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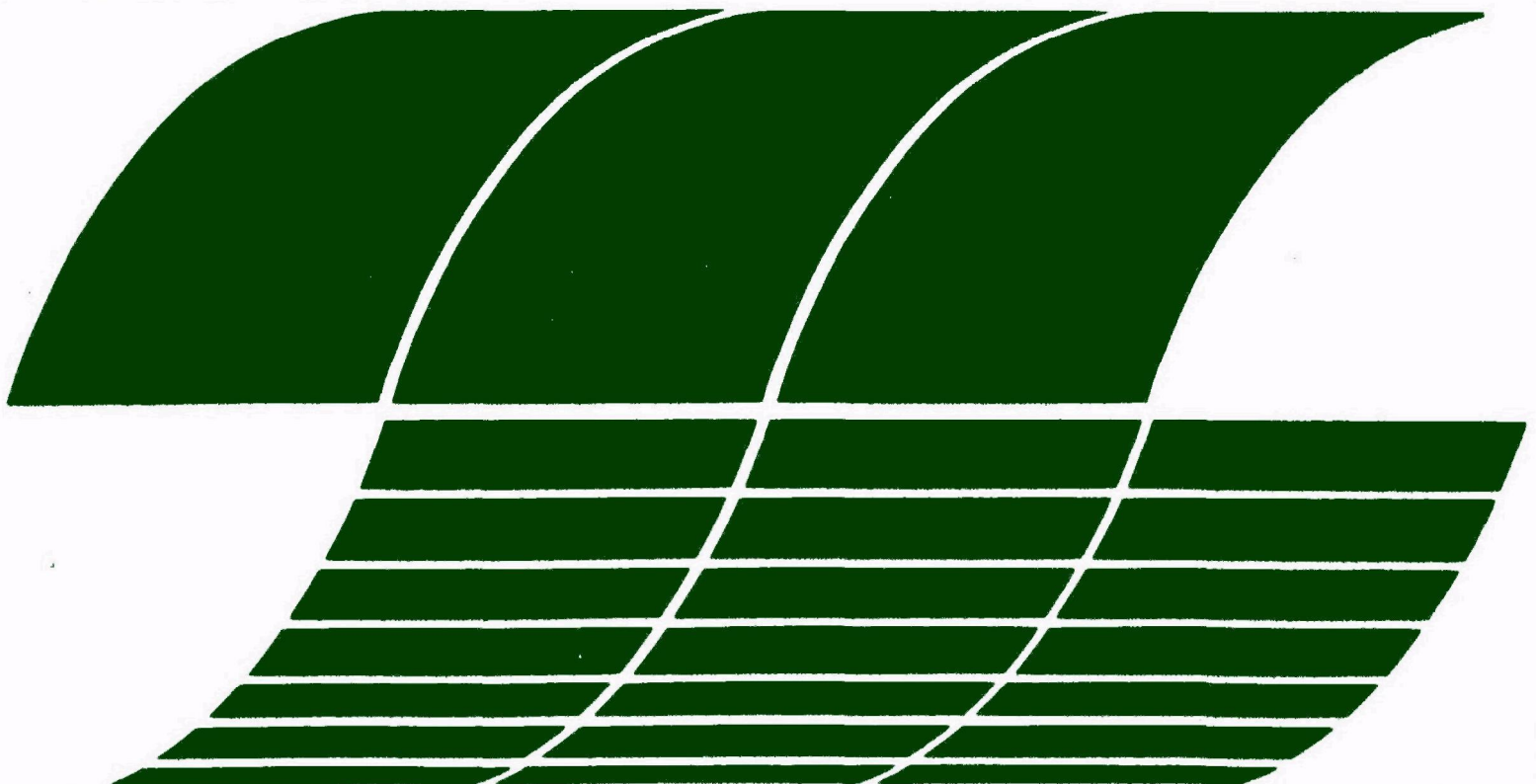
Preliminary Cost Estimates of Pollution Control Technologies for Geothermal Developments

Interagency Energy/Environment R&D Program Report

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PRELIMINARY COST ESTIMATES OF POLLUTION CONTROL
TECHNOLOGIES FOR GEOTHERMAL DEVELOPMENTS

by

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FOREWORD

When energy and material resources are extracted, processed, converted, and used, the related pollutional impacts on our environment and even on our health often require that new and increasingly more efficient pollution control methods be used. The Industrial Environmental Research Laboratory - Cincinnati (IERL-Ci) assists in developing and demonstrating new and improved methodologies that will meet these needs both efficiently and economically.

This report provides preliminary pollution control cost estimates for developers and regulators of geothermal energy. The report and similar ensuing reports are intended to develop the technical basis for eventual regulations.

Further information on the subjects of this report can be obtained from the Power Technology and Conservation Branch, Industrial Environmental Research Laboratory, Cincinnati, Ohio 45268.

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ABSTRACT

The utilization of geothermal resources for electrical power generation may contribute to energy production in the near future. A substantial capital investment will be required to control air and water pollutant emissions from geothermal power plants. This study is a preliminary investigation of the costs incurred in controlling H_2S emissions and treating waste fluids with a variety of applicable control technologies. Estimates include capital and operational/maintenance costs.

Air pollution control cost estimates for H_2S abatement utilizing the Stretford, EIC, Dow oxygenation, and iron catalyst processes have been developed. These process control technologies are in various stages of development, ranging from laboratory testing of a pilot unit to operating field installations. The location of a H_2S abatement unit in the power production process is dictated by the specific control technology used. Condenser ejector gases are controlled utilizing the Stretford process; the EIC process scrubs geothermal steam upstream of the power plant; geothermal brine is treated by the Dow oxygenation process; and the iron catalyst process is applied to geothermal steam condensation equipment (direct contact condenser and cooling tower water). The cost is 2.1 mills per KWH for the Stretford process and is 1.25 mills per KWH for the Iron Catalyst process under the following conditions: H_2S concentration of 220 ppm, steam flow of 907,000 kg/hr, pressure of 7.8 atm, and a temperature of 180°C. The cost for the EIC process is 3.6 mills per KWH at 830 ppm H_2S, 150°C, 11.9 atm and 71,000 kg/hr of steam. Depending on the type of mixers used, the cost for the Dow Oxygenation process is 9.2 mills per KWH for in-line mixers and is 8.6 mills per KWH for concurrent packed tower under the following operating conditions: 500 ppm H_2S, double flash conversion system, brine temperature of 177°C and pressure and brine flow of 11.2 atm and 100,000 lpm respectively. Due to the variability of application of the control technologies and the "site-specific" data base, it is difficult to make a conclusive comparison of H_2S control technology costs with the information presently available.

Cost estimates for water treatment technologies were developed based on: brine flow rates, raw geothermal brine concentrations, and discharge brine concentrations. Sedimentation, chemical precipitation and filtration process costs were generated for preliminary treatment of geothermal brines. Cost estimates for additional treatment including reverse osmosis, electrodialysis, ion exchange, and evaporation processes for the reuse of treated brine were developed. Injection, ocean disposal, evaporation ponds, and land application (utilized for brine disposal) costs were estimated. Costs for treatment and disposal of sludge generated by brine treatment technologies were also determined.

The type and cost of brine treatment technologies required for geothermal energy conversion processes are dependent upon the concentration of constituents and the degree to which these constituents must be removed. Subsurface injection of geothermal brines appears to be the most economical and technically feasible alternative. Minimal treatment is necessary for subsurface injection; sedimentation may be sufficient for low or moderate salinity brines; high salinity brine requires additional treatment.

Treatment of brine for disposal by sedimentation is significantly less costly than technologies required for brine reuse. For a geothermal wastewater flow of 120,000 liters per minute, the cost of treatment is most economical for sedimentation (at 0.4 cents/1000 liters) and is most expensive for reverse osmosis (at 12.9 cents/1000 liters).

Existing discharge regulations result in prohibitive treatment costs for ocean disposal of geothermal brines. Treatment of geothermal brines with evaporative systems (multiple stage evaporators or compression stills) is not economically attractive or technically feasible. Exorbitant costs result for sludge disposal if brines are treated sufficiently for reuse purposes.

As stated previously, the cost estimates presented in this report are preliminary and should not be construed as firm estimates. Further investigation and study are required to develop more accurate costs information. The costs for brine treatment processes were derived from data based principally on municipal wastewater treatment systems. The use of these data to develop costs for geothermal -rine treatment systems requires additional investigation to validate their technological and economic applications.

Treatability studies to demonstrate the performance of air and water control technologies evaluated in this report are recommended over a range of operating conditions expected to exist at geothermal sites. Research and development of additional control technologies should be encouraged to continue. For example, with additional investigation, the Deuterium H_2S removal process may prove to be economically and technically attractive.

The long-term feasibility of subsurface injection of brine should be determined. If long-term injection is not practical, the alternatives (drilling additional wells to continue subsurface injection or more efficient brine treatment) may incur additional pollution control costs.

Removal of boron (frequently existing in geothermal brines) has not been demonstrated to be technically feasible at the present time. Recent research and development studies utilizing specific adsorbents or foam separation have shown significant promise.

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SECTION 1

INTRODUCTION

The development of geothermal resources as an alternative energy resource
is not without environmental concern. Extraction and processing of geothermal
fluids can result in undesirable air emissions, toxic pollutant discharges
and contamination of surface and subsurface waters, potential land subsidence,
seismic activity, noise pollution, and possible blowouts of producing wells.
The uncertainties of these potential environmental problems have led to vari-
ous studies undertaken by the geothermal industries, the Department of Energy
(DOE) and academia (through federal and industrial funding) to develop control
measures.

The Environmental Protection Agency (EPA) has taken an initial step
towards the establishment of regulatory standards for the geothermal industry
by preparing a document entitled "Pollution Control Guidance for Geothermal
Energy Development". This report supports that document by providing pollu-
tion control cost information.

The objective of this report is to provide preliminary cost estimates for
air and water pollution treatment and disposal technologies applicable for geo-
thermal energy conversion systems. Cost estimates include both annualized
capital investment and operation and maintenance (O&M) costs for various levels
of environmental requirements.

SECTION 2

CONCLUSIONS

This study is an effort undertaken by TRW to provide preliminary cost estimates for applicable air and water pollution control systems. The associated costs for the handling and disposal of solid wastes were also evaluated. The culmination of the study produced the following conclusions.

AIR POLLUTION CONTROL

- The control technologies available for hydrogen sulfide (principal air pollutant) abatement are in various stages of technical development, ranging from field installations to preliminary design concepts. At present, the Stretford, EIC and Dow Oxygenation processes appear to be the most feasible control technologies.

- The cost of H_2S abatement at a power generation rate of 10,000 KW is 2.1 mills per KWH for the Stretford process and is 1.25 mills per KWH for the Iron Catalyst process under the following conditions: H_2S concentration of 220 ppm, steam flow of 907,000 kg/hr, pressure of 7.8 atm, and a temperature of 180°C. The cost for the EIC process is 3.6 mills per KWH at 830 ppm H_2S, 150°C, 11.9 atm and 71,000 kg/hr of steam and a power generation rate of 10,000 KW. Depending on the type of mixers used, the cost for the Dow Oxygenation process is 9.2 mills per KWH for in-line mixers and is 8.6 mills per KWH for cocurrent packed tower under the following operating conditions: 500 ppm H_2S, double flash conversion system, brine temperature of 177°C and pressure and brine flow of 11.2 atm and 100,000 lpm respectively. Comparisons of cost estimates should not be made without consideration of these baseline differences.

- The primary sources of hydrogen sulfide emissions from geothermal electric power generation processes are cooling towers and condenser ejectors. Hydrogen sulfide dissolved in the cooling water and condensate steam can be removed by the iron cataoyst process. Efficiency of H_2S removal by this process is still under investigation.

- Hydrogen sulfide emissions from condenser ejector gases can be effectively controlled by utilizing the Stretford process. However, the Stretford process for H_2S abatement requires the use of a surface condenser rather than the conventional direct contact condenser.

- Dow Oxygenation process can remove H_2S from unflashed geothermal brine and is applicable only to liquid-dominated resources.

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WATER POLLUTION CONTROL

● Wastewater treatment technologies required for geothermal conversion processes are highly dependent on the concentrations of constituents present and the quantities of pollutants in the waste stream that must be removed.

● Because of the prevailing environmental discharge regulations, disposal of high salinity geothermal wastewaters by direct ocean discharge or land application will probably not be generally adequate without costly treatment.

● Subsurface injection appears to be economically and technically feasible for the disposal of excess wastewater and concentrated brine. Minimum treatment of the geothermal wastewater is required by this method. For low and moderate salinity geothermal fluids, sedimentation may be the only treatment necessary prior to its disposal. High salinity geothermal fluids may require additional treatment.

● Costs for chemical precipitation and filtration are substantially less than costs for treatment processes such as reverse osmosis, ion exchange or electrodialysis. For a geothermal wastewater flow of 120,000 liters per minute, the cost of treatment is most economical for sedimentation (at 0.4 cents/1000 liters) and is most expensive for reverse osmosis (at 12.9 cents/1000 liters). The cost for subsurface injection at the same wastewater flow is estimated at 1.3 cents/1000 liters of water injected.

● The use of evaporative systems (multiple stage evaporators or compression stills) for complete treatment of geothermal fluids is both economically unattractive (more than 10 times as costly as ion exchange process) and technically infeasible because of corrosion and scaling problems. The direct cost for disposal of geothermal fluids by ocean disposal is likely to be prohibitive even when environmental regulatory requirements are ignored, because the distance from the ocean, and large quantities of water compared to the economic value of energy produced.

● In arid regions, where fresh water supplies are at a premium, treatment of a portion of the spent geothermal fluid for reuse may be a viable alternative to complete disposal by subsurface injection.

● The wastewater control technologies discussed in this report are applicable for low and medium salinity geothermal fluids. The economics as well as the applicability of these treatment systems for removing pollutants from high salinity geothermal fluids need further investigation. In particular, the technology for boron removal (from geothermal fluids) is still under research and development; as such there is no proven technology for effective removal of boron at present.

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SOLID WASTE DISPOSAL

- Sludge generated from the various wastewater treatment systems is highly dependent on the flow rate and the nature of the geothermal brine to be treated.

- Sludge disposal cost can be prohibitive for the treatment of high level waste for either disposal or reuse purposes. If subsurface injection is to be used for the disposal of spent geothermal fluids, the cost of sludge disposal can be substantially reduced by incorporating treatment processes which can minimize sludge production such as acidification and/or chelation of the wastewater.

SECTION 3

RECOMMENDATIONS

Based on the findings and conclusions of this report, the following recommendations are made:

- The pollution control cost estimates presented in this report should be viewed only as preliminary. Since most of the data were derived from municipal and industrial applications other than geothermal developments, cost data should be updated and revised as necessary when they become available.

- Air pollution cost estimates presented in this report for the Stretford, EIC, Dow Oxygenation and Iron Catalyst processes in H_2S control were based on specific plant operating conditons. In order to prepare cost estimates for other geothermal manifestations with varied operating conditions, site-specific detailed laboratory or pilot plant data should be developed.

- Research and development of additional H_2S emission control technologies should continue. H_2S is highly reactive making it especially amenable to process research.

- Additional study should include an evaluation of the technical and economic feasibility of combining individual air pollution control technologies to abate H_2S emissions from geothermal conversion processes. As an example, Dow Oxygenation and EIC processes might be applied to the brine and steam respectivley in a flash energy conversion system.

- For air emission control from geothermal developments, further investigation should include both technical and economic evaluations of abatement of other air pollutants such as methane, ammonia and carbon monoxide.

- The effectiveness and economics of the water pollution control systems were based primarily on data derived from treatment of wastewater significantly different from that expected in a geothermal development. Additional programs should, therefore, include treatability studies of the geothermal waste fluid including low, medium and high salinity waters by using the control technologies discussed in this report, and other innovative technologies.

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- The applicability of reverse osmosis as a candidate geothermal wastewater treatment system for achieving the various effluent quality requirements needs to be demonstrated.

- Although effective boron removal has not been demonstrated in commercial scale operations, recent research and development studies by specific adsorbents and foam separation processes have shown significant promise. These programs should be encouraged to include both technical and economic feasibility evaluations.

- Sludge disposal appears to be a major cost constraint in waste brine treatment for compliance with potential effluent quality requirements, particularly for high salinity geothermal waters. Additional studies should focus on technologies minimizing or eliminating sludge production, such as acidification, metal chelation, etc.

- Subsurface injection appears to be the best alternative for waste brine disposal. To minimize the cost for injection, it is recommended that pumping tests be performed at potential geothermal sites to determine the minimum wastewater quality requirements for injection.

- The cost of pollution control processes depends greatly on the quality of the wastewaters and steam. Additional research is needed to characterize the geothermal fluid resources. This information may permit the assessment of the cost of control equipment by specific geographic region.

- Detailed studies are needed to better understand subsurface injection cost as a function of the various cost determinants (e.g., drilling depth, lithology, etc., which would affect drilling costs).

- The cost and feasibility of pollution control equipment is dependent greatly on the energy conversion efficiency of the plant and the utilizable energy in the geothermal resources. Data are needed to relate plant efficiency, geothermal resource quality, and pollution control cost per unit of energy generation.

- Before definitive effluent discharge requirements for the geothermal industry are adopted, it is recommended that demonstration studies be conducted to validate the effectiveness and economics of the control technologies presented in this report.

SECTION 4

POTENTIAL GEOTHERMAL CONVERSION PROCESSES AND ASSOCIATED WASTE STREAMS

GEOTHERMAL ENERGY CONVERSION SYSTEMS

Geothermal resources may exist as steam (vapor-dominated resources), but the major geothermal resources to be developed in the United States will most likely exist as hot water (liquid-dominated resources). Air and water pollutant emission sources from geothermal developments are dependent upon the resource type and the energy conversion process used. In addition to direct heating, five types of power generation are under development: (1) direct steam turbine, (2) flash steam turbine, (3) flash steam binary cycle, (4) hot water binary cycle, and (5) total flow turbine systems (Cheremisinoff, 1976, and Library of Congress, 1974).

Direct Steam Turbine Power Generation

Electrical power is produced from vapor-dominated geothermal resources by expanding the steam through a turbine coupled to a generator. Turbines are designed to operate on relatively low steam temperatures and pressures compared to those utilized for conventional fuel-fired power generation. A centrifugal separator, upstream of the turbine, removes particulate matter. A barometric contact condenser or a surface condenser is generally used to condense the steam from the turbine at sub-atmospheric pressure to increase turbine efficiency. The condenser is equipped with an ejector to remove the noncondensible gases. The condenser fluid, condensate plus cooling water, may be pumped to a forced-draft cooling tower, and then back to the condenser to cool incoming steam. Excess cooling water may be subsurface-injected or discharged to the surface. A surface condenser or dry cooling tower can be substituted for the respective equipment described above. A schematic diagram of the direct process is shown in Figure 1.

Flashed Steam Turbine Power Generation

Steam for power generation from liquid-dominated geothermal resources is obtained by partial flashing of the liquid to a lower pressure. The flash chamber also acts as a centrifugal separator to remove liquid and particulates from the steam. The remaining brine can be: flashed again if its temperature is sufficiently high; subsurface-injected; or discharged on the ground surface. The separated steam is expanded through a turbine coupled to a generator. A barometric condenser and cooling tower may be utilized in the same manner as described for the previous system. A surface condenser may not be appropriate because of a two-phase flow. This condition may create problems for

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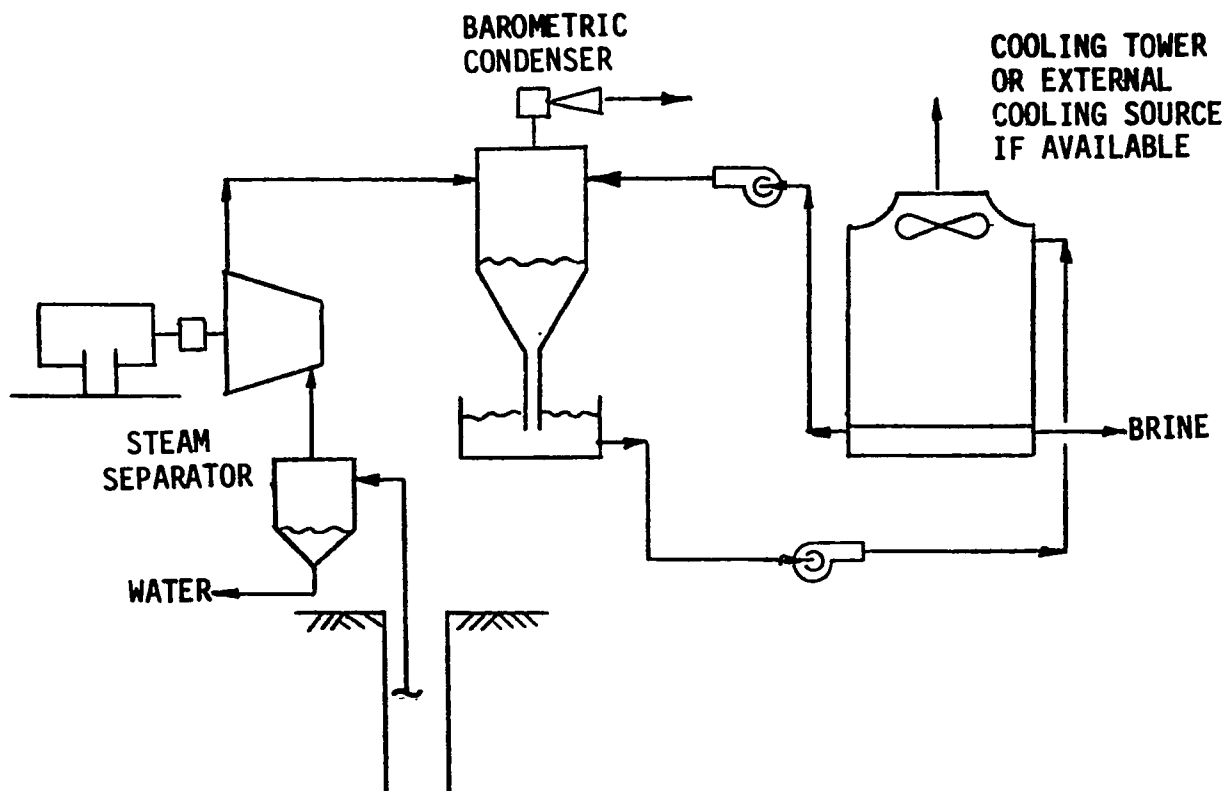
Figure 1. Direct steam process.

the condenser and the pump because of the non-condensible gases. A flashed steam system is depicted in Figure 2.

Binary Cycle - (Flashed steam to heat a secondary working fluid)

Steam obtained by flashing geothermal liquid, is passed through heat exchangers (boiler and superheater) to vaporize a low boiling point secondary fluid, such as isobutane. The high-pressure secondary fluid vapor is expanded through a turbine coupled to a generator. The secondary fluid vapor exhausted from the turbine is condensed and pumped back to the heat exchangers at a high pressure. The steam used in the heat exchangers is condensed and the noncondensible gases removed. Brine from the flash separator and steam condenser is injected or discharged above ground. The steam condensate may be passed through a cooling tower and recycled to provide condenser coolant for both the working fluid and flashed steam. A binary cycle flashed steam system is shown in Figure 3.

Binary Cycle - (Hot water to heat a secondary working fluid)

Hot geothermal water is used directly to vaporize a secondary fluid by circulating both countercurrently through a boiler and superheater. The high pressure secondary fluid vapor is expanded through a turbine coupled to

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BAROMETRIC
CONDENSER

COOLING
TOWER

FLASH CHAMBER
AND SEPARATOR

BRINE

BRINE

DOWN HOLE
PUMP

<u>NOTE:</u> SEPARATOR DISCHARGE BRINE
CAN BE FLASHED IF ITS
TEMPERATURE IS SUFFICIENTLY
HIGH

Figure 2.   Flashed steam process.

