hours	

Figure A-2.**Cost Reduction Worksheet—Concluded****C. Operating cost estimation for rinse and recycle system (Figure 21)**

1. Surface evaporation rate (gal/h) _____	
2. Recycle rinse ratio (surface evaporation rate/drag-out rate) or (C.1/A.1)	
3. Number of countercurrent rinse tanks in recovery rinse	
4. Percent recovery of drag-out (from Figure 22)	
5. Concentration of drag-out from recovery rinse (mg/l) _____ (from Figure 22)	
6. Number of countercurrent rinse tanks in final rinse	
7. Dilution ratio in final rinse $\frac{C_r}{C_f}$ or (C.5/A.7)	
8. Rinse ratio in final rinse (from Figure 17)	
9. Rinse water requirements in final rinse; [gal/h (rinse ratio \times drag-out rate) or (C.8 \times A.1)]	
10. Drag-out chemical losses (\$/h) (drag-out rate) \times (plating solution value) \times (100 - percent recovery/100) or (A.1 \times B.1) \times [100 - (C.4/100)]	
11. Rinse water use cost (\$/h) (water use rate \times cost factor) or (C.9 \times B.2)	
12. Bath heating load due to surface evaporation (Btu/h) _____	
13. Heating load cost (\$/h) (C.12 \times B.3)	
14. Annual operating cost (\$/yr) (C.10 + C.11 + C.13) \times B.4	

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