WORKING FLUID VAPOR

CONDENSER

COOLING
TOWER

WORKING
FLUID
LIQUID

SUPER-
HEATER

BOILER

BRINE

STEAM

STEAM

BAROMETRIC
CONDENSER

FLASH CHAMBER
AND SEPARATOR

BRINE

DOWN HOLE
PUMP
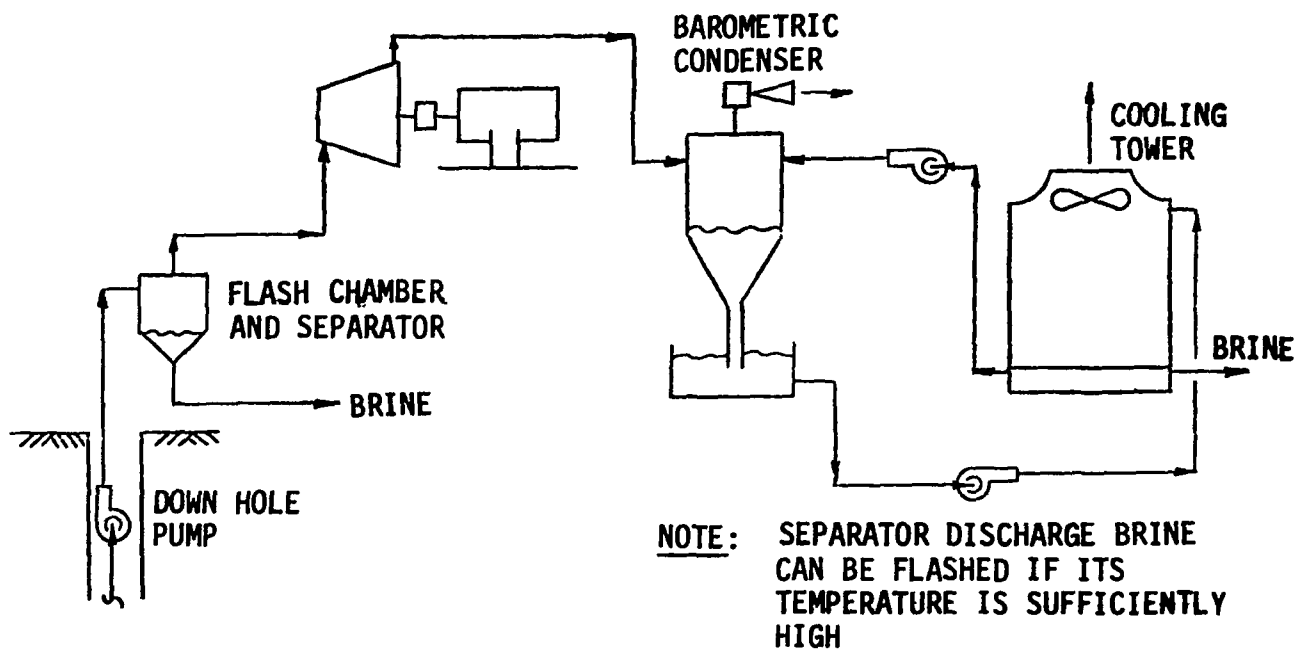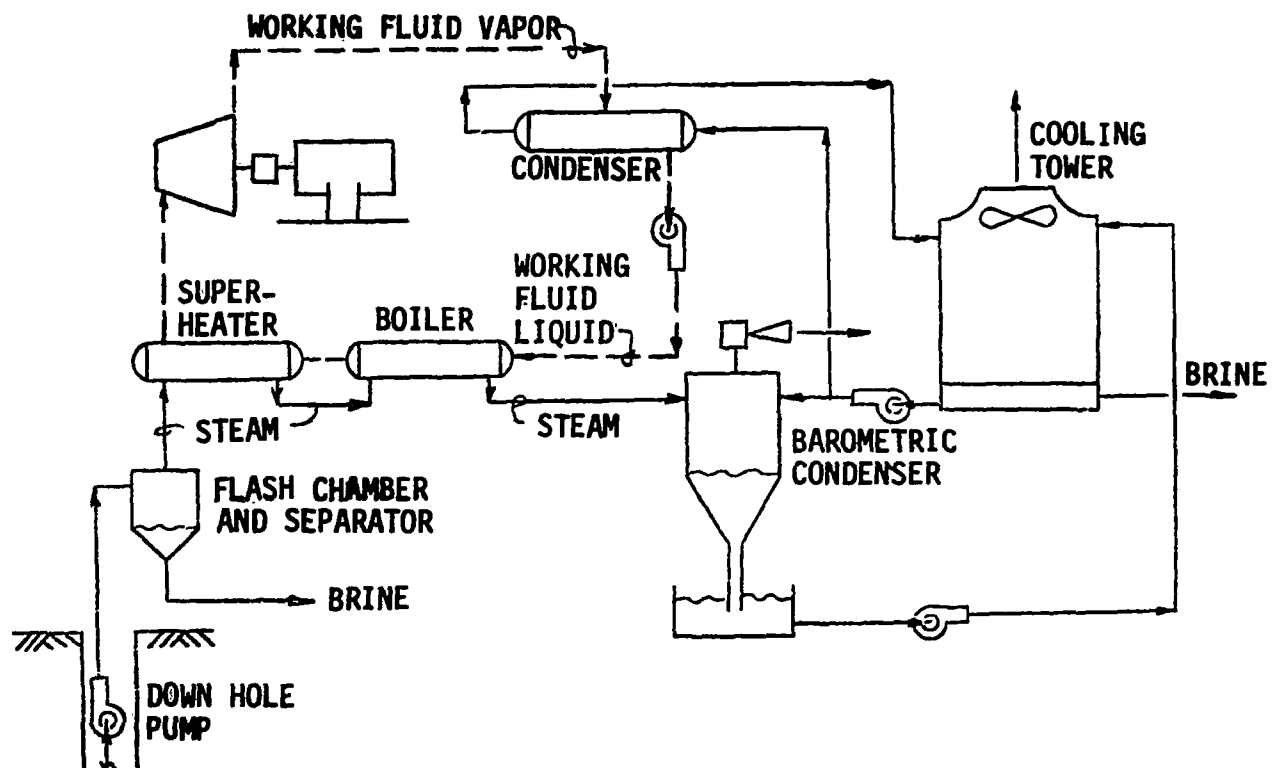
Figure 3.   Binary cycle (flashed steam) process.

a generator. The secondary fluid vapor exhausted from the turbine is condensed with coolant from a cooling tower, and pumped back to the heat exchangers at a high pressure. The spent geothermal water is injected or discharged on the surface. This system differs from those described previously in that cooling water must be supplied from an outside source. A flow diagram of the binary cycle hot water system is illustrated in Figure 4.

Total Flow Turbine Power Generation

A total flow geothermal energy conversion system utilizes an over-pressurized hot water resource. The geothermal fluid is allowed to expand as it ascends to the surface. The fluid then passes through a pressure-reducing nozzle to increase its velocity. The kinetic energy of the fluid drives an impulse turbine coupled to an electric generator. The impulse turbine requires special design and materials of construction to minimize erosion and corrosion caused by direct contact with the geothermal fluid. The turbine discharge fluid may be injected or discharged above ground. Coolant is not required in the system; thus there is no cooling water discharge. A schematic diagram of this total flow system is shown in Figure 5.

Direct Heating

Low to moderate temperature (<90°C) geothermal resources, not suitable for power generation, may be used in a variety of direct heating applications. Such applications include space heating, industrial process heat, crop drying, soil warming, etc. In most cases the heat is extracted by heat exchangers and the spent fluid is injected or discharged on the surface. An example process for direct heating is shown in Figure 6.
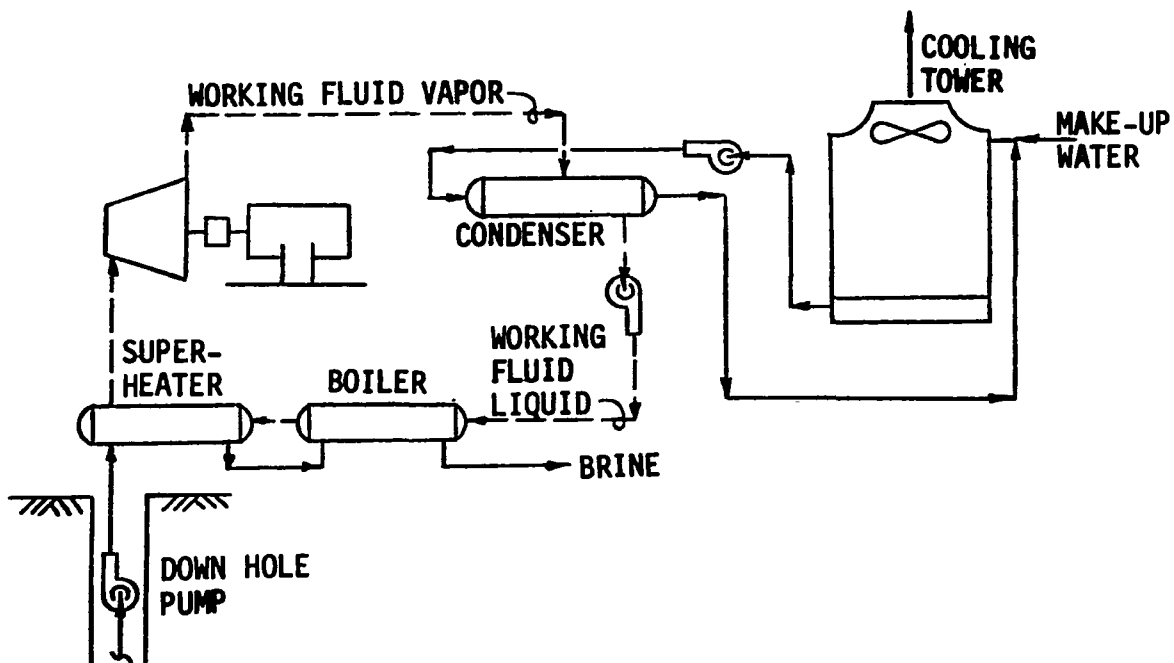
Figure 4.   Binary cycle (hot water) process.
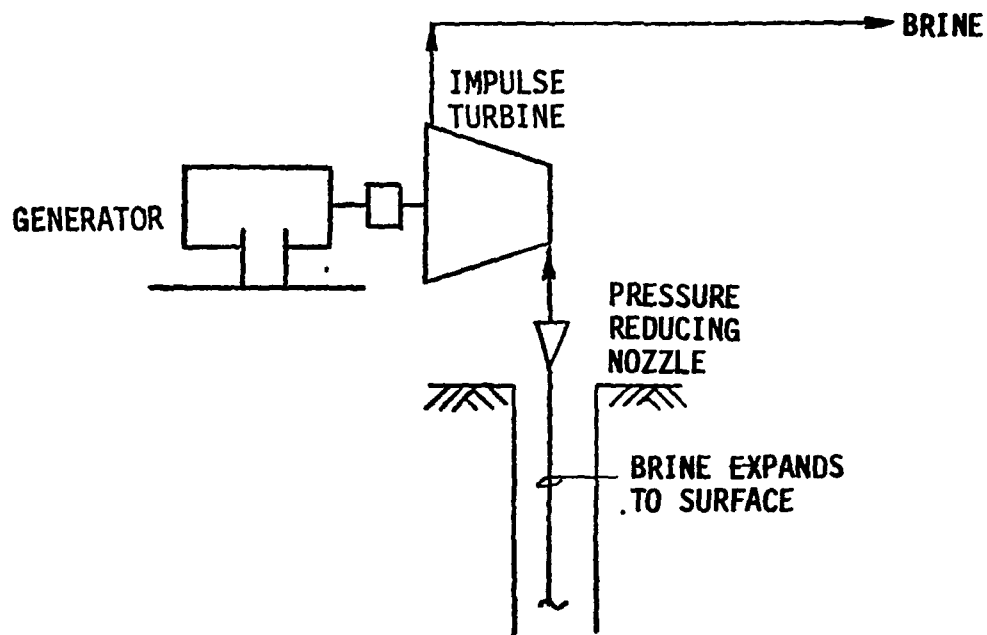
Figure 5.   Total flow process.
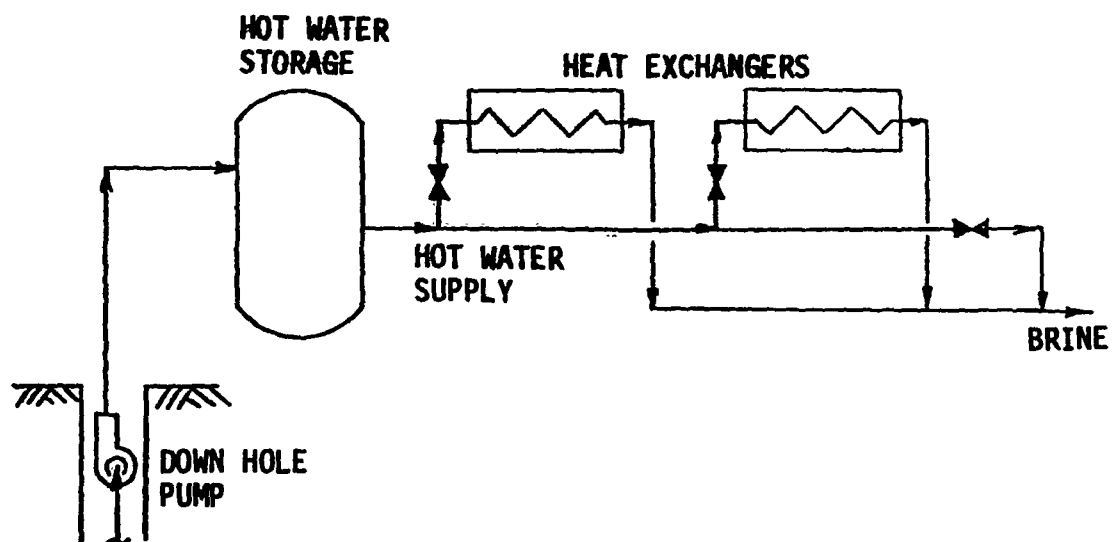
Figure 6.   Direct heating (closed system).

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Low-temperature geothermal resources can be used for agricultural land application, providing the salt content is suitable for plant life. Purposes may include frost prevention, plant nutrition and simple irrigation. The heat and carbon dioxide from the geothermal resource can be contained in a greenhouse to increase plant growth or crop yield throughout the year.

Desalination - (For mineral and/or water recovery)

Due to the high temperatures of many geothermal resources, they may contain high concentrations of valuable minerals. If economically profitable, mineral recovery alone could justify geothermal resource development. Mineral recovery may, in some cases, reduce the cost of a wastewater treatment process for power production. Various desalination processes can produce fresh water suitable for secondary uses. The Bureau of Reclamation has several programs under investigation at the East Mesa plant, California, for desalting geothermal fluids to supplement the Colorado River. Evaporation, distillation, reverse osmosis, electrodialysis, etc. are applicable desalination processes.

IDENTIFICATION OF AIR POLLUTANT EMISSION SOURCES

For geothermal energy conversion systems the major sources of air emissions are from operations relating to:

- well drilling
- well cleanout
- pipeline venting
- power plant by-pass
- condenser ejector
- cooling tower

Air pollutants will be emitted from energy conversion systems operating on geothermal steam; they will not be emitted, however, from conversion systems that do not condense steam (e.g., binary cycle using hot geothermal brine to heat a secondary working fluid). The principal air pollutants include noncondensible gases and particulate material such as metals. Noncondensible gases (those that do not condense at normal operating temperatures) in geothermal steam vary in concentration from one resource to another and include the following major constituents: hydrogen sulfide, carbon dioxide, methane, ammonia, nitrogen and hydrogen. Hydrogen sulfide, which exists in essentially all geothermal steam, is the most likely to cause an environmental hazard due to its toxicity and noxious odor. The control technologies for hydrogen sulfide are discussed in detail in Section 5.

During well drilling and development of vapor-dominated resources, steam is released through well venting to remove debris. Wells may also be vented periodically during flow testing. Steam traps and separators are vented to remove condensate. If electrical load decreases or a power unit fails, steam may be by-passed by venting upstream of the turbine. Free noncondensible

gases are removed from the turbine condenser by a gas ejector to prevent their accumulation. Those dissolved in the condensate can be emitted when the condensate-cooling water is evaporated in the cooling tower. Table 1 summarizes the sources of air pollutant emissions for various geothermal energy conversion systems.

IDENTIFICATION OF WATER POLLUTION DISCHARGE SOURCES

Water pollutants from geothermal energy conversion systems can be discharged from any one of the following sources:

- steam separator
- flash separator
- cooling tower overflow
- cooling tower blowdown
- brine discharged directly from energy cycle
- venting of wells and pipelines
- once-through cooling

A variety of potentially hazardous materials may be contained in the spent fluids from each of the conversion processes. In addition to the major chemical constituents of sodium, chloride and silica, the fluid may also contain dissolved solids, iron, manganese, boron, zinc, barium, fluoride, lead, copper, arsenic, mercury, selenium, chromium, silver and cadmium. These constituents, if not probably contained, may create environmental problems when discharged to receiving streams. In general, the higher temperature geothermal resources contain higher concentrations of dissolved solids. The control technologies for water pollutants are discussed in Section 6.

Water produced in the steam separator contains significant amounts of dissolved solids and suspended particulates. This water is normally combined with the cooling tower overflow and treated for surface discharge or subsurface injection. Conversion systems utilizing flashed steam will emit brine from the flash separator and cooling tower. Cooling tower water needs to be blown down periodically to prevent the accumulation of dissolved solids in the cooling water. Wastewater in the form of concentrated brine is also discharged from the energy cycle involving the use of boilers, impulse turbine and heat exchangers. In addition, the venting of wells and pipelines and once-through cooling systems also produce liquid wastes which are potential sources of pollutants discharge. Table 2 summarizes the probable sources of water pollutants discharge as they relate to the energy conversion systems identified in this section.

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TABLE 1.   AIR POLLUTANT EMISSION SOURCES

EMISSION SOURCES

Conversion system	Well drilling	Well cleanout	Pipeline vent	Power plant by-pass	Condensor ejector	Cooling tower
Direct steam	X	X	X	X	X	X
Flashed steam		X	X	X	X	X
Binary cycle (flashed steam)		X	X		X	X
Binary cycle (hot water)	Does not utilize steam					
Total flow	Does not separate steam					
Direct heating (closed system)	Does not utilize steam					

TABLE 2.   WATER POLLUTANT DISCHARGE SOURCES

DISCHARGE SOURCES

Conversion system	Steam separator	Flash separator	Cooling tower overflow	Cooling tower blowdown	Brine discharged directly from energy cycle
Direct steam	X		X	X	
Flashed steam		X	X	X	
Binary cycle (flashed steam)		X	X	X	
Binary cycle (hot water)					X
Total flow					X
Direct heating (closed system)					X

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SECTION 5

AIR POLLUTION CONTROL TECHNOLOGY EVALUATIONS AND COST ESTIMATES

This section discusses air pollution control technologies that are or may be applicable to air emissions from geothermal energy conversion systems. It also examines the costs of those technologies to the extent that they can be determined.

Because the geothermal industry is still in its early stages of development, most of the control technologies described herein either have not been applied or have been utilized only on a limited scale to geothermal developments. Thus, their applicability and cost must be considered preliminary judgements based primarily on the use of those technologies in related industries.

Technologies to control air pollution from geothermal operations are directed primarily at incoming steam, condenser vent emissions and cooling tower emissions. Although a number of pollutants can be emitted from these sources, a single pollutant that has caused significant environmental concern is hydrogen sulfide (H_2S). For this reason, the control technologies to be discussed in the following sub-section are directed towards H_2S removal.

Published cost data for the H_2S control technologies, directly applicable to the range of conditions occurring at potential geothermal sites, is limited and in many cases non-existent. These cost estimates were derived from a particular set of operating parameters at a given flow rate. Several assumptions were made to facilitate the evaluation of costs for a control technology within the range of hydrogen sulfide concentrations and steam or brine flow rates possibly existing at future geothermal energy conversion sites. Operation and maintenance cost estimates were based on a control system with stable operation and did not include any major upset conditions or extended repair periods. The reference costing data for each control technology and the assumptions utilized to develop cost estimates for additional operating conditions are delineated in the following discussion. Costs for disposal of residuals generated by the hydrogen sulfide control technologies are estimated in Section 7.

Cost estimates were determined for total, installed capital and operation/maintenance costs. These costs were estimated for (1) varying geothermal steam (or brine) flow rates with a specific hydrogen sulfide concentration, and (2) a specific geothermal steam (or brine) flow rate with varying hydrogen sulfide concentrations. Normalized control costs (cost per killowatt-hour or cost per unit flow rate) were also estimated for varying power generation rates with a specific hydrogen sulfide concentration.

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Costs have been standardized to the end of the second quarter 1977 dollars by utilizing the Marshall and Stevens process industries average equipment cost index. Capital costs have been amortized by the capital recovery factor (CRF) over the estimated life of the control technology equipment at an annual interest rate of 8 percent. Amortizing the cost of the control system over its anticipated life, and not calculating the reinvestment necessary for a 30-year operating period, is valid because the cost (in present dollars) for future replacement of the system is equivalent to that incurred for the initial system installation (Grant, 1970). Operating cost for electrical energy consumption is assumed to be 4 cents per killowatt-hour.

STRETFORD PROCESS

The Stretford process is designed to remove H_2S from a gaseous stream and is applicable to geothermal steam conversion processes.

Process Description

A simplified flow diagram of the Stretford process is shown in Figure 7. The process produces elemental sulfur and is applicable to those geothermal energy conversion processes condensing steam (Laszlo, 1976). Noncondensible gases from the condenser ejector are scrubbed with an aqueous solution containing sodium carbonate, sodium metavanadate, and anthraquinone disulfonic acid (ADA). An alkaline solution of sodium carbonate and bicarbonate is produced with the carbon dioxide present in the scrubbed gas stream. The gas stream is scrubbed countercurrently with the alkaline solution in the absorber, and hydrosulfide (HS^-) is formed:

$$H_2S + Na_2CO_3 \rightarrow NaHS + NaHCO_3$$

$$NaHS(aq) \rightarrow Na^+ + HS^-$$
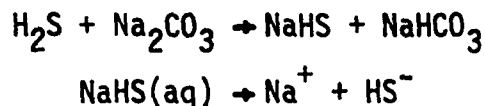
The hydrosulfide is oxidized by 5-valent state vanadate to form elemental sulfur and 4-valent state vanadate:

$$HS^- + V^{+5} \rightarrow S + V^{+4}$$
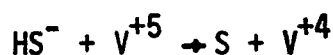
The above reaction is hindered by pH over 9.5, thus the pH is controlled in the optimum range of 8.5 to 9.5 by adding sodium hydroxide. Scrubbing solution is regenerated by blowing air into the oxidizer, and the reduced vanadate is restored to the 5-valent state through a mechanism involving oxygen transfer by the ADA:

$$V^{+4} + ADA \rightarrow V^{+5} + \text{reduced ADA}$$

$$\text{reduced ADA} + O_2 \rightarrow ADA + H_2O$$
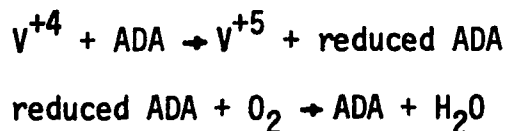
Air blown into the oxidizer brings the suspended elemental sulfur to the surface. The sulfur froth is removed to the skim tank and is filtered, centrifuged, or washed and melted to produce high quality sulfur. The Stretford
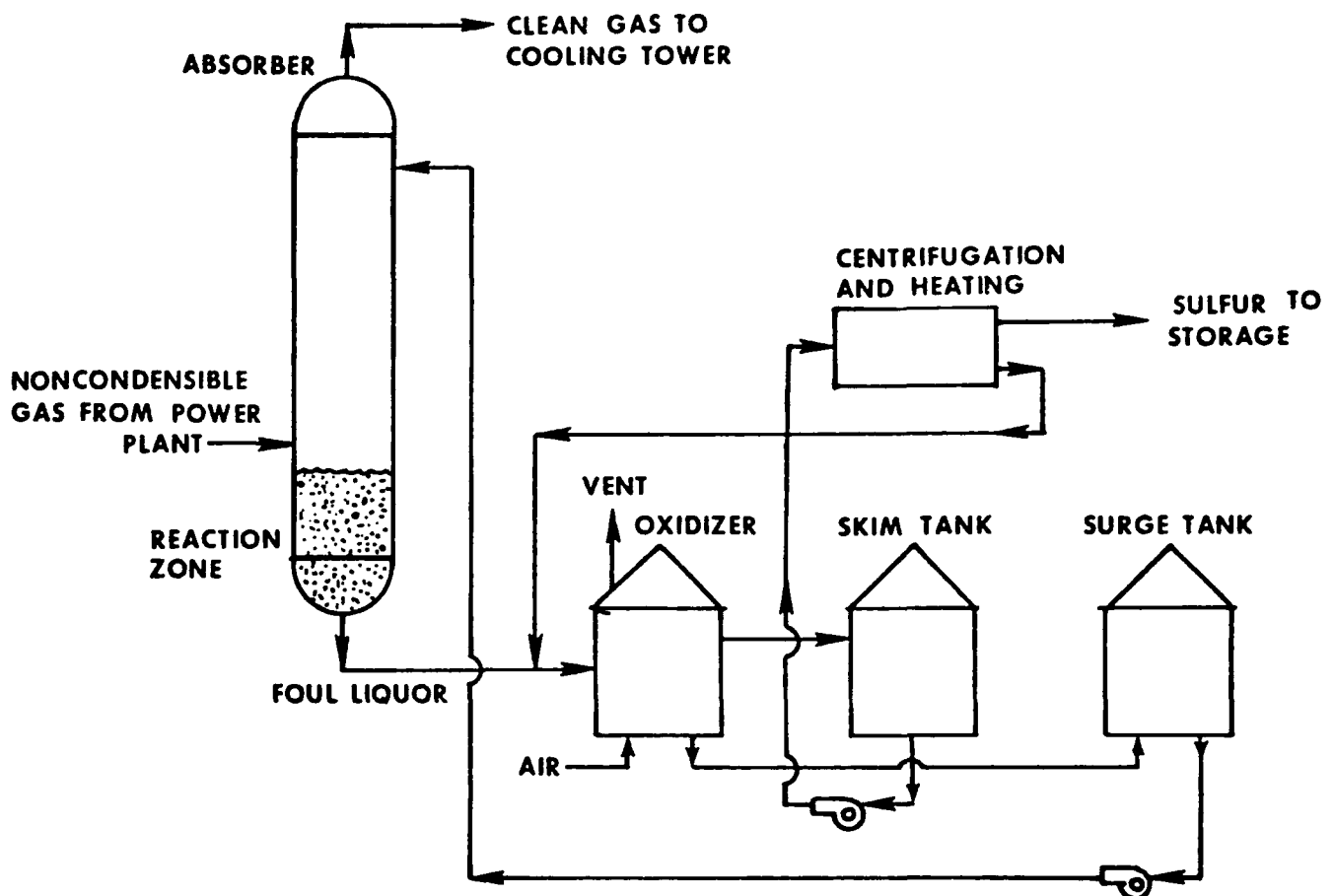
Figure 7. Flow diagram of a Stretford process.

process removes over 99 percent of the hydrogen sulfide from the condenser off gases. The overall reaction is:

$$2H_2S + O_2 \rightarrow 2H_2O + 2S$$
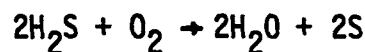
A surface condenser rather than a direct contact condenser must be used with the Stretford process to eliminate direct contact of the cooling water with the condensate. Thus, the amount of water (condensate only - not cooling water) available for hydrogen sulfide to dissolve in is significantly reduced. However, with a surface condenser approximately 10 to 20 percent of the hydrogen sulfide remains in solution with the condensate to be stripped out of solution in the cooling tower and emitted to the atmosphere. Therefore, if a Stretford process is applied to a geothermal energy conversion system designed with a surface condenser, 80 to 90 percent of the hydrogen sulfide existing in the turbine discharge can be removed. The Stretford process will effectively control hydrogen sulfide emissions without any direct detrimental influence on the power cycle. However, retrofitting the conventional geo-

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thermal energy conversion system requires redesigning to include a surface condenser.

Costs

Stretford process cost estimates are based on the process currently being designed for installation on the 117.5 MW, unit 14 power plant at the Pacific Gas and Electric Geysers facility in 1978 (Laszlo, 1976).

The installed capital cost of approximately $2,432,000 was used as a basis for the Stretford cost estimates presented herein (Laszlo, 1976). The Geysers unit 14 will produce electrical power from a vapor-dominated resource with the following operating conditions:

- Steam quality: 180°C (355°F)

    : 7.8 atm (114 psia)

    : 220 ppm average hydrogen sulfide concentration

    : 1200 Btu/lb

- Steam flow rate: 907,000 kg/hr (2,000,000 lb/hr)

- Scrubbing efficiency: 99 percent, or greater

All capital costs for the Stretford process include the differential investment required for a surface condenser in lieu of a direct contact condenser. Capital costs for units with hydrogen sulfide concentration or steam flow rates differing from those given for The Geysers unit 14 base case can be computed utilizing the following formulas obtained from Mark Griebe of the Ralph M. Parsons Company (Griebe, 1977):

$$IA = IB \left(\frac{SA}{SB}\right)^{0.4} \text{ for: } 0.5 < SA < 5 \text{ metric tons of sulfur per day}$$

$$IA = IB \left(\frac{SA}{SB}\right)^{0.5} \text{ for: } 5 < SA < 250 \text{ metric tons of sulfur per day}$$

SA = metric tons of sulfur produced per day in the desired case

SB = metric tons of sulfur produced per day by the base case (The Geysers unit 14) Stretford process.

I = Capital investment for the desired or base (A or B) Stretford process

Based on the above equations, the capital cost for a Stretford unit is assumed to be exponentially dependent upon the quantity of elemental sulfur produced. Ninety percent of the total sulfur entering the power plant as hydrogen sulfide is assumed to be removed by the Stretford process.

The following assumptions were used to estimate the annual capital and operating/maintenance costs for a Stretford unit:

● Amortization period: 15 years (SRI, 1977)

● Maintenance materials: 2 percent of the installed capital cost (Griebe, 1977)

● Maintenance labor: 10 percent downtime, requiring a two-man maintenance crew, earning approximately $30 per hour per person

● Electrical power usage: 66 operating BHP per metric ton of sulfur produced per day (Griebe, 1977)

● Chemical cost: $35 per metric ton of sulfur produced per day (Griebe, 1977)

● Sulfur credit: $20 per metric ton

● Construction site: The Geysers

The accepted market value of commercial grade elemental sulfur is approximately $40 per metric ton. However, the market value is dependent upon the demand in the vicinity of the geothermal site, and in some areas could be as low as $3 to $4 per metric ton (Griebe, 1977). Since geothermal sites are likely to be located in remote areas, a market value of $20 per ton was used to compensate for transportation and other costs. A credit for the elemental sulfur produced by the Stretford process was deducted from the annual operation and maintenance cost.

The Stretford annual costs as a function of steam flow rates ranging from 100,000 kg/hr to 907,000 kg/hr for a constant hydrogen sulfide concentration are presented in Table 3. Costs as a function of hydrogen sulfide concentration varying between 220 ppm and 10,000 ppm at a constant steam flow rate are given in Table 4. Table 3 is based on a hydrogen sulfide concentration of 220 ppm, equivalent to that normally found at The Geysers. Table 4 is based on a steam flow rate of 907,000 kg/hr, equivalent to that of The Geysers 117.5 MW unit 14. Normalized total, capital, and operation/maintenance annual costs, based on the estimates given in Table 3 and 4, are presented in Figures 8, 9, and 10. The costs are based specifically on the design conditions for The Geysers unit 14 power plant and do not apply to geothermal energy conversion systems in general. At other geothermal sites, greater or lesser quantities of steam may be required to produce the same amount of electrical energy. Since the cost of a Stretford process is a function of the sulfur mass flow rate, costs will vary from those presented for other geothermal applications.

Figure 8 gives the costs, in dollars per kg/hr of steam, for steam flow rates varying from 100,000 kg/hr to 907,000 kg/hr and a hydrogen sulfide concentration of 220 ppm. Costs, in mills per KWH, for power generation capacities ranging from 12.95 MW to 117.5 MW and a hydrogen sulfide concentration of 220 ppm are presented in Figure 9. The dependency of the Stretford process costs, in mills per KWH, on hydrogen sulfide concentration in the steam is shown in Figure 10. Costs were estimated for hydrogen sulfide concentrations ranging from 220 ppm (0.022 percent) to 10,000 ppm (1.0 percent). A cost es-

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TABLE 3. STRETFORD ANNUAL COSTS VS. STEAM FLOW RATE.
(220 ppm H_2S)

Costs ($)	Steam Flow Rate (kg/hr)			
	100,000	400,000	700,000	907,000*
Annual Capital	122,800	213,700	267,400	284,200
Maintenance Material	21,800	36,600	45,800	48,700
Maintenance Labor	50,000	50,000	50,000	50,000
Electrical Power	7,200	28,700	50,200	58,500
Chemicals	5,500	21,800	38,200	44,500
Sulfur Credit	-3,100	-12,500	-21,800	-25,400
Total O & M Cost	80,600	124,600	162,400	176,300
Total Annual Cost	203,400	338,300	429,800	460,500

NOTE: *Based on The Geysers unit 14 steam flow rate (Laszlo, 1976)

TABLE 4. STRETFORD ANNUAL COSTS VS. HYDROGEN SULFIDE CONCENTRATION IN STEAM
907,000 kg/hr STEAM FLOW RATE

Costs ($)	ppm H_2S		
	220	2000	10,000
Annual Capital	284,200	857,400	1,916,700
Maintenance Material	48,700	146,800	328,100
Maintenance Labor	50,000	50,000	50,000
Electrical Power	58,500	531,900	2,658,300
Chemicals	44,500	405,300	2,025,500
Sulfur Credit	-25,400	-231,600	-1,157,400
Total O & M Cost	176,300	902,400	3,904,500
Total Annual Cost	460,500	1,759,800	5,821,200

NOTE: A 50,000 ppm (5% by weight) H_2S concentration at the given steam flow rate, results in a sulfur production rate beyond the range of this cost estimate.

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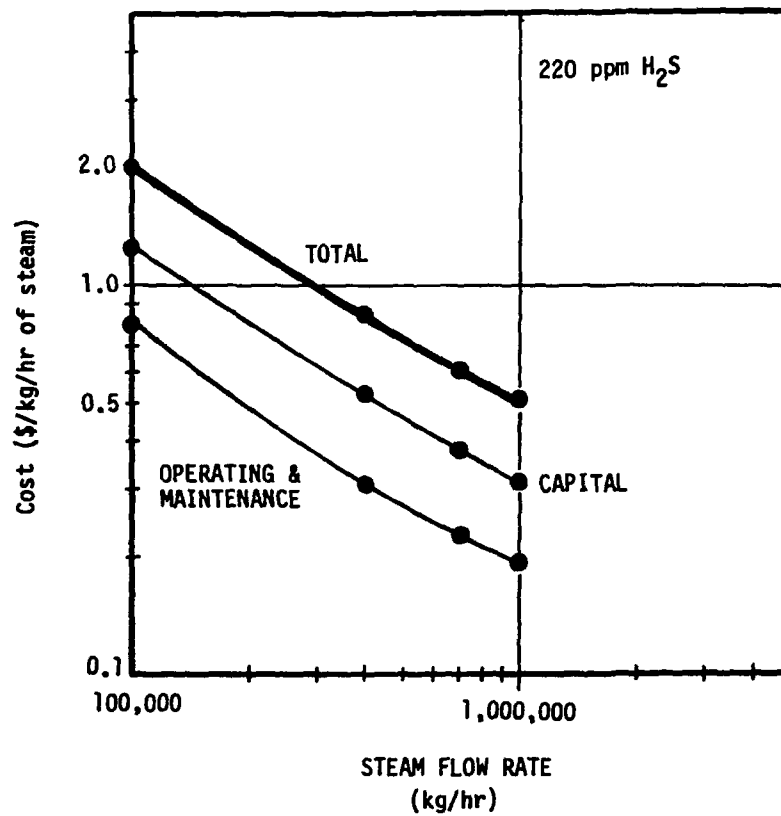
Figure 8. Stretford annual cost vs. steam flow rate
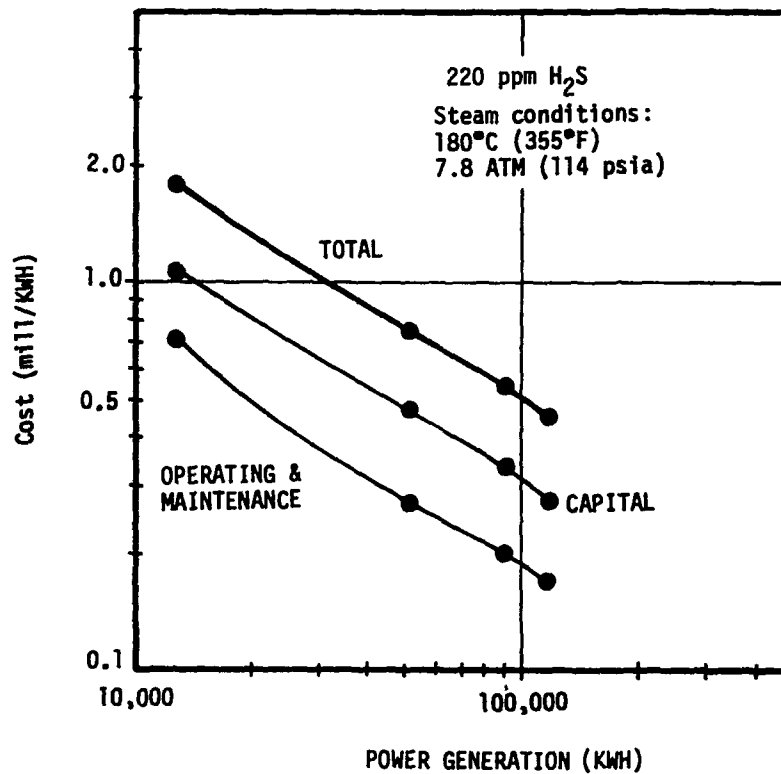
Figure 9. Stretford annual cost vs. power generation

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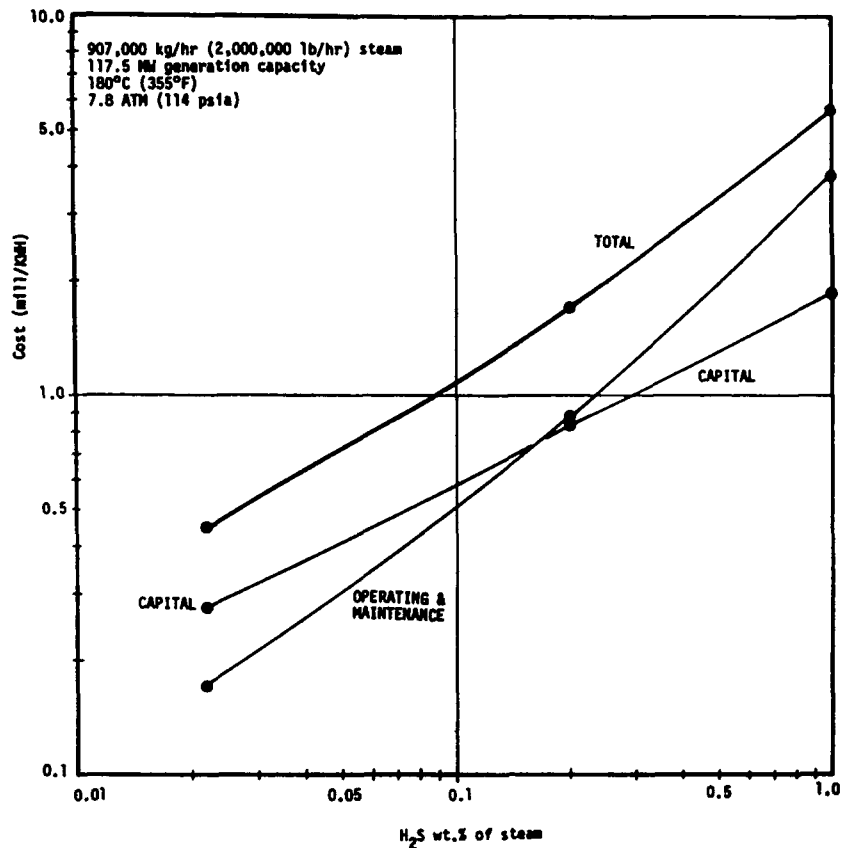
Figure 10. Stretford annual cost vs. H_2S concentration

timate for a 5.0 percent (50,000 ppm) hydrogen sulfide concentration and 907,000 kg/hr steam flow rate is not presented because the resultant sulfur production rate is beyond the valid range of the cost equation.

IRON CATALYST PROCESS

The iron catalyst (or Ferrifloc) system was developed by the Pacific Gas and Electric Company and is presently in use for hydrogen sulfide control at The Geysers geothermal field (Fairfax and McCluser, 1972). This system is applicable to geothermal conversion systems equipped with direct contact condensers.

Process Description

A simplified flow diagram of this process is shown in Figure 11.

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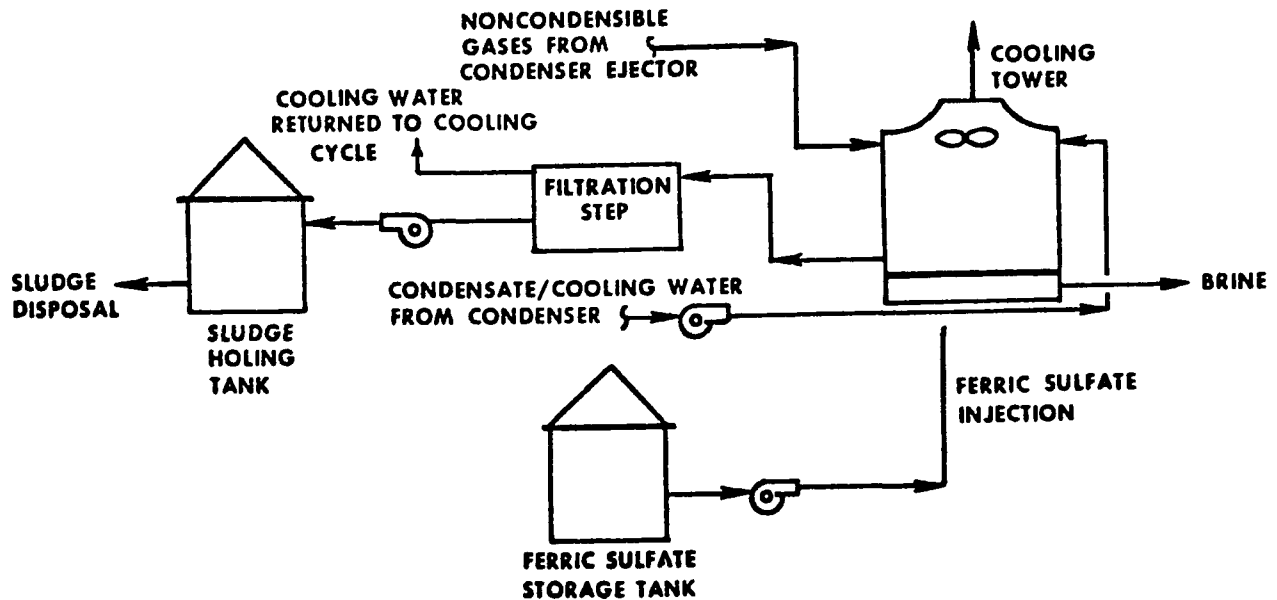
Figure 11. Iron catalyst hydrogen sulfide removal process

Ferric sulfate, in solution, is added to the cooling water, thus oxidizing the hydrogen sulfide contained in the aqueous phase. The noncondensible condenser ejector gases are ducted to the cooling tower and hydrogen sulfide is scrubbed by the falling water containing the ferric sulfate catalyst. Operational experience at The Geysers indicates that, without control, practically all of the hydrogen sulfide dissolved in the cooling water/condensate stream is stripped out into the air stream as it passes through the cooling tower. Therefore, any process controlling hydrogen sulfide emissions must be applied to the cooling water upstream of the cooling tower. The addition of ferric sulfate makes ferric ions available to react with the dissolved hydrogen sulfide, thus forming elemental sulfur, water, and ferrous ions. The reaction mechanism is given below:

$$H_2S(aq) \rightarrow 2H^+ + S^{-2}$$

$$Fe_2(SO_4)_3 \rightarrow 2Fe^{+3} + 3SO_4^{-2}$$

$$2Fe^{+3} + S^{-2} \rightarrow 2Fe^{+2} + S\!\!\downarrow$$

$$2Fe^{+2} + \tfrac{1}{2}O_2 + 2H^+ \rightarrow 2Fe^{+3} + H_2O$$
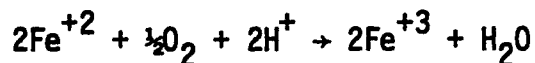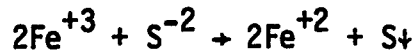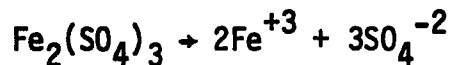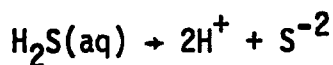
The ferrous ions react with the oxygen encountered in the cooling tower to regenerate the ferric ions. Thus, the regenerated ferric ions are available and the hydrogen sulfide reaction repeats continuously to form elemental sulfur which is removed from the cooling water by filtration. The original design for this system at The Geysers facilities included the use of sand filters; however, significant plugging and maintenance problems have been encountered. To resolve these difficulties, alternative filtration systems are being investigated. The filtration step generates large quantities of toxic sludge that may cause disposal problems. An industrial waste disposal site or appropriate landfill disposal site is required.

The iron catalyst system causes significant corrosion rate increases in the condenser, cooling tower, and associated piping. Plugging problems will be similarly increased in all of the equipment in contact with the cooling water/condensate. The direct contact condensers, presently operating at The Geysers with an iron catalyst system, are clad with stainless steel. It is anticipated that the accelerated corrosion rate will reduce condenser life to seven years. Insoluble salts carried over into the cooling tower blowdown may cause plugging problems in the injection well, if the blowdown is injected.

The iron catalyst system is the only present control technology in use to control hydrogen sulfide emissions from both the cooling tower and condenser ejector. The overall hydrogen sulfide removal efficiency from the power cycle for the iron catalyst system was originally projected to be 90 to 92 percent. Actual field demonstrations by PG&E, however, indicated that the process is at best only 50 percent efficient in H₂S removal. This discrepancy is a result of two major problems which were not accounted for in the original pilot scale tests. The first problem was due to H₂S concentration differences. In the original pilot test unit the H₂S content was substantially lower than that found in the demonstration units. The cooling towers in the demonstration units were not able to adequately oxygenate the cooling waters to achieve more than 50 percent control. Experiments are currently underway to improve the iron oxide process removal efficiency by utilizing caustic soda and hydrogen peroxide as an oxidant source. The second problem was due to mechanical difficulties associated with the use of the process. Plugging of cooling tower nozzles/heat exchangers and corrosion of condenser tubes/pipings are major elements contributing to mechanical failures and loss of efficiency.

Costs

The iron catalyst (or Ferrifloc) system is currently in operation at the Pacific Gas and Electric Geysers facility on the 110 MW unit 11 and the 27 MW units 3 and 4. This system has experienced operational difficulties in the filtration of precipitated sulfur from the cooling water stream (Galeski, 1977). Due to this filtration problem, sludge thickeners will replace the sand filters used in the present installation, thereby resulting in an increased capital investment. The installed capital cost of The Geysers unit 11 iron catalyst system is $1,718,000 and was used as a basis for the cost estimates presented (Laszlo, 1976). The installed capital cost includes a differential estimated investment of $300,000 for sludge thickeners in lieu of sand filters (Galeski,

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1977). The operating conditions of the unit 11 power plant are as follows:

- Steam quality : 180°C (355°F)
  - : 7.8 atm (114 psia)
  - : 220 ppm average hydrogen sulfide concentration
  - : 1200 Btu/lb
- Steam flow rate : 907,000 kg/hr (2,000,000 lb/hr)

Capital costs for iron catalyst systems with steam flow rates differing from that given above can be calculated using the following formula:

$$IA = IB \left(\frac{STA}{STB}\right)^{0.6}$$

STA = Steam flow rate of desired case

STB = Steam flow rate of base case (907,000 kg/hr)

I = Capital investment for the desired or base (A or B) case.

The above equation assumes that the capital investment depends exponentially on the steam flow rate according to the Williams sixth-tenths rule (Hesketh, 1973). The cost of the iron catalyst system is a function of the cooling water/condensate flow rate, which is directly proportional to the steam flow rate. Therefore, the steam flow rate is an acceptable variable in the cost equation. Capital costs were assumed not to be affected by variations in hydrogen sulfide concentration. Operation and maintenance costs for electrical power and chemical usage were assumed to be linearly dependent upon: steam flow rate (with constant hydrogen sulfide concentration) and hydrogen sulfide concentration (with constant steam flow rate). Operation and maintenance costs are difficult to estimate due to the operational problems encountered at The Geysers (Allen, 1977).

The following assumptions were used for the iron catalyst annual capital and operation/maintenance cost estimates:

- Amortization period: 15 years
- Maintenance materials: 1 percent of the installed capital cost
- Maintenance labor: 10 percent down time, requiring a two man crew, earning approximately $30 per hour per person
- Electrical power usage: 68 KW per hour (Galeski, 1977)
- Ferric sulfate usage: 0.5 kg ferric sulfate per kg of hydrogen sulfide, with a loss factor of 20 percent (Laszlo, 1976)
- Ferric sulfate cost: $0.05 per lb or $0.11/kg (Galeski, 1977)
- Removal efficiency: 90 to 92 percent
Construction site: The Geysers

The annual costs for the iron catalyst system as a function of steam flow rate ranging from 100,000 kg/hr to 907,000 kg/hr for a constant hydrogen sulfide concentration, and as a function of hydrogen sulfide concentration varying from 220 ppm to 50,000 ppm for a constant steam flow rate, are given in Tables 5 and 6, respectively. Table 5 is based on a hydrogen sulfide concentration of 220 ppm, equivalent to that normally found at The Geysers. Table 6 is based on a steam flow rate of 907,000 kg/hr equivalent to that of The Geysers 110 MW unit 11. Based on these tables, normalized total, capital and operation/maintenance annual costs are shown in Figures 12, 13 and 14. Generation capacities are based specifically on the operating conditions for The Geysers unit 11 power plant and cannot be applied to geothermal energy conversion systems in general. Figure 14 shows the costs, in mills per KWH, for a 907,000 kg/hr steam flow rate and hydrogen sulfide concentrations varying from 220 ppm (0.022 percent) to 50,000 ppm (5.0 percent).

TABLE 5. IRON CATALYST ANNUAL COST VS. STEAM FLOW RATE. (220 ppm H_2S)

Costs ($)	Steam Flow Rate (kg/hr)			
	100,000	400,000	700,000	907,000*
Annual Capital	53,400	122,800	171,800	200,700
Maintenance Material	4,600	10,500	14,700	17,180
Maintenance Labor	50,000	50,000	50,000	50,000
Electrical Power	2,100	8,400	14,700	19,000
Chemicals	10,200	40,800	71,300	92,400
Total O & M Cost	66,900	109,700	150,000	178,600
Total Annual Cost	120,300	232,500	322,500	379,300

NOTE: * Based on The Geysers unit 11 steam flow rate (Laszlo, 1976)

TABLE 6. IRON CATALYST ANNUAL COST VS. HYDROGEN SULFIDE CONCENTRATION IN STEAM. (907,000 kg/hr STEAM FLOW RATE)

Costs ($)	H_2S ppm			
	220	2000	10,000	50,000
Annual Capital	200,700	200,700	200,700	200,700
Maintenance Material	17,200	17,200	17,200	17,200
Maintenance Labor	50,000	50,000	50,000	50,000
Electrical Power	19,000	19,000	19,000	19,000
Chemicals	92,400	923,800	4,199,800	21,999,800
Total O & M Cost	178,600	1,010,000	4,286,000	21,086,000
Total Annual Cost	378,300	1,210,700	4,486,700	21,286,700
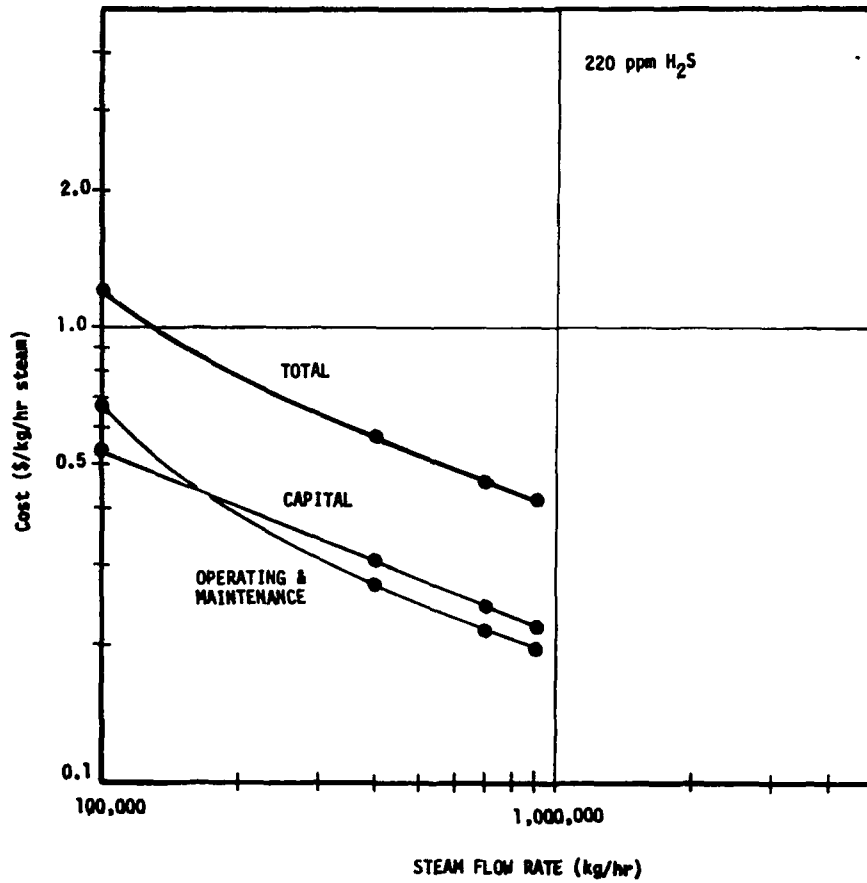
Figure 12. Iron catalyst cost vs. steam flow rate
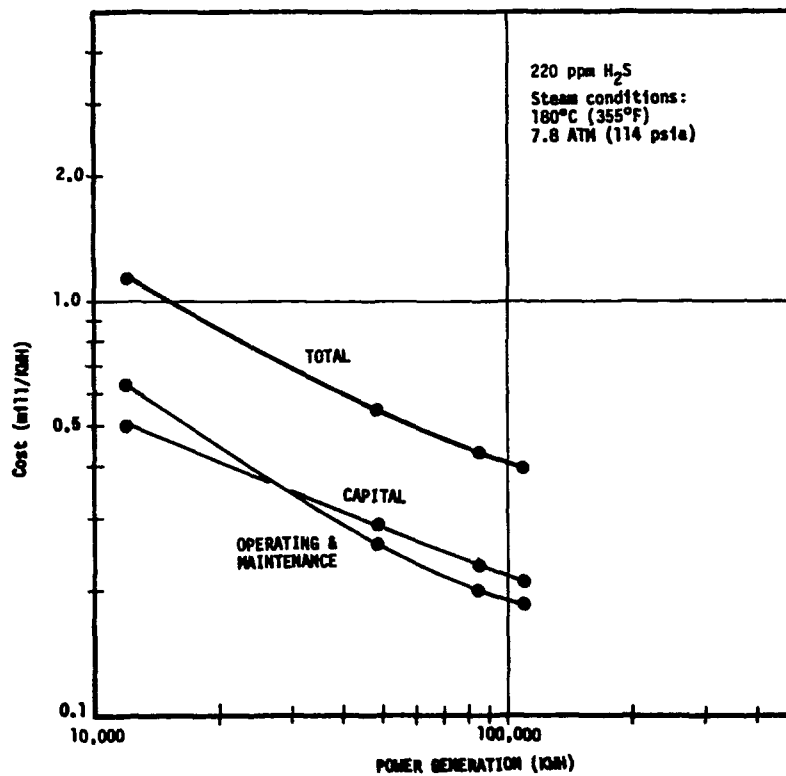
Figure 13. Iron catalyst annual cost vs. power generation

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907,000 kg/hr (2,000,000 lb/hr) steam
110 MW generation capacity
180°C (355°F)
7.8 ATM (114 psia)
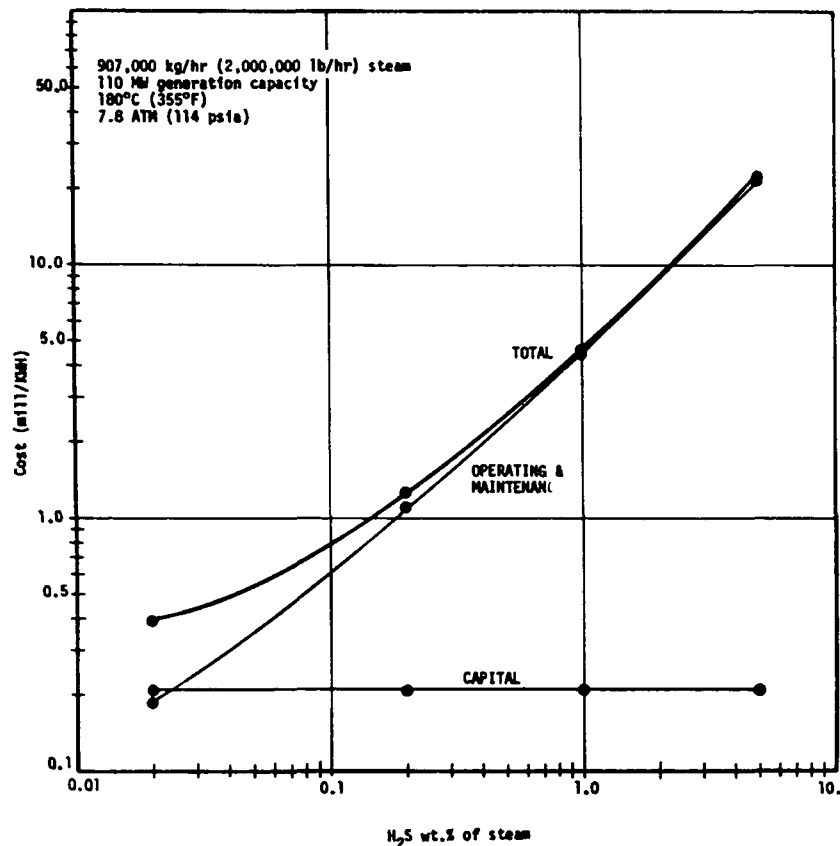
Figure 14. Iron catalyst annual cost vs. H₂S concentration

EIC PROCESS

The EIC process removes hydrogen sulfide (H₂S) from raw geothermal steam by scrubbing it with a aqueous solution of copper sulfate upstream of the power plant (EIC Corp., 1976). The hydrogen sulfide and copper sulfate react in a scrubber, forming a copper sulfide precipitate. The process is potentially valuable because it can remove hydrogen sulfide from the plant input steam thus controlling emissions even while the plant may be shut down and bypassing steam. Another benefit of an upstream scrubbing process is the reduction of corrosive effects of H₂S on the turbine and condensing/cooling cycle equipment. This enables the use of standard materials of construction for the power plant equipment and piping. The EIC process removes hydrogen sulfide without significant degradation of steam quality (temperature and pressure).

Process Description

A simplified flow diagram of the EIC process, with copper sulfate regeneration by roasting, is shown in Figure 15. Figure 16 shows the process with regeneration by leaching.

The process consists of three primary operations: scrubbing, liquid/solid separation, and regeneration. A packed column, sieve tray column, venturi scrubber, or spray scrubber could be used to provide sufficient contact

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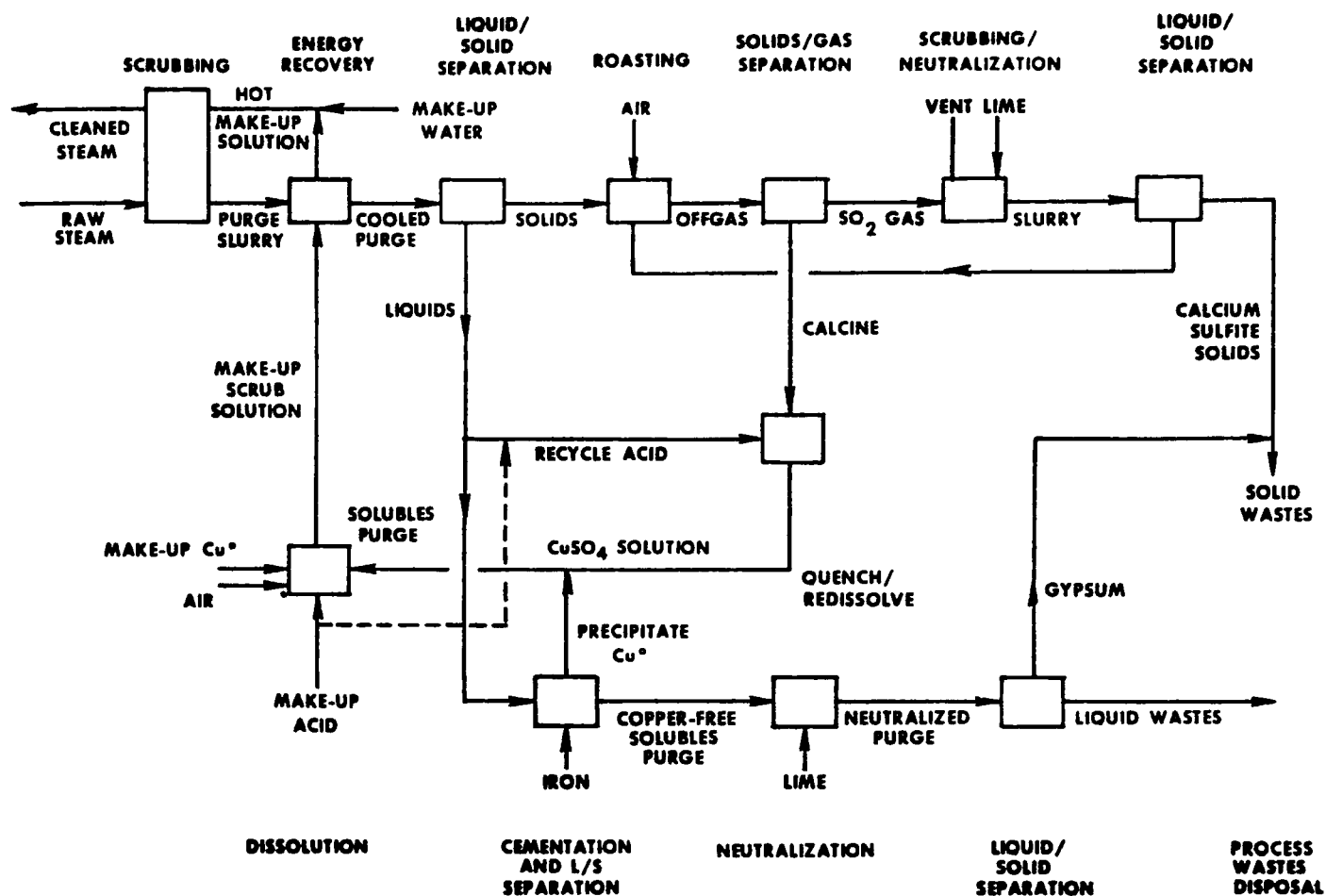
Figure 15. EIC hydrogen sulfide removal process
with regeneration by roasting

time and interfacial area for mass transfer between the hydrogen sulfide and
copper sulfide to occur. An eight-inch diameter single sieve tray column has
been used in field tests at The Geysers. Hydrogen sulfide gas in the geotherm-
al steam is absorbed in an aqueous solution containing dissolved copper sul-
fate and suspended copper oxide particles by the following reaction sequences:

$$H_2S(aq) \rightarrow H^+ + HS^-$$

$$HS^- \rightarrow H^+ + S^{-2}$$

$$CuSO_4(aq) \rightarrow Cu^{+2} + SO_4^{-2}$$

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

$$Cu^{+2} + S^{-2} \rightarrow CuS$$

Overall
reaction $\overline{CuSO_4 + H_2S \longrightarrow CuS + H_2SO_4}$

SCRUBBING   ENERGY RECOVERY   LIQUID/ SOLID SEPARATION   LEACHING   LIQUID/ SOLID SEPARATION   PROCESS WASTES DISPOSAL

CLEANED STEAM

HOT MAKE-UP SOLUTION

MAKE-UP WATER

AIR VENT

RAW STEAM   PURGE SLURRY   COOLED PURGE   SOLIDS   LEACHED SLURRY   SULFUR   SOLID WASTES

LIQUIDS

MAKE-UP SCRUB SOLUTION

RECYCLE ACID

REGENERATED CuSO₄ SOLUTION

MAKE-UP Cu°

SOLUBLES PURGE

AIR

PRECIPITATE Cu°

GYPSUM

MAKE-UP ACID

COPPER-FREE SOLUBLES PURGE

NEUTRALIZED PURGE

LIQUID WASTES

IRON   LIME

DISSOLUTION   CEMENTATION AND L/S SEPARATION   NEUTRALIZATION   LIQUID/ SOLID SEPARATION

Figure 16. EIC hydrogen sulfide removal process with regeneration by leaching.

$$H_2S(aq) \rightarrow H^+ + HS^-$$

$$HS^- \rightarrow H^+ + S^{-2}$$

$$CuO \rightarrow Cu^{+2} + O^{-2}$$

$$Cu^{+2} + S^{-2} \rightarrow CuS$$

$$2H^+ + O^{-2} \rightarrow H_2O$$

Overall reaction $\quad CuO + H_2S \longrightarrow CuS + H_2O$
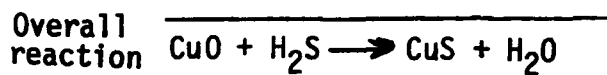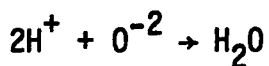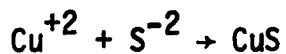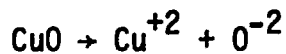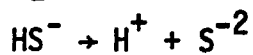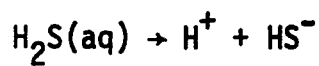
The two reaction chains given above produce a highly insoluble copper sulfide precipitate. The reactions given may be only part of the total reaction chain mechanism. In addition, some reduction of cupric ions occurs, resulting in a cuprous sulfide precipitate. The overall reaction for this mechanism is:

$$Cu_2SO_4 + H_2S \rightarrow Cu_2S + H_2SO_4$$
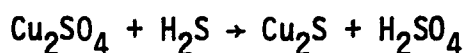
The scrubbed steam passes through a mist eliminator to remove particulate matter before expansion in the turbine.

Copper sulfide slurry purged from the scrubber column is pumped to a centrifuge for liquid-solid separation. The regeneration technique used will determine further requirements of the separation step. If roasting is used, a polishing filter may be necessary to remove fines entrained in the recycle stream. If leaching is used for regeneration, unreacted copper sulfides and elemental sulfur will be contained in the residues, thus requiring chemical flocculants together with filtration to obtain acceptable separation and clarification. To reduce copper sulfate losses, washing of the cake may be required. Clear liquid from the liquid/solid separation process is returned to the scrubber.

Fluid-bed roasting burns the copper sulfide/cuprous sulfide cake from the liquid/solid separation step with air to produce recoverable copper compounds. The roasting regeneration reactions are as follows:

$$CuS + 2 O_2 \rightarrow CuSO_4$$

$$CuS + 3/2 O_2 \rightarrow CuO + SO_2$$

$$SO_2 + 1/2 O_2 \rightarrow SO_3$$
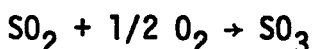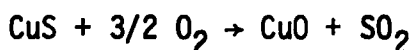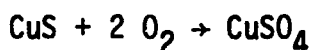
The reactions are highly exothermic, and self-sustaining after start-up. The solid copper sulfate and copper oxide are slurried and reintroduced into the scrubber for continued hydrogen sulfide removal. The sulfur dioxide and sulfur trioxide produced in the regeneration step are scrubbed by an ammoniacal solution. The liquid discharge stream from the sulfur dioxide scrubber is mixed with cooling tower blowdown and injected.

Oxygen pressure leaching is another alternative for recovering copper compounds. The copper sulfide/cuprous sulfide cake requires approximately two to four hours contact time with pressurized oxygen (100 psia) to obtain acceptable conversion rates. The copper sulfide is oxidized to copper sulfate and elemental sulfur, the ratio being a function of residence time, pH, and temperature. If desirable, operating conditions can be controlled to increase elemental sulfur production in the leaching step. The possible reactions for copper sulfate regeneration by leaching are:

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$$CuS + 2\ O_2 \rightarrow Cu^{+2} + SO_4^{-2}$$

$$Cu_2S + 5/2\ O_2 + 2H^+ \rightarrow 2\ Cu^{+2} + SO_4^{-2} + H_2O$$
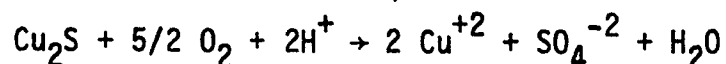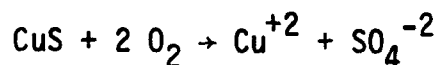
The possible reactions for elemental sulfur formation are:

$$CuS + 1/2\ O_2 + 2H^+ \rightarrow Cu^{+2} + S + H_2O$$

$$Cu_2S + O_2 + 4H^+ \rightarrow 2\ Cu^{+2} + S + 2H_2O$$
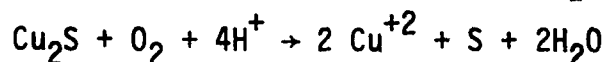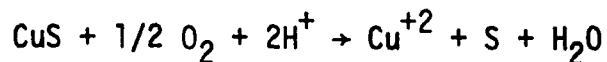
The EIC process was field tested at The Geysers, unit 7 in December 1976. An eight inch diameter single sieve tray scrubbing column was used. Continuous scrubbing of 1000 lb/hr (450 kg/hr) of steam, containing 220 ppm hydrogen sulfide, was accomplished for 30 hours with H_2S removal efficiencies generally over 97 percent. Entrainment of copper from the scrubbed solution into the steam was less than measurable (<0.05 ppm). In addition to hydrogen sulfide, approximately 80 percent of ammonia and boric acid were removed. The field test scrubber was constructed from Carpenter 20 Cb 3 and showed excellent service under the field test operating conditions. Corrosion tests with various stainless steels have shown corrosion rates of less than 5 mils per year.

Costs

Installed capital cost and operation/maintenance costs are summarized in the EIC Corporation Annual Status Report (EIC, 1976) for a 50 MW geothermal power plant with the following design parameters:

Steam to be treated : 17,100 kg/hr (37,700 lb/hr) steam

                         : 53,900 kg/hr (118,900 lb/hr) inerts

                         : 830 ppm H_2S

                         : 150°C (300°F)

                         : 11.9 atm (160 psig)

Scrubbing efficiency : 97.5 percent, or greater

Regeneration process : Leaching

Construction site   : Well-developed site, adjacent to existing facilities

The above steam conditions are based on those encountered for vent gases at the Niland geothermal loop experimental facility, located in Imperial Valley, California. The estimated installed capital cost of an EIC process removing hydrogen sulfide from steam utilized in a 50 MW geothermal power plant, operating with the above conditions, is $4,400,000 (EIC, 1976). Total annual operating cost, including capital, is given in the EIC Annual Status Report as 1.5 mills per KWH for a 500 MW power plant.

Capital costs for EIC units with hydrogen sulfide concentrations differing from that given above for the Niland facility can be computed utilizing the following formula:

$$IA - 0.85 \ IB \ \left(\frac{HA}{HB}\right)^{0.6} + 0.15 \ IB$$

HA = Hydrogen sulfide concentration of the desired case

HB = Hydrogen sulfide concentration for the base case (830 ppm)

I = Capital investment for the desired or base (A or B) case

Eighty-five percent of the capital investment for the EIC process involves reactors, tanks, vessels, heat exchangers, filters, pumps, and other associated process equipment. The remaining 15 percent of the capital investment is allocated for the scrubbing tower. It is assumed that the capital investment for equipment associated with the liquid/solid separation and regeneration operations (85 percent of total) vary exponentially with hydrogen sulfide concentration according to William's sixth-tenth rule (Hesketh, 1973). The capital investment for the scrubbing tower (15 percent of total) is assumed to depend upon steam flow rate and is relatively independent of hydrogen sulfide concentration.

Assumptions used to estimate the annual capital and operation/maintenance costs for an EIC unit are (EIC, 1976):

- Amortization period: 10 years

- Maintenance materials: 2 percent of the installed capital cost

- Maintenance labor: 4 operators at $18,000 per year per person
  1 maintenance man at $20,000 per year
  1 supervisor at $22,000 per year

- Electrical power usage: 2,200,000 KWH per year

- Water usage: 10,000,000 gallons (37.85 x 10^6 liters) per year at $0.50 per 1000 gallons (3785 liters)

- Chemical and process materials:

  sulfuric acid - 300 tons (273 metric tons) per year at $33 per ton ($36.30/metric ton)

  limestone - 250 tons (227 metric tons) per year at $8 per ton ($8.80/metric ton)

  precipitated copper - 37.5 tons (34 metric tons) per year at $1600 per ton ($1760/metric ton)

  detinned scrap - 45 tons (41 metric tons) per year at $200 per ton ($220/metric ton)

  miscellaneous - $19,000

The EIC process annual costs for a 50 MW and 500 MW geothermal power plant are given in Table 7. The annual costs for a 50 MW power plant with hydrogen sulfide steam concentrations varying from 830 ppm to 50,000 ppm are presented in Table 8. The cost estimates given in Table 7 were derived from the basic EIC data given for a 50 MW power plant with a hydrogen sulfide concentration in the steam of 830 ppm. Capital costs as a function of increased hydrogen sulfide concentration were calculated, based on the formula given previously. Operation and maintenance costs for electrical power, water, chemicals, and process materials were assumed to increase linearly with an increase in hydrogen sulfide concentration. Normalized total, capital and operation/maintenance annual costs given in Tables 7 and 8 are shown in Figures 17, 18 and 19. The cost estimates presented for the EIC process were developed from the specific set of operating conditions previously outlined, and may not necessarily apply to geothermal resources with different operating conditions.

Figure 17 gives the cost, in dollars per kg/hr, for steam flow rates of 71,000 kg/hr to 7,100,000 kg/hr (corresponding to 50 MW and 500 MW) and a hydrogen sulfide concentration of 830 ppm. Costs, in mills per KWH, for power generation capacities ranging between 50 MW and 500 MW and a hydrogen sulfide concentration of 830 ppm are given in Figure 18. Costs, in mills per KWH, estimated for a generating capacity of 50 MW and hydrogen sulfide concentrations from 830 ppm (0.083 percent) to 50,000 ppm (5.0 percent), are shown in Figure 19.

TABLE 7. EIC ANNUAL COST FOR 50 MW AND 500 MW PLANTS (EIC, 1976). 830 PPM H_2S

| Costs ($) | Plant Generating Capacity (MW) | |
	50	500
Annual Capital	655,700	3,725,800
Maintenance Material	88,000	--
Maintenance Labor	114,000	--
Electrical Power	88,000	--
Water	5,000	--
Chemicals and Process Material	100,000	--
Total O & M Cost	395,000	3,413,000 *
Total Annual Cost	1,050,700	7,138,800

NOTE: * Derived from EIC cost data for total annual operating costs (EIC, 1976)

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TABLE 8. EIC ANNUAL COST FOR 50 MW PLANT VS. HYDROGEN SULFIDE CONCENTRATION
IN STEAM

Costs ($)	H_2S ppm			
	830	2000	10,000	50,000
Annual Capital	655,700	1,040,300	2,573,000	6,613,700
Maintenance Material	88,000	170,300	631,700	2,451,600
Maintenance Labor	114,000	114,000	114,000	114,000
Electrical Power	88,000	211,200	1,056,000	5,297,600
Water	5,000	12,000	60,000	301,000
Chemicals and Process Material	100,000	240,000	1,200,000	6,020,000
Total O & M Cost	395,000	747,500	3,061,700	14,184,200
Total Annual Cost	1,050,700	1,787,800	5,634,700	20,797,900
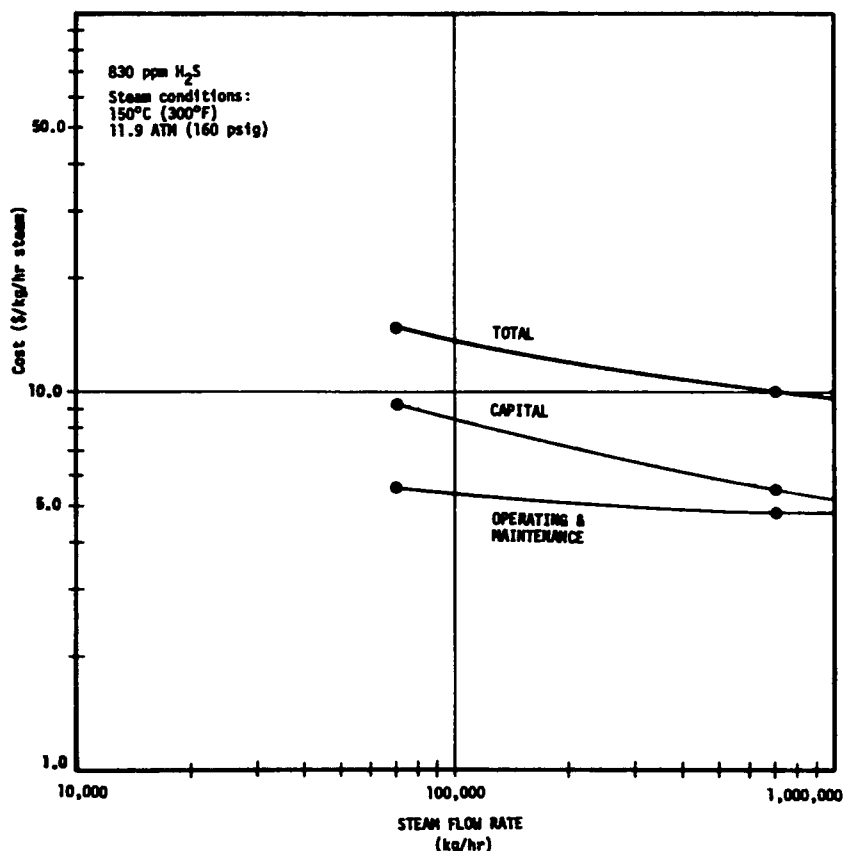
Figure 17. EIC annual cost vs. steam flow rate
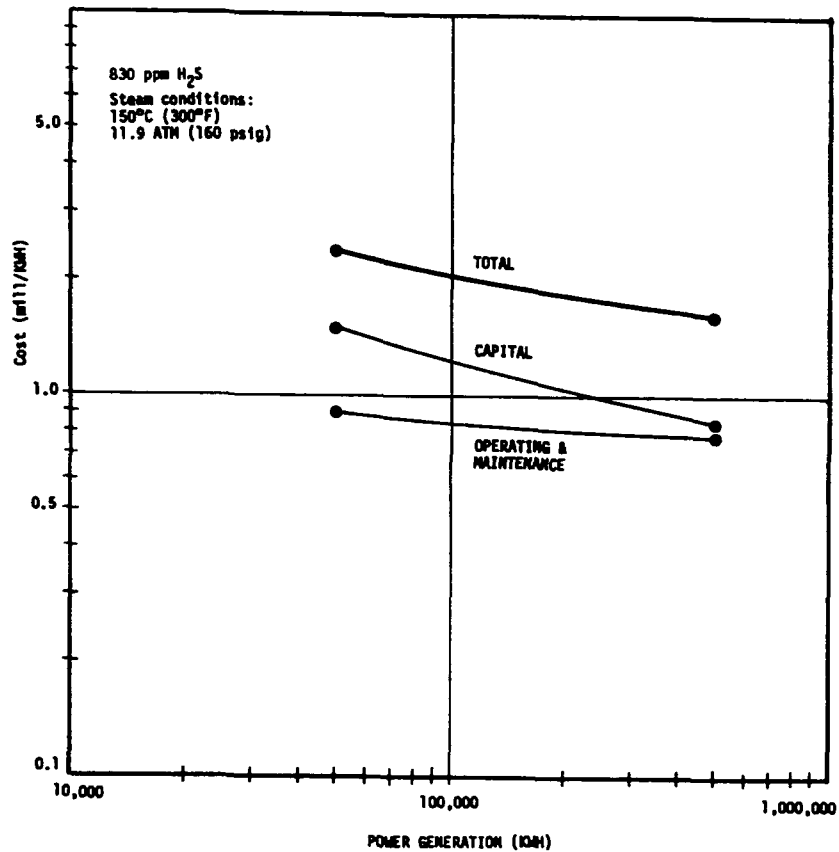
Figure 18. EIC annual cost vs. power generation
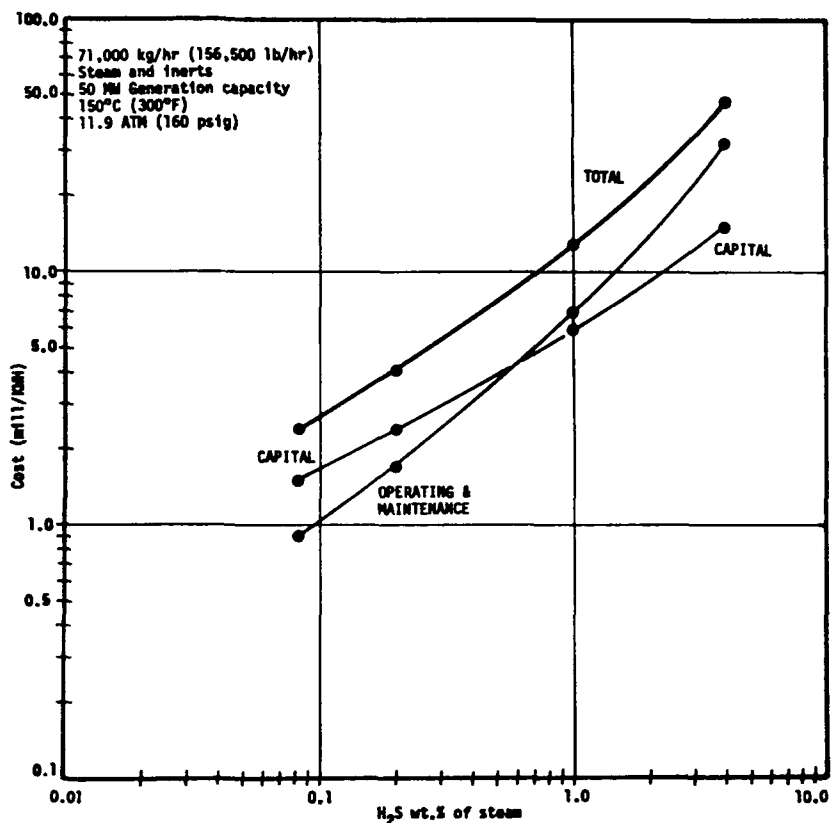
Figure 19. EIC annual cost (mill/KWH) vs. H2S concentration

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DOW OXYGENATION PROCESS

The Dow oxygenation process removes hydrogen sulfide from geothermal liquid at the wellhead; thus, it is applicable only to liquid-dominated resources. Removal of hydrogen sulfide at the wellhead would provide a less corrosive liquid in the pipelines and in the power cycle. The Dow process oxidizes the aqueous hydrogen sulfide by injecting oxygen directly into the geothermal brine (Dow Chemical Co., 1976). Thorough mixing to facilitate contact of the brine and oxygen can be accomplished by using either in-line mixers or a cocurrent packed tower. Although this process appears conceptually feasible, its operation is still in the experimental stage. Full-scale operations are needed to demonstrate its technical as well as economic feasibilities.

Process Description

Simplified flow diagrams of these two systems (in-line mixers and packed tower) are shown in Figures 20 and 21, respectively. Figure 20 shows that ten in-line mixers are required for a geothermal well with a 1000 gallon (3785 liters) per minute flow rate. This design utilizes the largest valuable in-line mixer at an acceptable pressure drop.

In general the oxidation reaction occurs very rapidly, less than one minute for temperatures expected for geothermal fluids. One proposed
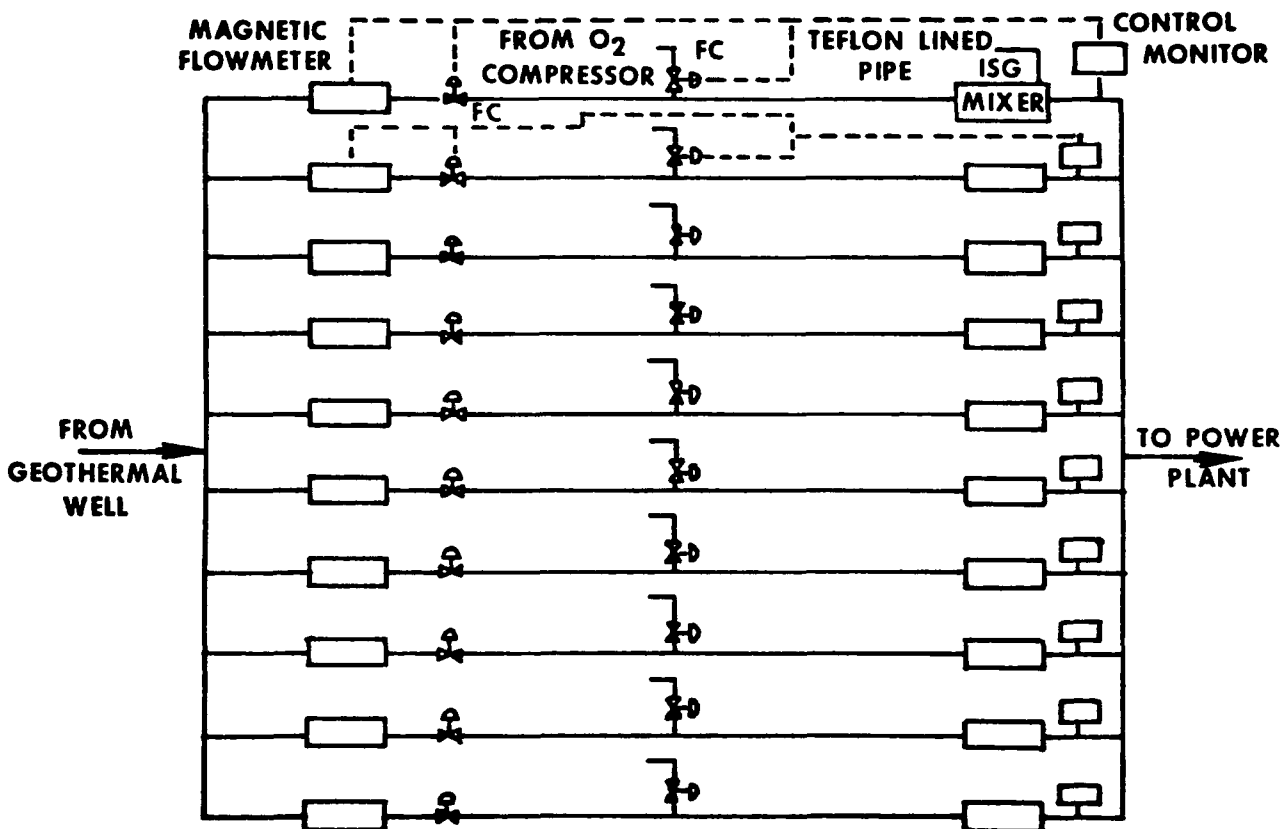
Figure 20. Dow oxygenation hydrogen sulfide
removal process with in-line mixers

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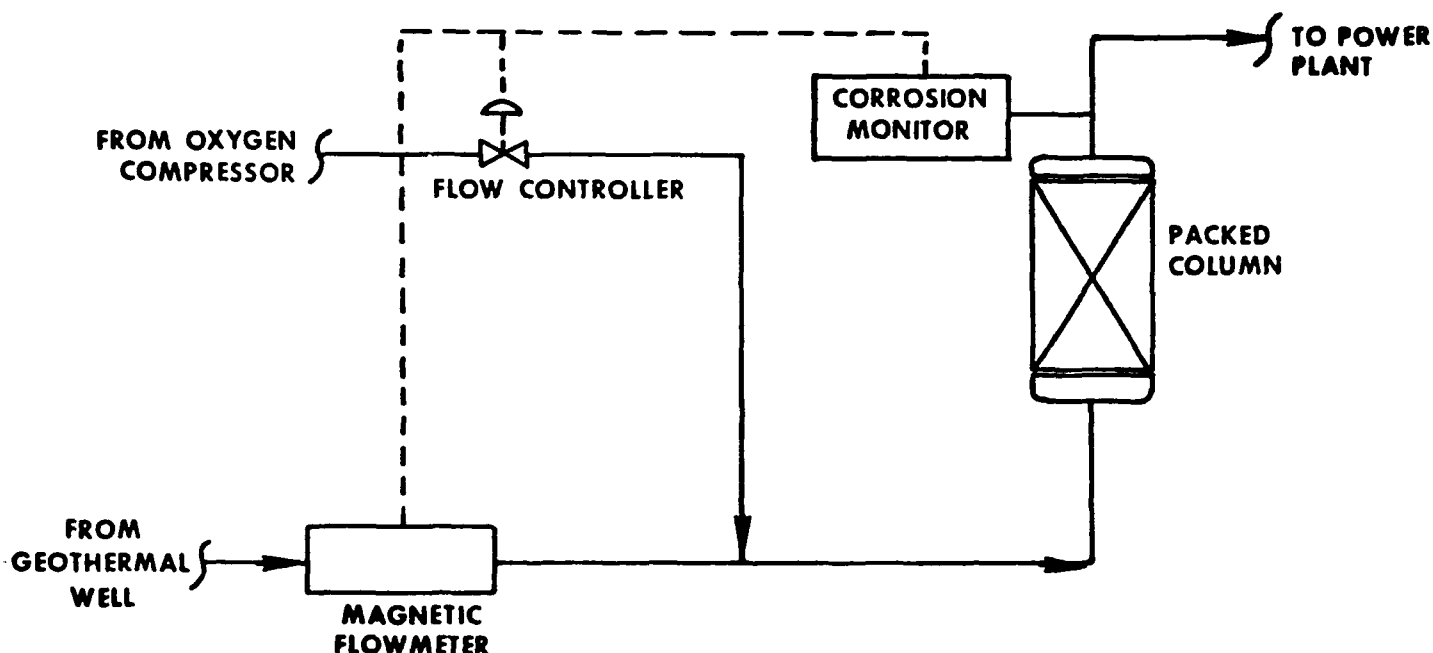
Figure 21. Dow oxygenation sulfide removal
process with cocurrent packed tower.

.reaction chain for the aqueous oxidation of hydrogen sulfide is as
follows:

$$H_2S(aq) \rightarrow H^+ + HS^-$$

$$2HS^- + 3\ O_2 \rightarrow 2\ SO_3^{-2} + 2H^+$$

$$SO_3^{-2} + 1/2\ O_2 \rightarrow SO_4^{-2}$$

$$SO_3^{-2} + HS^- + 1/2\ O_2 \rightarrow S_2O_3^{-2} + OH^-$$

$$S_2O_3^{-2} + 1/2\ O_2 \rightarrow SO_4^{-2} + S$$
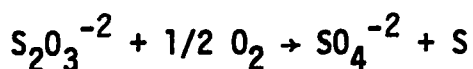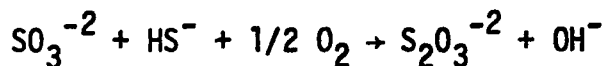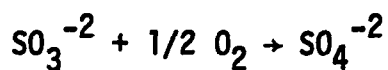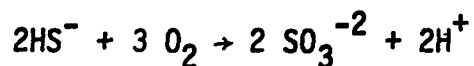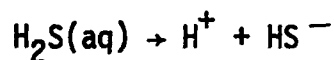
The second reaction given above has an oxygen/sulfide mole ratio of 3:2 (or
1.5:1). However, Dow's laboratory experiments yielded results indicating
complete sulfide oxidation occurred at oxygen/sulfide mole ratios of 1.25:1

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to 1.5:1, depending on temperature and total dissolved salts in the simulated geothermal brine. Thus, it would appear that other reactions, such as the following, must occur:

$$HS^- \rightarrow H^+ + S^{-2}$$

$$S^{-2} + 1/2 \, O_2 + H_2O \rightarrow S + 2 \, OH^-$$

$$S^{-2} + O_2 + H_2O \rightarrow H_2 \, SO_3^{-2}$$
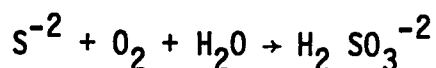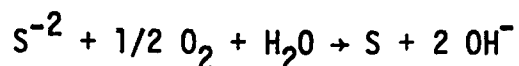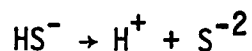
The oxygen/sulfide mole ratios for these two reactions are 0.5:1 and 1:1, respectively. The amounts of elemental sulfur, sulfite, and sulfate formed depend upon the oxygen/sulfide mole ratio, but generally 80 percent or more of the sulfide is converted to sulfate ion, approximately 10 percent to elemental sulfur, and 10 percent or less to sulfite.

After oxygen is injected into the geothermal fluid, and until it reacts with the sulfide, the corrosivity of the fluid increases. This condition requires special materials of construction for both mixing and contact systems. Piping in both systems is teflon-lined between the point of oxygen injection and the mixers or packed tower. The packed tower requires use of a corrosion-resistant alloy. The internal components of the mixers are constructed of teflon.

The in-line mixer system shown in Figure 20 was designed for a well flow rate of 1000 gallons (3785 liters) per minute, thus necessitating the use of ten in-line mixers in parallel, as described previously. In each of the ten lines, a magnetic flowmeter measures the brine flow rate. Each flowmeter is electrically interlocked with a control valve, to ensure that each line has an equal brine flow rate, and interlocked with a control valve injecting compressed cryogenic oxygen into the brine. The brine-oxygen stream passes through the in-line mixers to ensure complete reaction. Injection of excess oxygen is detected with a corrosion rate monitor downstream of the mixers. The brine streams are combined after mixing and the brine is sent to the power plant in mild steel piping.

The packed tower system shown in Figure 21 does not require the duplication of equipment and instrumentation necessary for the in-line system. The geothermal well fluid flow rate is measured with a magnetic flowmeter and oxygen injection is controlled as described for the in-line system. The brine-oxygen stream passes through a packed tower to ensure complete reaction. The piping downstream of the tower can be mild steel.

The Dow oxygenation process has been tested and shown to be technically feasible on a small 3gpm (11.3 lpm) laboratory pilot-plant scale utilizing the in-line mixer system. Initially, catalytic agents were believed necessary to achieve acceptable reaction rates; however, additional catalysts had no measurable effect. Hydrogen sulfide removal efficiencies, at 350°F (175°C) and oxygen/sulfide mole ratio of 1.5:1, generally varied from 90 to 100 percent over a pH range of 5.2 to 11.3.

Costs

Preliminary capital cost estimates for both the in-line and packed column systems have been developed by The Dow Chemical Company based on the results of the laboratory investigation and the following process conditions (Dow, 1977):

- Brine to be treated: 3785 liters per minute (1000 gpm)
    - : 40 ppm H_2S
    - : 177°C (350°F)
    - : 11.2 atm (150 psig)
- Brine phase : single-phase liquid
- Oxygen : hydrogen sulfide mole ratio: 1.25 : 1.0
- Construction site : Imperial Valley, California

The preliminary installed capital cost estimates for an in-line and for a packed column system, with the above operating conditions, are \$373,600 and \$216,500, respectively. Over 31 percent of the capital cost investment for the in-line system is required for instrumentation. Ten in-line mixer trains are required due to the 100 gpm (378 lpm) mixer capacity limitation; thus necessitating a duplication of instrumentation. Capital costs would be significantly reduced if larger capacity mixers could be utilized to minimize duplication.

Capital cost for an in-line system was assumed to depend linearly on brine flow rates. This is due to the required duplication of equipment, negating any possible savings resulting from economies of scale. Capital costs for a packed column system with differing brine flow rates can be computed utilizing the following formula:

$$IA = IB \left(\frac{BA}{BB}\right)^{0.85}$$

BA = Brine flow rate of desired case

BB = Brine flow rate of base case (1000 gpm or 3785 lpm)

I = Capital investment for the desired or given (A or B) case

The capital cost of a packed column system is therefore assumed to be exponentially dependent upon the brine flow rate. The exponential factor was based on that for stainless steel packed towers, 36 to 100 inches in diameter (Hesketh, 1973). The Dow preliminary cost estimate was based on a Carpenter 20 alloy column, packed with teflon pall rings. The exponent utilized in the cost calculation applies to these materials of construction. Capital costs for the in-line and packed column systems were assumed to be independent of the hydrogen sulfide brine concentration. Operation and maintenance costs for electrical power usage and cryogenic oxygen consumption were assumed to be linearly dependent upon the hydrogen sulfide brine concentration.

The following assumptions were utilized to estimate the annual capital and operation/maintenance costs for the in-line and packed column Dow oxygenation systems:

- Amortization period: 15 years

- Maintenance materials: 1 percent of the installed capital cost

- Maintenance labor: 10 percent down time, requiring a two-man crew, earning approximately $30 per hour per person

- Electrical power usage: 5 horsepower oxygen compressor required for 1000 gpm (3785 lpm) system (Galeski, 1977)

- Cryogenic oxygen usage: Calculated for an oxygen/hydrogen sulfide mole ratio of 1.25 : 1.0, an additional 20 percent required to account for system losses.

- Cryogenic oxygen cost: $0.65 per 100 cubic feet ($0.23 per cubic meter).

The annual cost of maintenance materials was taken as 1 percent of the installed capital cost because of the relative simplicity of equipment and design for the Dow process.

The annual costs of the in-line and packed column Dow oxygenation processes as a function of brine flow rates ranging from 15,000 to 350,000 l/min, with a hydrogen sulfide concentration in the brine of 40 ppm, are given in Tables 9 and 10. Annual costs for the Dow processes for a 100,000 l/min brine flow rate and hydrogen sulfide concentrations of 40 ppm, 500 ppm and 1000 ppm are presented in Tables 11 and 12. Figures 22 through 27 are graphs of the normalized total, capital and operation/maintenance costs given in Tables 9 through 12. Cost estimates for the Dow oxygenation in-line and packed column systems have been developed from specific data and conditions, thus cannot be applied to geothermal resources in general.

Figures 22 and 23 give the cost, in dollars per l/min, for brine flow rates from 15,000 l/min to 350,000 l/min with a hydrogen sulfide concentration of 40 ppm for the in-line and packed column systems. Costs, in mills per KWH, for power generation capacities varying from 14.9 MW to 347 MW and with a 500 ppm hydrogen sulfide concentration are shown in Figures 24 and 25. Generation capacities were computed based on a double flash energy conversion system with 8 percent overall efficiency, operating with brine conditions given previously. Figures 26 and 27 represent the costs, in mills per KWH, of the Dow processes as a function of hydrogen sulfide brine concentration at a 100,000 l/min flow rate (98.2 MW).

OTHER H_2S REMOVAL PROCESSES

Several other processes are available for the treatment of hydrogen sulfide emissions. At the present time, they do not appear attractive for geothermal applications because of high costs, low efficiency, proprietary nature of the process, or questionable process reactions under geothermal conditions.

TABLE 9. DOW OXYGENATION ANNUAL COST* FOR IN-LINE SYSTEM VS. GEOTHERMAL BRINE
FLOW RATE 40 ppm H_2S

Costs ($)	Brine Flow Rate (1/min)			
	15,000	100,000	225,000	350,000
Annual Capital	173,100	1,154,600	2,597,800	4,041,000
Maintenance Material	14,800	98,800	222,400	345,900
Maintenance Labor	50,000	50,000	50,000	50,000
Electrical Power	2,000	13,300	30,000	46,700
Oxygen	71,300	475,200	1,070,000	1,664,500
Total O & M Cost	138,100	637,300	1,372,400	2,107,100
Total Annual Cost	311,200	1,791,900	3,970,200	6,148,100

NOTE: * Derived from Dow Chemical Co. cost data for 1000 GPM in-line system (Dow, 1977)

TABLE 10. DOW OXYGENATION ANNUAL COST* FOR PACKED COLUMN SYSTEM VS.
GEOTHERMAL BRINE FLOW RATE 40 ppm H_2S

Costs ($)	Brine Flow Rate (1/min)			
	15,000	100,000	225,000	350,000
Annual Capital	81,500	409,000	814,900	1,186,400
Maintenance Material	7,000	35,000	69,800	101,600
Maintenance Labor	50,000	50,000	50,000	50,000
Electrical Power	2,000	13,300	30,000	46,700
Oxygen	71,300	475,200	1,070,000	1,664,500
Total O & M Cost	130,300	573,500	1,219,800	1,862,800
Total Annual Cost	211,800	982,500	2,034,700	3,049,200

NOTE: * Derived from Dow Chemical Co. cost data for 1000 GPM packed column system.
(Dow, 1977)

TABLE 11. DOW OXYGENATION ANNUAL COST FOR 100,000 1/MIN IN-LINE SYSTEM VS. HYDROGEN SULFIDE CONCENTRATION IN GEOTHERMAL BRINE

Costs ($)	ppm H_2S		
	40	500	1000
Annual Capital	1,154,600	1,154,600	1,154,600
Maintenance Material	98,800	98,800	98,800
Maintenance Labor	50,000	50,000	50,000
Electrical Power	13,300	166,300	332,500
Oxygen	475,200	5,940,000	11,880,000
Total O & M Cost	637,300	6,255,100	12,361,300
Total Annual Cost	1,791,900	7,409,700	13,515,900

TABLE 12. DOW OXYGENATION ANNUAL COST FOR 100,000 1/min PACKED COLUMN SYSTEM VS. HYDROGEN SULFIDE CONCENTRATION IN GEOTHERMAL BRINE

Costs ($)	ppm H_2S		
	40	500	1000
Annual Capital	409,000	409,000	409,000
Maintenance Material	35,000	35,000	35,000
Maintenance Labor	50,000	50,000	50,000
Electrical Power	13,300	166,300	332,500
Oxygen	475,200	5,940,000	11,880,000
Total O & M Cost	573,500	6,191,300	12,297,500
Total Annual Cost	982,500	6,600,300	12,706,500
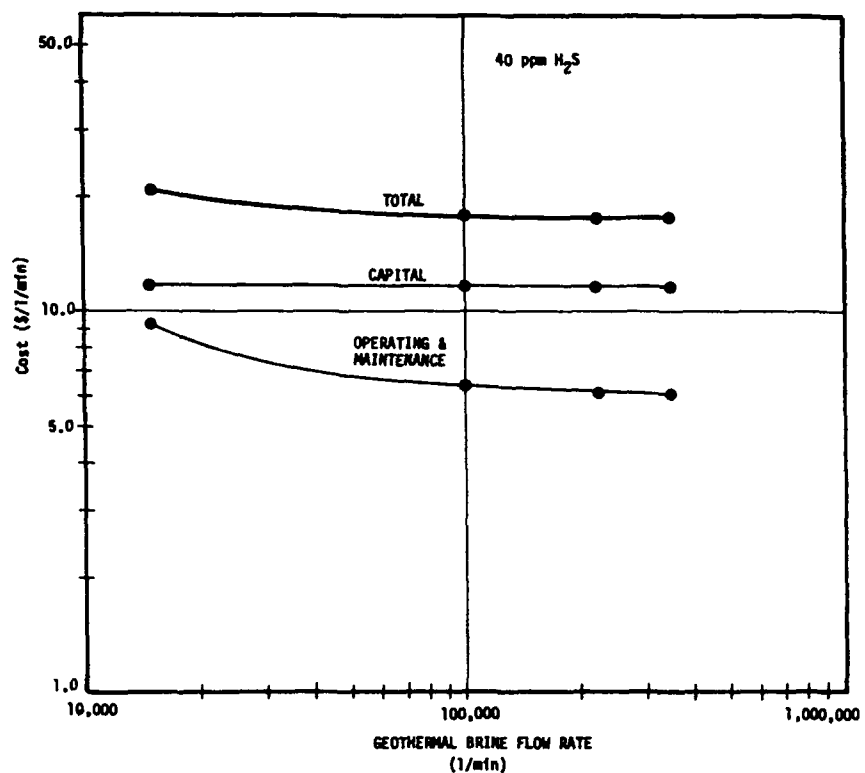
Figure 22. Dow oxygenation - in-line system annual cost vs. brine flow rate
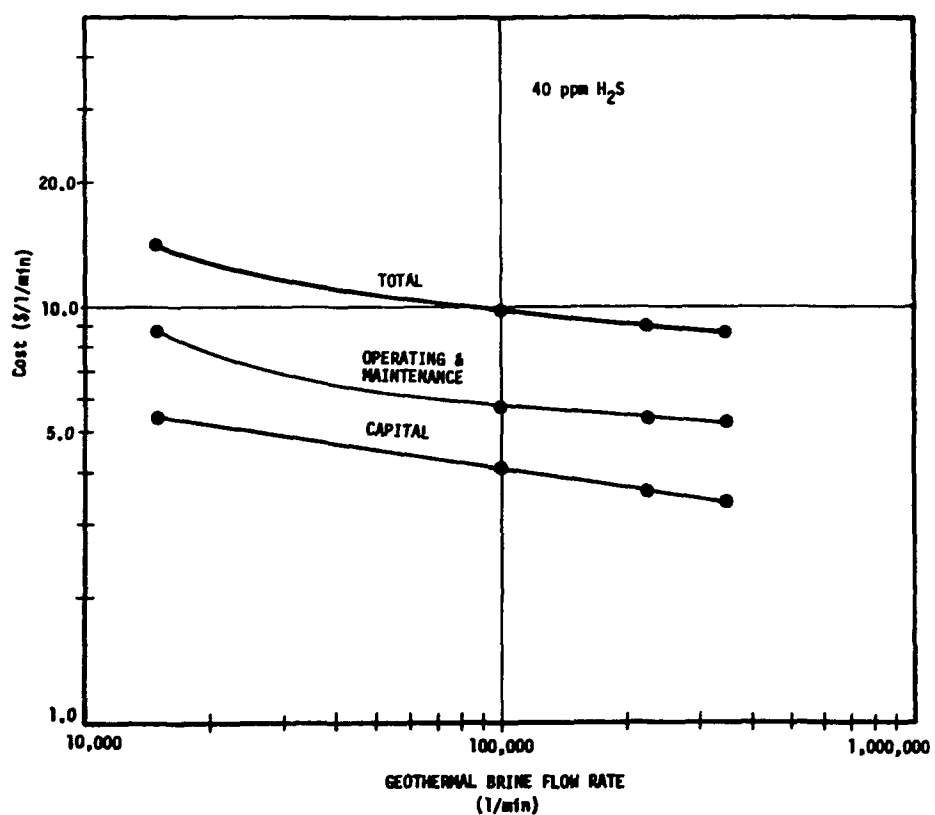
Figure 23. Dow oxygenation - packed column system annual cost vs. brine flow rate
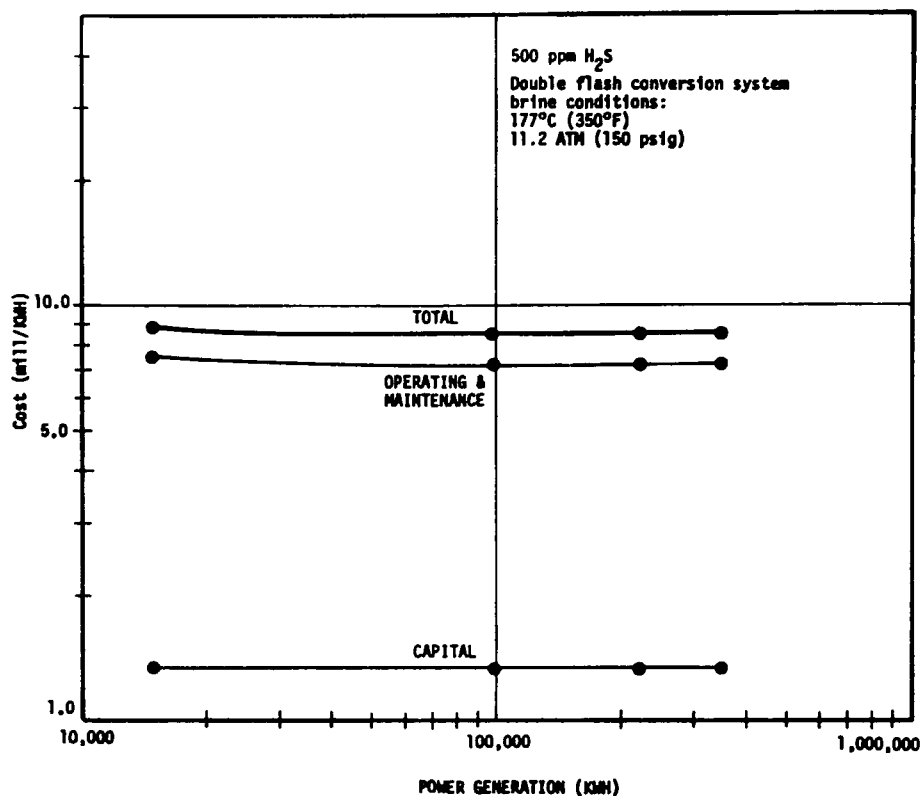
Figure 24. Dow oxygenation - in-line system annual cost vs. power generation
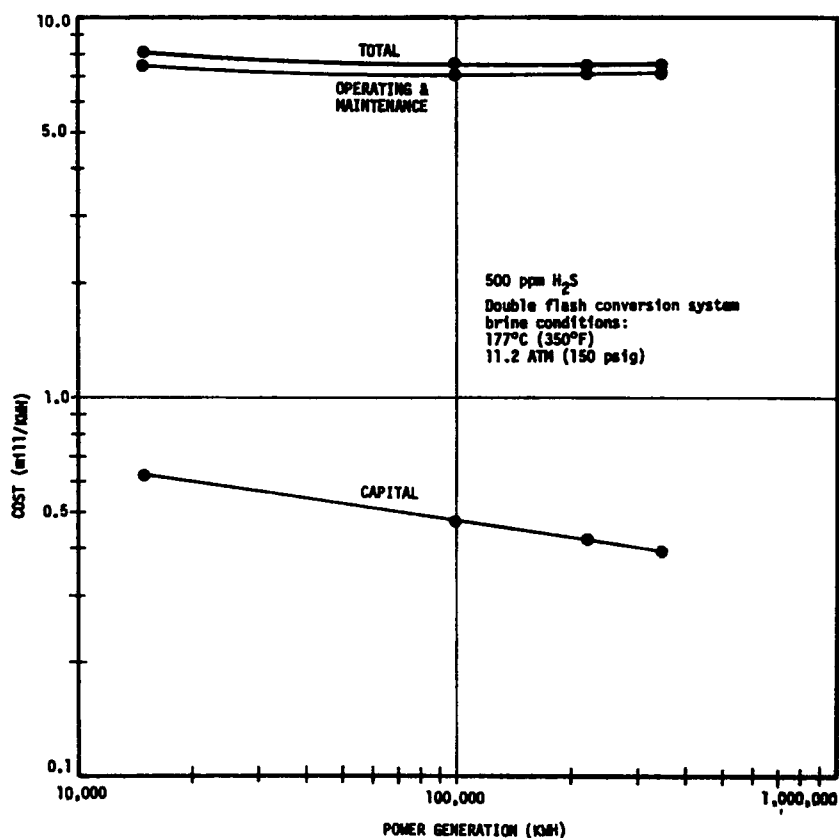
Figure 25. Dow oxygenation - packed system annual cost vs. power generation

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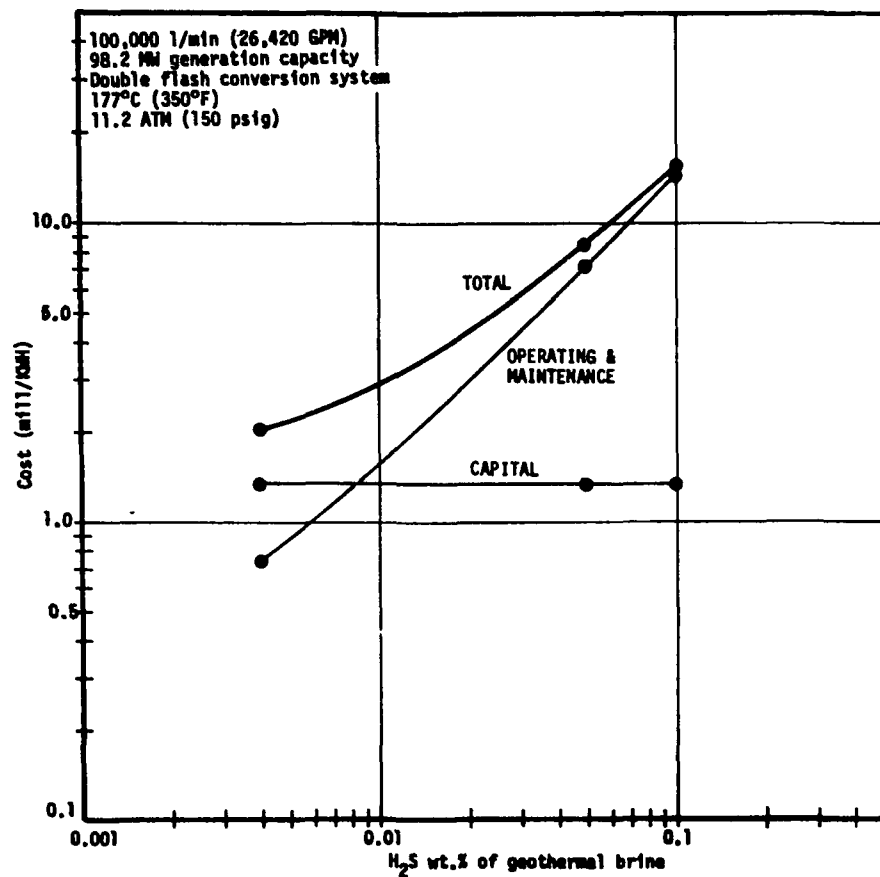
Figure 26. Dow oxygenation - in-line system annual cost vs. H₂S concentration
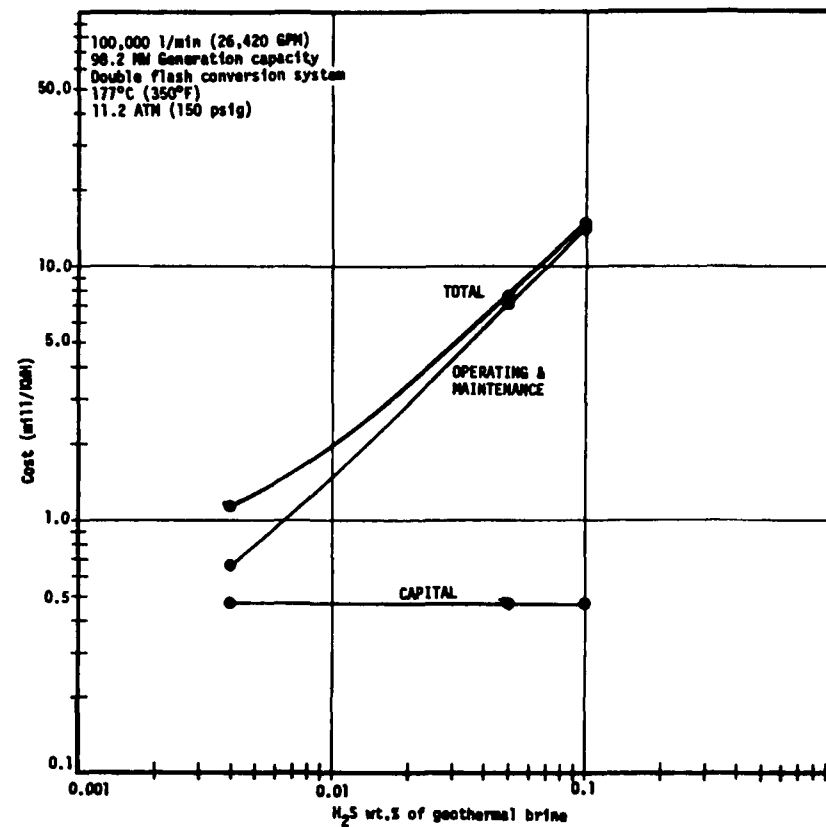
Figure 27. Dow oxygenation - packed column annual cost vs. H₂S concentration

Solid Sorbent Process

Battelle Pacific Northwest Laboratories has investigated numerous solid sorbents for the removal of hydrogen sulfide from geothermal steam (Battelle, 1976). Battelle assumed that the following conditions should be satisfied to establish a technically and economically feasible hydrogen sulfide control process: minimum degradation of steam; regenerable sorbent; reasonably high sorption capacity; simple regeneration process; quick regeneration; and a stable or useful by-product of regeneration. Using simulated geothermal steam, zinc oxide produced the most favorable results among the numerous metal oxide and organic amine sorbents tested. The zinc oxide-hydrogen sulfide adsorption reaction is given below:

$$ZnO + H_2S \rightarrow ZnS + H_2O$$
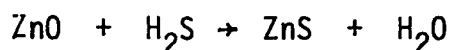
Regeneration is accomplished by reactions with oxygen:

$$ZnS + 3/2\ O_2 \rightarrow ZnO + SO_2$$
$$ZnS + 2\ O_2 \rightarrow ZnSO_4$$
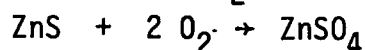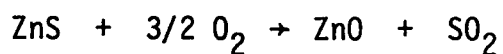
Low temperatures, with oxygen or air regeneration, favor the second reaction producing zinc sulfate rather than zinc oxide. Temperatures in excess of 1200°C are necessary to regenerate zinc oxide directly from zinc sulfide. However, at those temperatures, zinc oxide loses its capacity for adsorbing hydrogen sulfide.

A flow diagram for a sorbent hydrogen sulfide removal process proposed by Battelle is shown in Figure 28. Geothermal steam is introduced to the bottom of a fluidized bed gas-solid contact vessel and hydrogen sulfide is adsorbed by the solid sorbent (zinc oxide). The solid sorbent particles suspended in the steam are removed in a cyclonic separator and, if required, a baghouse. The steam is then utilized in the energy conversion system. Solid sorbent is continuously removed from the fluidized bed contractor to the regenerator. Regenerated sorbent is returned pneumatically to the top of the contractor vessel for reuse. Sulfur dioxide generated in the regeneration process requires treatment in a separate sulfur recovery process. Battelle's laboratory investigation determined that a zinc oxide solid sorbent process is not economically viable for the removal of hydrogen sulfide from geothermal steam and recommends that no further work on solid sorbents be undertaken.

Claus Process

The Claus process is probably the best known process for recovering sulfur from gas streams containing hydrogen sulfide and sulfur dioxide. There are several variations of the process; a specific version of the Claus process flow diagram is shown in Figure 29.

The process requires a specific concentration ratio of hydrogen sulfide to sulfur dioxide. Sulfur dioxide, obtained by combusting part of the hydrogen sulfide, is mixed with the feed stream. The hydrogen sulfide and sulfur dioxide are reacted with each other in a series of converters to produce ele-

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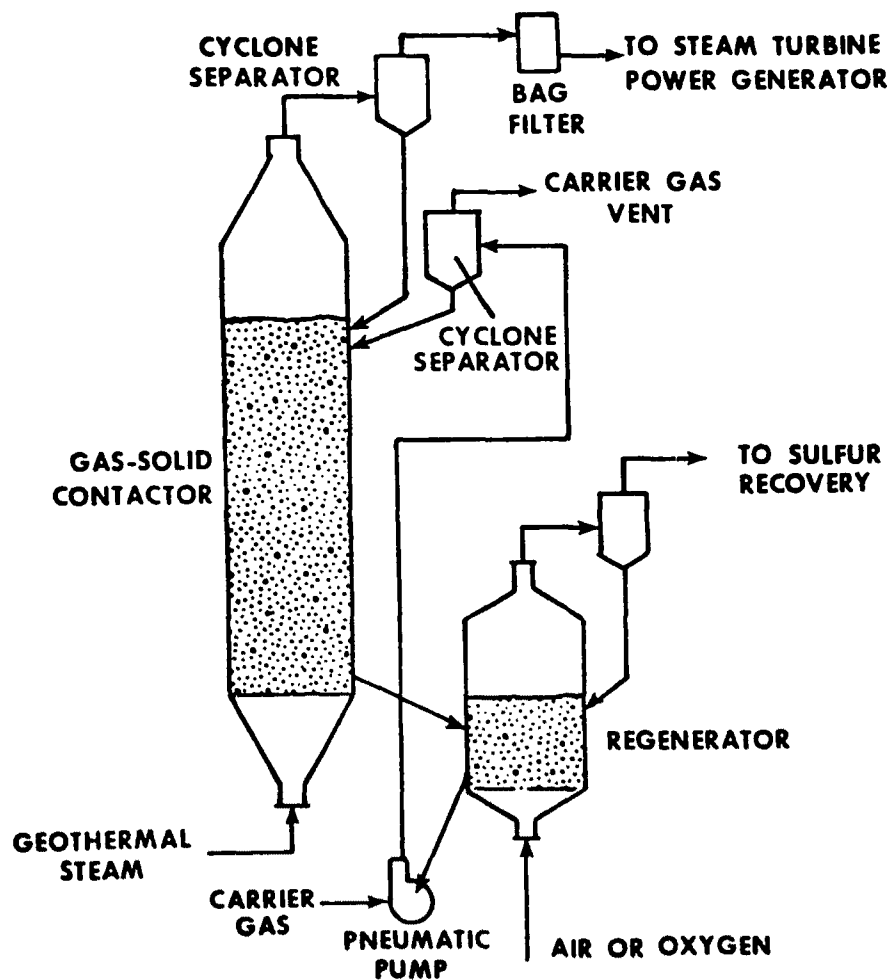
Figure 28. Solid sorption hydrogen sulfide removal process
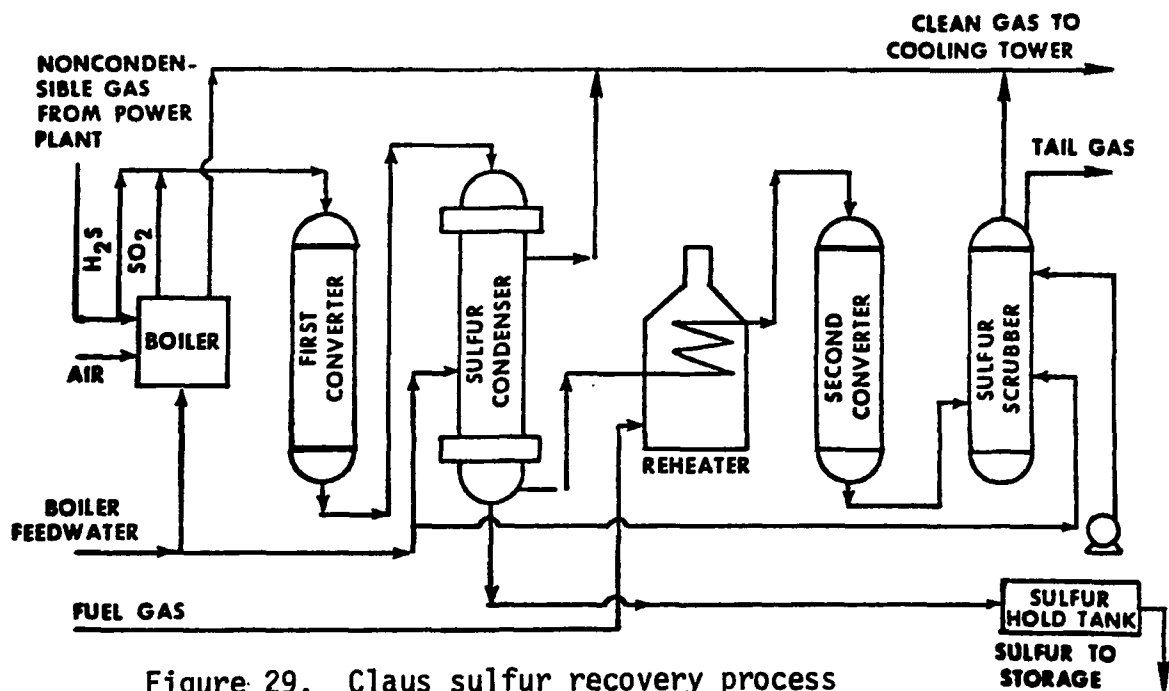
Figure 29. Claus sulfur recovery process

mental sulfur, which is condensed out of the main gas stream. The converters contain an activated bauxite catalyst that accelerates the following reaction:

$$2 \ H_2S \ + \ SO_2 \rightarrow 3 \ S \ + \ 2H_2O$$
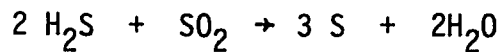
A tail gas containing residual amounts of hydrogen sulfide and sulfur dioxide in moderate concentrations is treated by one of the following processes: recycling into the main process upstream of sulfur separation; sent to another treatment process; or diluted into a large volume of stack gases.

It is doubtful that the Claus process is suitable for removal of hydrogen sulfide from condenser ejector gases. The presence of moisture and carbon dioxide in the feed gas is detrimental to the Claus reaction. Carbon dioxide causes the following side reactions:

$$CO_2 \ + \ H_2S \rightarrow COS \ + \ H_2O$$

$$CO_2 \ + \ H_2S \rightarrow CS_2 \ + \ 2H_2O$$
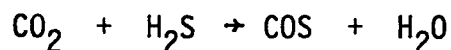
The ejector gases will be saturated and the presence of water tends to reverse the catalyzed Claus reaction.

Hydrogen Peroxide Process

Hydrogen peroxide (H_2O_2) has been used to remove hydrogen sulfide from various wastewater streams. The applicability of H_2O_2 to geothermal cooling water/condensate is somewhat questionable because of the high temperature environment. Hydrogen peroxide reacts with hydrogen sulfide in an acidic or neutral aqueous solution to produce elemental sulfur and water:

$$H_2O_2 \ + \ H_2S \rightarrow S \ + \ 2H_2O$$
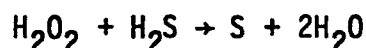
In alkaline solutions (pH >8), the sulfide ion reacts with hydrogen peroxide to produce sulfate and water:

$$H_2S(aq) \rightarrow H^+ \ + \ HS^-$$

$$HS^- \rightarrow H^+ \ + \ S^{-2}$$

$$S^{-2} \ + \ 4H_2O_2 \rightarrow SO_4^{-2} \ + \ 4H_2O$$
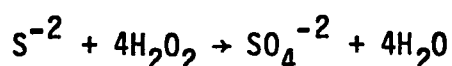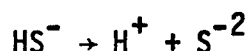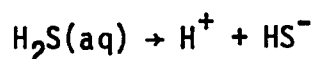
The acidic or neutral reaction is catalyzed by a metal ion, such as the ferrous ion. The rate of the acidic reaction is greatly increased by an increase in temperature. It is interesting to note that four times the hydrogen peroxide is theoretically required to oxidize hydrogen sulfide in an alkaline solution than is required to oxidize that in an acidic solution.

The FMC Corporation has conducted laboratory experiments on oxidation of hydrogen sulfide in samples of cooling water/condensate streams taken from The Geysers power plant (SRI, 1977). The results indicate that the hydrogen sulfide oxidation rate increases as a result of increases in (ranges tested given in parenthesis): initial hydrogen sulfide concentration (2.3-12.5 ppm) temperature (40°-51°C), hydrogen peroxide/hydrogen sulfide weight ratio (0.9-3.9 and 400), and ferric sulfide concentration (0-2.0 ppm). Oxidation of 88 percent of H_2S was obtained in less than three minutes, without the use of a catalyst, and using a hydrogen peroxide/hydrogen sulfide weight ratio of 1.9 and an initial hydrogen sulfide concentration of 12.5 ppm. The results from the FMC experiments indicate that the use of hydrogen peroxide for oxidation of hydrogen sulfide in geothermal cooling water/condensate may be feasible.

Ozone

Oxidation of hydrogen sulfide with ozone in aqueous solutions has not been adequately investigated to evaluate its applicability for controlling geothermal emissions. Ozone has previously been used to oxidize hydrogen sulfide in the gaseous phase. Elemental sulfur and sulfate are the most likely products of the hydrogen sulfide-ozone aqueous reaction:

$$3H_2S + O_3 \rightarrow 3S + 3H_2O$$

$$3H_2S + 4 O_3 \rightarrow 3H_2SO_4$$
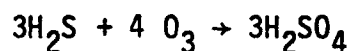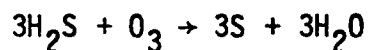
Four times as much ozone is required to produce sulfate as is required to produce elemental sulfur. Because of the high cost of producing ozone, the economic feasibility of this process may depend on which of the two reactions dominate.

Burner-Scrubber Process

The burner-scrubber process incinerates the noncondensible condenser ejector gases and scrubs the combustion products with cooling water. The hydrogen sulfide contained in the ejector gases is burned to sulfur dioxide. The combustion gases are ducted to a scrubber where contact is made with cooling water, thus dissolving the sulfur dioxide. The dissolved sulfur dioxide reduces the pH of the cooling water, which increases the amount of hydrogen sulfide being removed with the noncondensible gases from the condenser. Thus, more hydrogen sulfide is incinerated, rather than remaining dissolved and being stripped from the cooling water into the air stream in the cooling tower. The sulfur dioxide may also oxidize the hydrogen sulfide dissolved in the cooling water to produce elemental sulfur, providing further abatement of hydrogen sulfide emissions. The burner-scrubber system has been field tested on The Geysers 27 MWe Unit 4, with approximately 50 percent of the hydrogen sulfide entering the power plant being removed (Laszlo, 1976).

Catalyst-Scrubber Process

The catalyst-scrubber process is essentially the same as the burner-scrubber system, except the hydrogen sulfide is selectively oxidized to sulfur dioxide with a catalyst developed by the Union Oil Company. Since the hydrogen sulfide is oxidized without combustion, this system is potentially less complex and safer than the burner-scrubber process. The efficiency of the catalyst-scrubber process is also expected to be approximately 50 percent. This process is projected to be installed on The Geysers 53 MWe Units 5 and 6 sometime in 1978.

Deuterium Process

The Deuterium process removes hydrogen sulfide from geothermal steam upstream of the power plant. This process is proprietary and a process description is not currently available. The Deuterium Corporation holds the patent for heavy water, production of which requires steam containing hydrogen sulfide.

SECTION 6

WATER POLLUTION CONTROL TECHNOLOGY
EVALUATIONS AND COST ESTIMATES

This section discusses water pollution control technologies that may have potential applications to wastewater discharges from geothermal conversion processes. It also presents preliminary cost estimates based on information derived from related industries. Although treatment cost is a function of many variables including wastewater quantity and quality, temperature and TDS, it would be very complicated to develop cost curves based on all these variables. For preliminary cost considerations, a simplifying assumption is made to consider treatment cost dependent only on flow. Other variables are assumed to only affect treatment efficiencies and not the cost.

In general, water pollution control technologies include wastewater treatment and wastewater disposal. The following discussion describes both. Depending on the constituents present and the quantities that must be removed, many of the treatment technologies may be used individually or in series. The treatment technologies to be discussed are those applicable primarily to the removal of suspended and dissolved inorganic solids characteristic of geothermal fluids. Treatment system costs have no provision for redundancy.

WASTEWATER TREATMENT TECHNOLOGIES

The major wastewater treatment technologies applicable to geothermal conversion processes are: sedimentation, chemical precipitation, filtration, reverse osmosis, electrodialysis, ion exchange, and evaporation-distillation. The following is a discussion of the technical as well as the economic analysis of these systems.

Sedimentation, Chemical Precipitation and Filtration

Sedimentation Process Description - Sedimentation is a physical treatment operation which removes settleable solids from wastewaters. It is generally applied to raw wastewaters and to wastewaters that have been chemically treated to precipitate constituents. Any one of several configurations of settling ponds, tanks, and gravity separators may be used for sedimentation. They may be used (particularly gravity separators) to concurrently remove floating materials such as oil. Without other treatment, they will not remove significant amounts of dissolved or emulsified materials.

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Sedimentation process efficiency is a function of temperature of the waste-water, the density and size of suspended particles, the amount and character-istics of the suspended material, and settling time. Gravity separation can normally remove 50-65 percent of the suspended solids (Bond, 1974, **U.S.D.O.I.,** **1967).**

Chemical Precipitation Process Description - Chemical precipitation is a chemical treatment process involving chemical addition, particle aggregation and particle precipitation. This treatment process is used to assist the sedi-mentation of colloidal and highly dispersed particles in the waste stream by aggregation and coalescence of small particles into larger more readily settle-able or filterable aggregates. Some dissolved inorganic constituents may also be precipitated by chemical coagulants.

The function of chemical coagulation and mechanical flocculation of wastewater is the removal of suspended solids by destabilization of colloids and removal of soluble inorganic compounds, such as trace metals and phospho-**rus, by chemical precipitation or adsorption on chemical floc. Coagulation in-**volves the reduction of surface charges of colloidal particles and the forma-tion of complex hydrous oxides or precipitates. Coagulation is essentially instantaneous in that the only time required is that necessary for dispersing the chemical coagulants throughout the liquid. Flocculation involves the bond-ing together of the coagulated particles to form settleable or filterable solids by agglomeration. Agglomeration is facilitated by stirring the water to in-crease the collision of coagulated particles. Unlike coagulation, flocculation requires definite time intervals to be accomplished.

The more common chemical coagulants used are lime, soda ash, filter alum, ferric or ferrous sulfate and ferric chloride. Among the coagulant aids used, the more popular ones are sodium aluminate, activated silica, and bentonite or other clays. Generally, chemical coagulants and coagulant aids are added to the waste in a separate chamber in which the waste is mixed rapidly with the chemicals. This system is followed by flocculation chambers and sedimentation tanks.

In general, coagulation reactions vary significantly with changes in pH; therefore for a given coagulant, pH adjustment of the wastewater may be re-quired to achieve optimum conditions. With proper design of the coagulation/ flocculation system and sedimentation tank, removal efficiencies of 80-90 per-cent of suspended solids and 20-40 percent of dissolved solids can be readily attained (Bond, 1974, **U.S.D.O.I., 1967).**

Filtration Process Description - Filtration is a solids-liquids separation technique to remove particulate matter from wastewater. It may be used instead of or in addition to sedimentation. In filtration, the wastewater to be treat-ed is passed through a porous medium. Solids separation is accomplished largely by sieving action. The mechanisms involved in the removal of suspended or col-loidal material from wastewater by filtration are complex and interrelated. The dominant mechanisms depend on the physical and chemical characteristics of the particulate matter and filtering medium, the rate of filtration, and the biological-chemical characteristics of the water. The mechanisms responsible for the removal of particulate matter vary with each treatment system.

Filtration can be accomplished by the use of: (1) a microstrainer, (2) a diatomaceous earth filter, (3) a sand filter, or (4) a mixed-media filter. The microstrainer is a screen in the form of a partially submerged rotating drum or cylinder. Water flows continuously by gravity through the submerged portion from inside the drum to a clear-water storage chamber outside the drum. Cleaning is carried out by backwashing with sprays of product water. Removal efficiencies have been reported for the following parameters (Bond, 1974): SS 50-80 percent; BOD 40-70 percent; and turbidity 60-76 percent.

Diatomaceous earth filtration is a mechanical separation system that employs a layer of filter aid such as diatomaceous earth. As filtration proceeds, deposited solids build up on the precoat, resulting in an increase in pressure drop. The filter run can be increased by the addition of a filter aid to the body feed to maintain the porosity of the cake. When the pressure drop becomes too great to continue filtration, the filter is backwashed and a new precoat applied. Turbidity and suspended solids removals in excess of 90 percent have been reported (Bell, 1962).

Sand filtration may be employed following chemical coagulation and preceding carbon adsorption or ion exchange. The length of the average filter run before backwashing is related to the solids loading on the filter. Generally, filtration rate is low, and backwashing is frequent because of the rapid build-up of headloss. However, removal efficiency for suspended solids is usually very good.

Mixed-media filtration was developed in an attempt to approach ideal filtration. Three to four types of media are layered into the filter, graded as to size and density, with coarse low density coal (sp. gr. about 1.0) on top, smaller regular density coal (sp. gr. about 1.6) and silica sand (sp. gr. about 2.6) in the middle two layers, and garnet sand or ilmenite (sp. gr. of 4.2 and 4.5, respectively) in the bottom layer. These different media provide decreasing, coarse to fine, void gradation down through the filter. Large suspended particles in the wastewater are stopped near the surface with finer suspended solids being entrapped in bottom layers, thus providing full bed depth filtration. Effluent suspended solids concentrations less than 1.0 mg/l are readily achieved (Culp & Culp, 1971).

A typical granular mixed-media filter is shown in Figure 30. The wastewater is passed through one or several layers of granular material and suspended solids are removed by physical screening, sedimentation, and interparticle action. Headloss increases until breakthrough or removal capacity is reached, and then the filter is cleaned by backwashing.

Cost of Sedimentation, Chemical Precipitation, and Filtration

The basis for the development of cost curves is derived from the Van Note, et al., publication "Guide to the Selection of Cost-Effective Wastewater Treatment Systems" (Van Note, 1975).

The cost curves for these treatment systems basically consist of three elements: total cost, capital investment, and operation and maintenance (O&M) costs. Total cost is defined as the sum of capital investment and O&M costs.
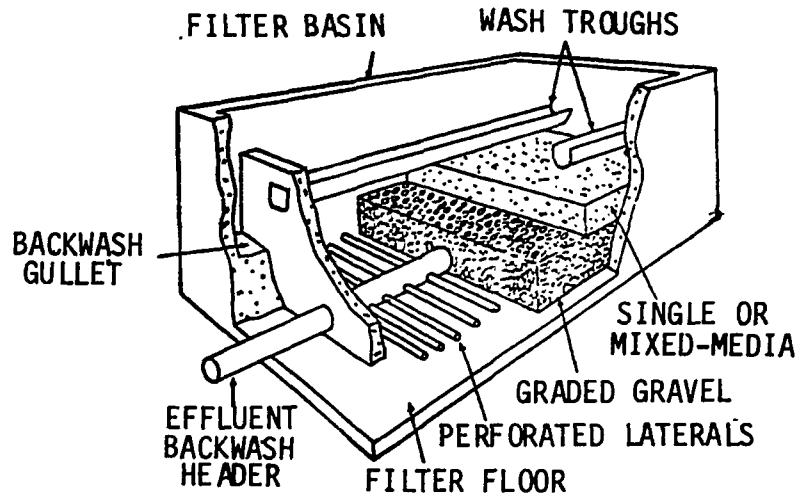
Figure 30. Cut-away view of a granular mixed-media filter.

Capital investment is the cost of purchasing and installing the pollution control systems. O&M costs are associated costs for the operation, repair, and routine maintenance of the pollution control equipment. Since the capital investment as well as the O&M costs are flow dependent, empirical equations have been developed for costing these pollution control systems. In addition to flow, the capital investment is a function of the base capital cost (BCC), land requirement (LR) and, service and interest factor (SIF), and capital recovery factor (CRF). The O&M costs, on the other hand, are functions of base man-hour requirements (BMH) and labor rate (MHR). The total amortized capital cost (TACC) in cents per thousand gallon is given by the following equation:

$$ \text{TACC} = [(\text{BCC})\,(\tfrac{\text{STP}}{177.5}) + (\text{LR})(\text{ULC})]\ [\tfrac{100 + \text{SIF}}{100}]\ [\tfrac{1}{3650\ Q}]\ \text{CRF} $$

and the operation and maintenance costs (O&M) are given by:

fixed operation and maintenance cost in ¢/1000 gal. is

$$ (\text{O\&M})_F = (\text{BMH})(\text{MHR})\ (\tfrac{1}{3650\ Q}), $$

and variable operation and maintenace cost in ¢/1000 gal. is

$$ (\text{O\&M})_V = (\text{BMC})\ (\tfrac{\text{WPI}}{120})\ (\tfrac{1}{3650\ Q}), $$

where

BCC = base capital cost (see Table 13)
STP = October 1977 cost index for average wastewater treatment plant
LR = land requirement (see Table 13)
ULC = land cost (see Table 14)
SIF = service and interest factor (see Table 14)
 Q = wastewater flow (mgd)
CRF = capital recovery factor = $\dfrac{i\,(1 + i)^n}{(1 + i)^n - 1}$

BMH = base man-hours (see Table 13)
MHR = labor rate (see Table 14)
BMC = base materials cost (see Table 13)
WPI = wholesale price index (see Table 14)

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TABLE 13. FLOW VARIABLE COST ELEMENTS FOR WASTEWATER TREATMENT

Process	Base capital cost (BCC)	Land requirement (LR)	Base manhours (BMH)	Base materials cost (BMC)
Pri. sed. conventional	$139753 + 17341.2Q$	$0.23 + 0.088Q$	$1852.8Q^{0.42}$	$1158.4Q^{0.62}$
Pri. sed. 2-stage lime add	$307785 + 33538.6Q$	$0.16 + 0.18Q$	$4259.3Q^{0.41}$	$2956.2Q^{0.66}$
Pri. sed. 1-stage lime add	$198801 + 19934.9Q$	$0.68 + 0.11Q$	$3260.8 + 161.1Q$	$1694.4Q^{0.65}$
Pri. sed. alum. add	$241226 + 33921.4Q$	$0.26 + 0.16Q$	$2783.4Q^{0.47}$	$\dfrac{Q}{(6.62 + 0.036Q)}10^{-5}$
Pri. sed. $FeCl_3$ add	$269563 + 33561.5Q$	$0.26 + 0.16Q$	$2805.5Q^{0.43}$	$2982.5 + 14255.3Q$
Filtration	$231495.0Q^{0.66}$	$0.024 + 0.028Q$	$\dfrac{Q}{(6.8 + 5.8Q)}10^{-4}$	$16491.9Q^{0.68}$
Ion exchange	$163270Q^{0.88}$	$-0.17 + 0.021Q$	$3746.2Q^{0.72}$	$15161.5Q^{0.86}$

where Q = plant capacity or flow (mgd)

TABLE 14. ASSUMPTIONS USED TO DETERMINE WASTEWATER COST CURVES
FOR PRETREATMENT AND ION EXCHANGE

Variables	Notations	Units	Value used to determine cost curve
Q	Wastewater flow	MGD	0.0038, 0.38, 1.52, 1.9 5.7, 11.4, 133
n	Amortized period	Years	20
i	Interest rates	%	8
SIF	Service & interest factor	%	27*
MHR	Labor rate	$/man-hour	12.31**
ULC	Land cost	$/acre	10,000***
WPI	Wholesale price index		199.1****
STP	National average waste-water treat-ment plant cost index		275.0*****

* This includes allowance for engineering, contingencies and interest
  during construction
** As of October 1977, including fringe benefits
*** As of October 1977
**** As of October 1977
***** As of October 1977 (Water Resources Council)

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Using these equations in conjunction with the assumptions presented in Tables 13 and 14, total capital costs and total operation and maintenance costs were computed in cents per thousand gallons for sedimentation, various chemical precipitation systems, and filtration. These values were then converted to cents per thousand liters and plotted in Figures 31 through 36. Costs for the disposal of sludges and brine have not been included. These curves are presented strictly for preliminary cost comparison purposes, based on the assumptions set forth above. Information contained in these curves should not be construed as absolute data points for costing a new or existing treatment system. In particular, variations of geographic locations, climatic conditions, land values and composition of waste streams may invalidate the application of these curves. However, new curves can be developed based on the equations and assumptions provided above.

In costing the sedimentation basins, a surface loading rate (overflow rate) of 800 gallons/day/ft^2 (32,600 1/day/m^2) was assumed. The required surface area of the basins is based on this loading rate. Depending on the nature and characteristics of the geothermal fluid, the overflow rate may not be adequate for complete settling of the suspended material.

The cost curves developed for chemical precipitation by the addition of lime, alum or ferric chloride are applicable for geothermal fluids with chemical characteristics approximating those found in municipal wastewaters. The actual amount of chemical dosage for geothermal fluids will have to be determined by jar test of the geothermal fluid. The chemical dosage in this cost analysis assumes a dosage rate of: 400 mg/l as CaO for a 2-stage lime treatment; 200 mg/l as CaO for a 1-stage lime treatment; 80 mg/l as FeCl$_3$; and 170 mg/l as alum for a 1-stage treatment. Capital costs for both the sedimentation basin and chemical precipitation system include costs for sludge removal devices, piping, pumps and equipment for sludge thickening. Normal allowances for operation and maintenance of chemical equipment are also included.

The costing curves developed for gravity filtration are based on filtration rate of 4 gal/min/ft^2. This rate is highly dependent on the nature of the filtered fluid and the characteristic of the filter media. The capital costs include both the filter and the facilities for storage of backwash water (all pumps and piping were also included). The O&M costs include all power and labor associated with filtration and backwash cycles.
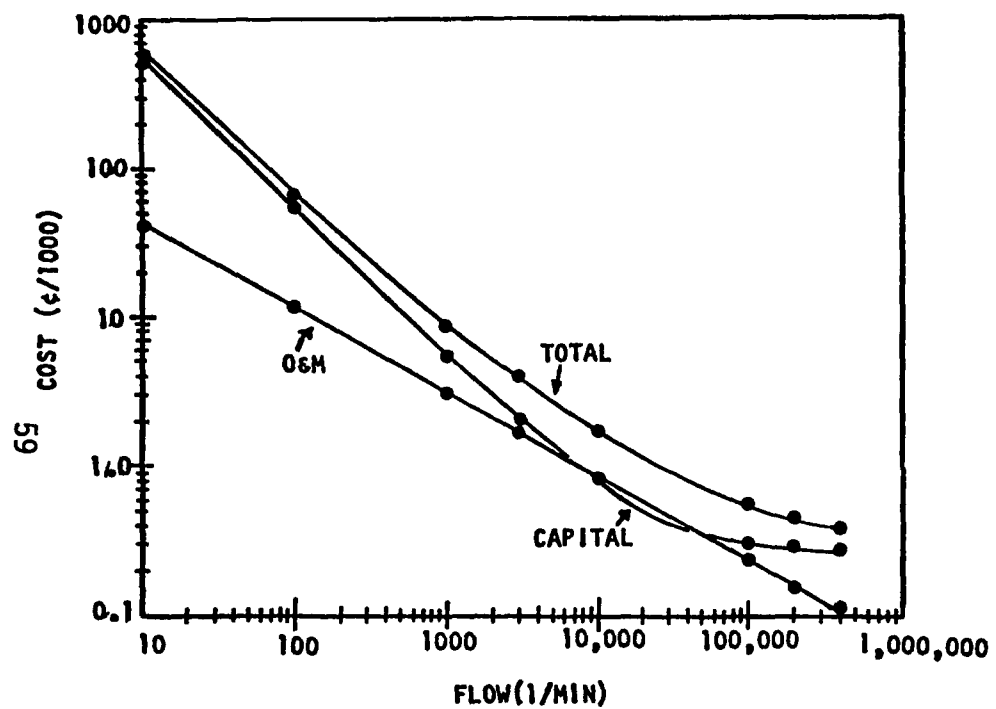
Figure 31.   Cost estimates for sedimentation
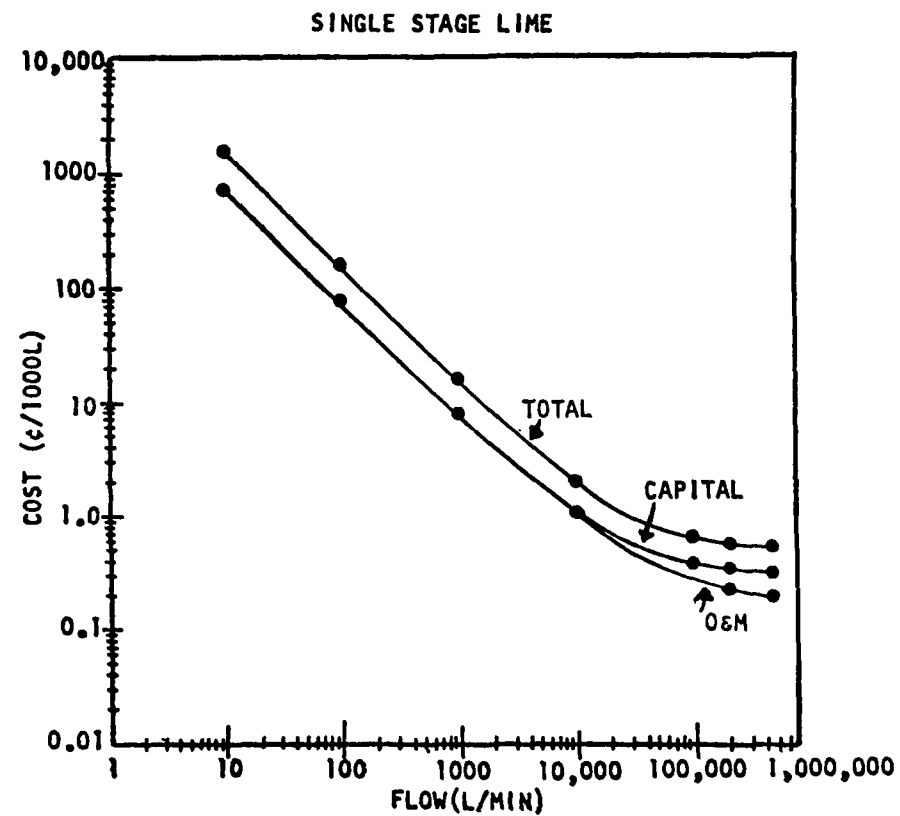
Figure 32.   Cost estimates for chemical precipitation with single stage lime addition.
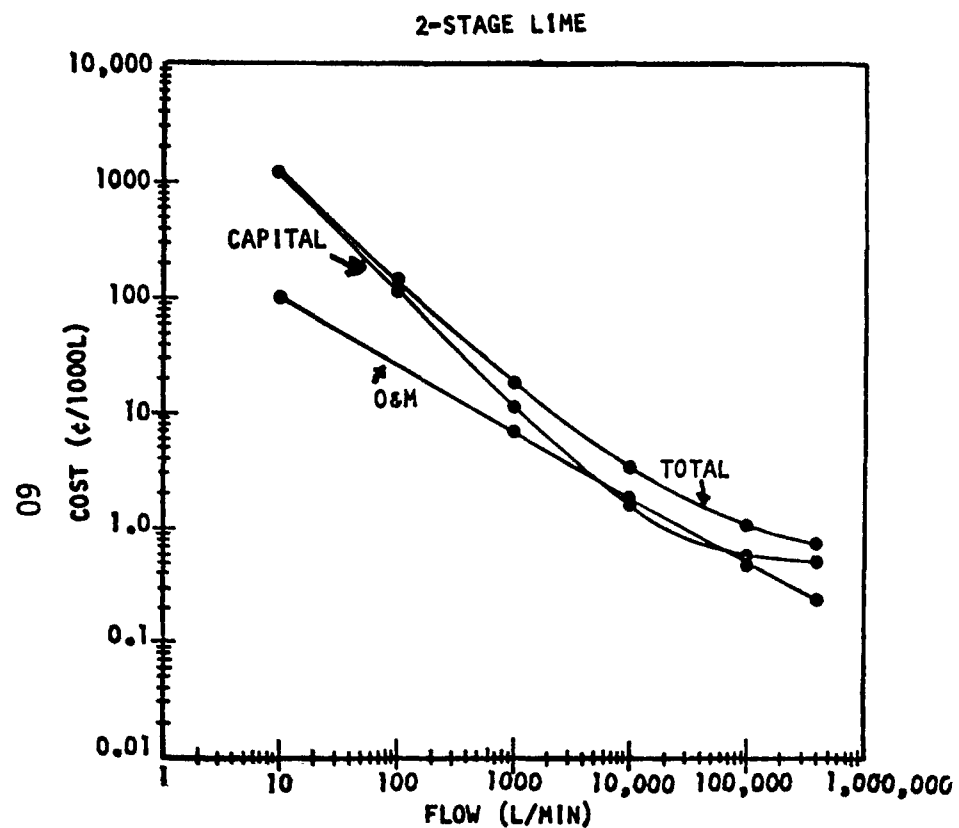
Figure 33. Cost estimates for chemical treat-
ment 2-stage lime addition
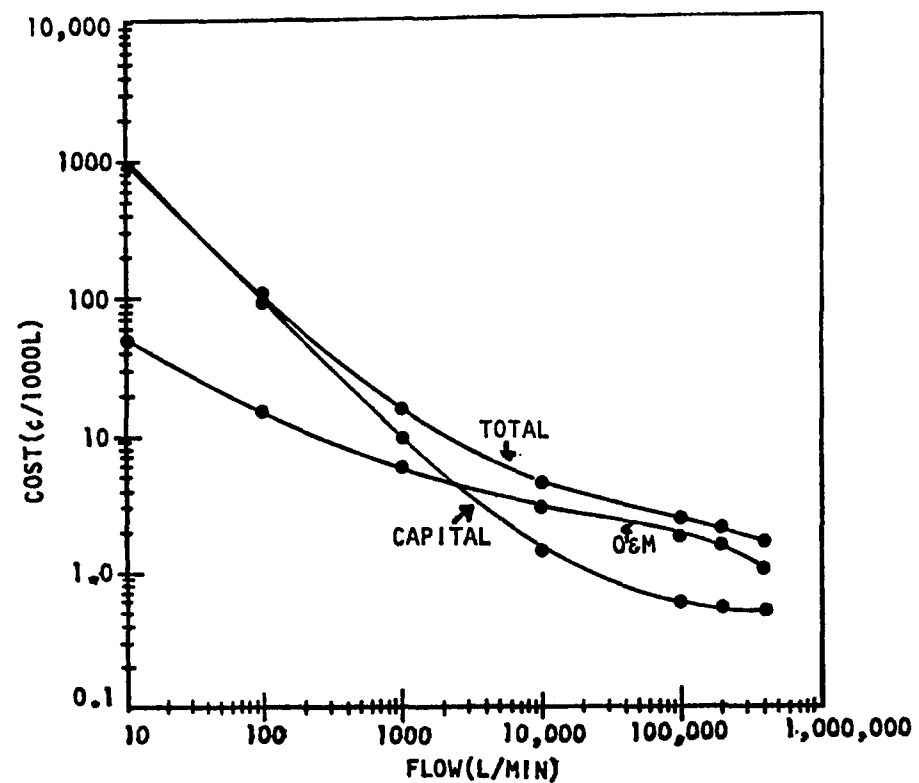
Figure 34. Cost estimates for chemical
precipitation with alum
addition
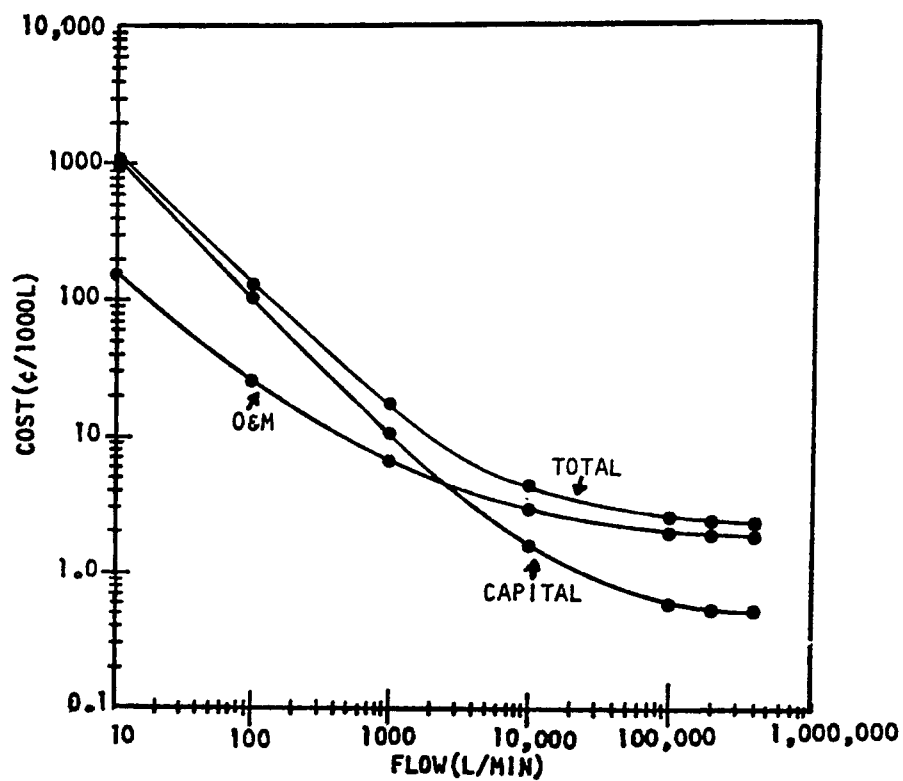
Figure 35.   Cost estimates for chemical
             precipitation with ferric
             chloride addition
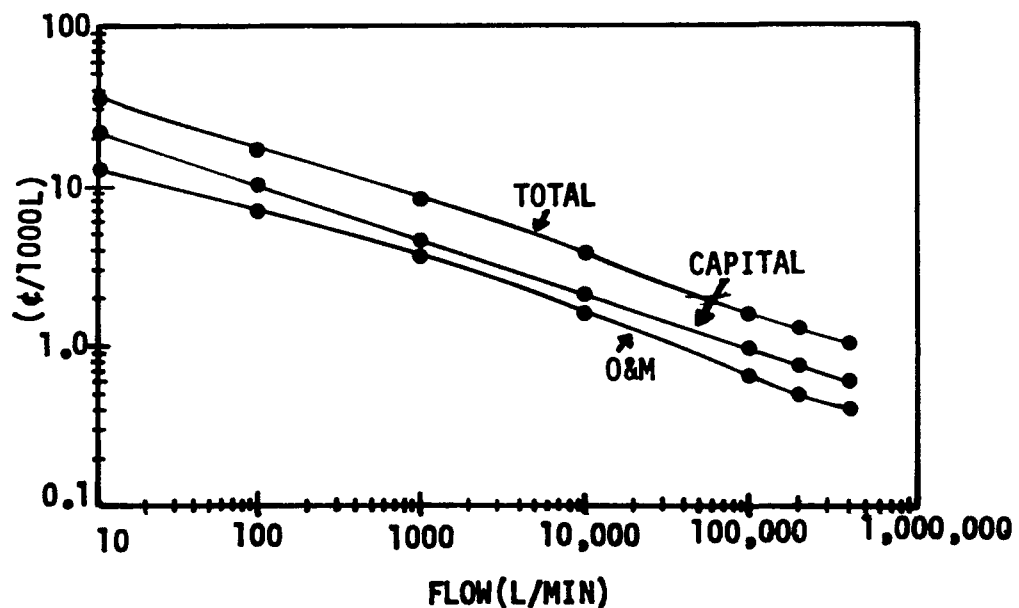
Figure 36.   Cost estimates for filtration

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Reverse Osmosis

Process Description - In this. process, a portion of the wastewater is forced through a semi-permeable membrane. **Figure 37 (Chen, 1977).** The membrane allows passage of water (permeate) while impeding passage of dissolved ions. The portion of the waste stream not forced through the membrane becomes more concentrated in dissolved solids than the original waste. This concentrated solution (retentate) must be disposed of in some manner such as reclaiming it or disposing of it in lined evaporation ponds.

The membrane is the heart of the reverse osmosis process. Most membranes in current use are cellulose acetate. However, properties of cellulose acetate membranes vary according to the method of manufacture. Therefore, different membranes have different permeabilities for various ions. The technical feasibility of reverse osmosis is determined by the availability of a membrane which sufficiently limits passage of the ion to be removed while allowing passage of a reasonable amount of water. Any solids (suspended or colloidal) present in the waste stream will impede the passage of water through the membrane. For proper operation, a filtration system is usually installed ahead of the reverse osmosis unit to minimize plugging of the membrane.
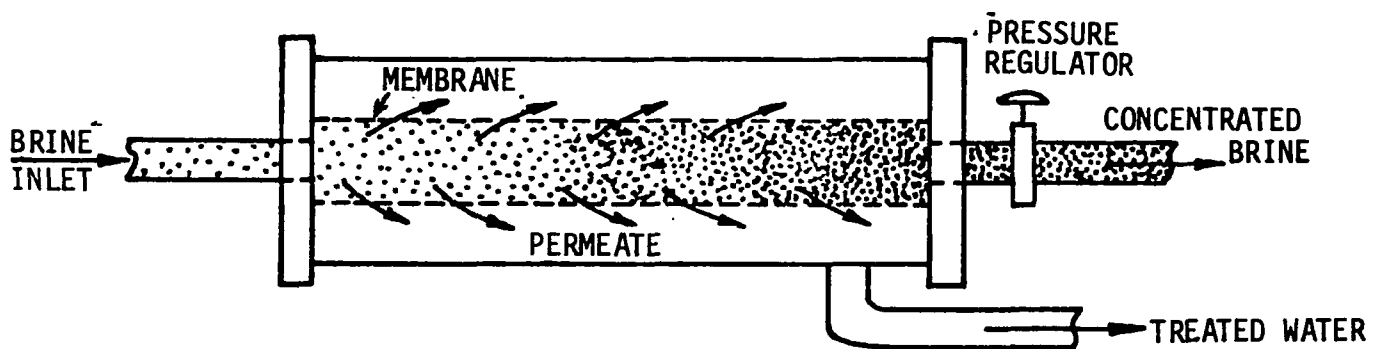
Figure 37. Schematic presentation of reverse osmosis.

Given a suitable membrane, the performance of a reverse osmosis unit is largely determined by the proportion of permeate to the retentate. As a larger fraction of the feed is removed as permeate, the concentration of the retentate increases. The increased concentration difference across the membrane tends to cause ion migration through the membrane. In general, reverse osmosis produces a permeate with a dissolved solids concentration approximately 10 percent of that in the feed solution. Industrial application of the process has shown the following removal efficiencies **(Argo, 1977, Chen, 1977, and Liptak, 1974):** SS 95-98 percent; BOD 90-95 percent; NH_3 95-99 percent; and org-N, NO_3N, PO_4-P, and TDS 95-99 percent.

Passage of individual ions varies according to the selectivity of the membrane, feed temperature, and pH. Water flux usually increases with increasing temperature, whereas salt rejection remains essentially constant over the normal operating temperature range of 15-30°C. The effect of pH on performance of the reverse osmosis unit is determined by membrane hydrolysis, which also influences salt rejection. Since the membrane is an organic ester, the rate of hydrolysis is pH dependent. Hydrolysis increases at both high and low ends of the pH scale. For this reason, a pH of 3 to 7 should be maintained for optimum membrane operation.

For reverse osmosis to be effective, it is essential that all large suspended particles be removed prior to its application. In addition, most membranes have a maximum tolerable temperature beyond which the membrane loses its effectiveness in retaining the dissolved constituents. Most commercial membranes have a maximum **temperature limitation of 200°F.** Geothermal fluids may require cooling prior to treatment by reverse osmosis.

Costs - Cost estimates for reverse osmosis were derived from a combination of studies prepared by the Fluids Systems Division of UOP, Inc. (UOP, 1974), Los Angeles County Sanitation (Chen, 1977), and the Orange County Water District Factory 21 (Argo, 1977). Figure 38 is a summary of cost estimates for the reverse osmosis system. The value plotted in the capital investment curve was calculated based on a 20-year plant life, using the capital recovery factor at 8 percent interest rate, 15 percent inflation, and 92 percent plant factor. The operation and maintenance costs include power, chemical, membrane replacement and maintenance, and labor costs. Cost data from the above references were adjusted to the 1977 second quarter costs by using the Marshall and Stevens Process Index (M&S) Index. The cost curves thus developed were found to be comparable to curves produced by Gumerman et al (1978).
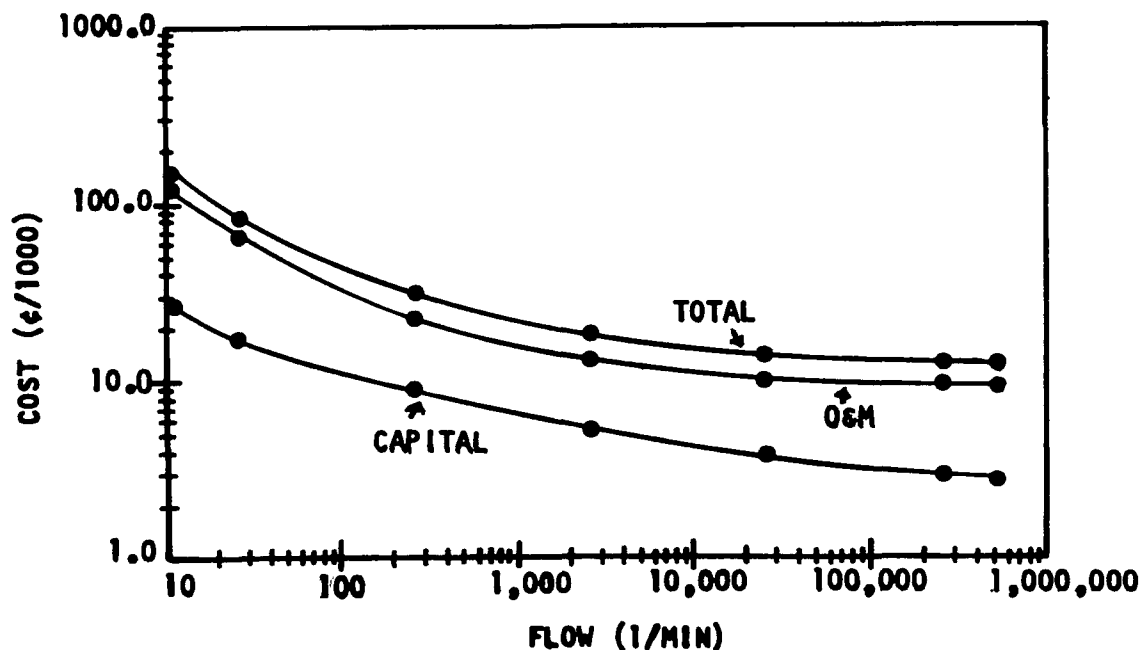
Figure 38. Cost estimate for reverse osmosis system.

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Costs of reverse osmosis depend largely upon the quality and quantity of wastewater to be treated. Pretreatment and disposal of residuals have not been included in the estimates. Membrane life is strongly influenced by the amount of total dissolved solids. The costs shown are for one stage. More than one stage may be required to achieve suitable effluent quality.

Electrodialysis

Process Description - This is an electrolytic process causing separation of ions in the presence of an imposed electrical field. Ions of opposite charge migrate through membranes toward their respective electrodes and the brine is separated into water and a concentrated brine (Chan, 1975). The basic principles of electrodialysis are illustrated in Figure 39.
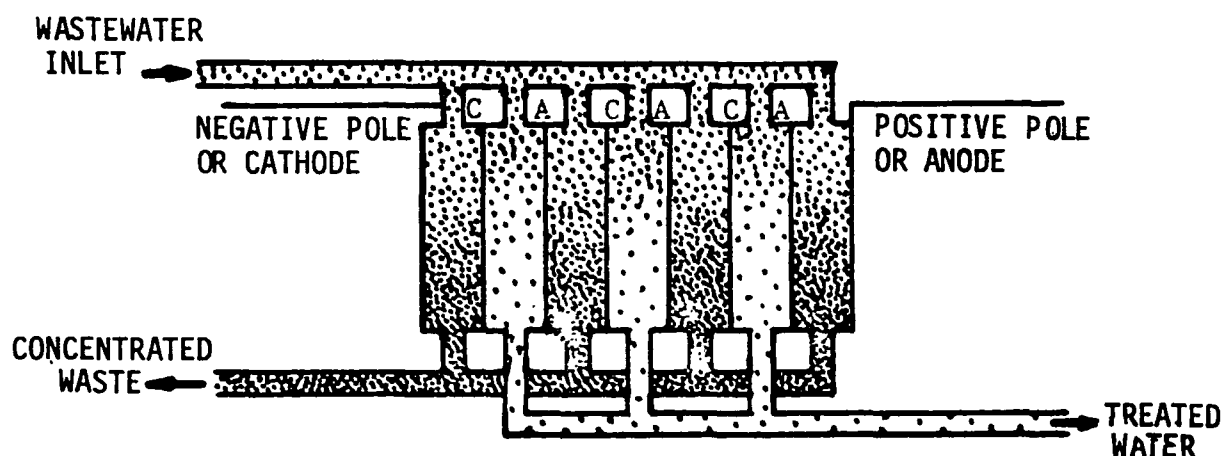
Figure 39.   Electrodialysis cell.

The electrodialysis system uses a series of compartments separated by alternately-placed anion and cation permeable membranes. The application of an electrical potential across the system causes the migration of cations to the cathode and of anions to the anode. The migration results in ion concentration and dilution in alternate compartments.

Electrodialysis as a wastewater treatment process is still in the preliminary development stages. It has been used for the desalination of brackish water, but has not been used extensively in the treatment of industrial wastes. As with reverse osmosis, electrodialysis produces a concentrate that, in turn, must be disposed of in some manner.

The membranes used in the process are subject to fouling by any suspended solids or oils (or other organics) in the waste. Such materials must be removed by pretreatment. Membrane life is usually determined experimentally. Electrodialysis has produced water having a total dissolved solids content of less than 500 ppm. This process is also found to be effective in removing 30 percent to 50 percent of NH_3-N and PO_4-P and approximately 40 percent of TDS (Bond, 1972).

Costs - For electrodialysis systems, the cost estimates were derived from information gathered in the San Francisco Bay-Delta Water Quality Control Program Study. Data points were extracted directly from the existing graphical plots. Units were converted to the metric system. These cost data (capital and O&M costs) were then updated to the present 1977 second quarter costs by again using the M & S Index. The plotted data (shown in Figure 40) were found to correlate relatively well with actual cost data presented by Los Angeles County Sanitation District (Dryden, 1970) and cost curves illustrated by Faber (Faber, 1972). Unfortunately the assumptions (amortization periods, interest rates, etc.) utilized in the original Bay-Delta Study on electrodialysis systems were not available for inclusion in this report; thus it was not possible to assess the accuracy and validity of these data points.
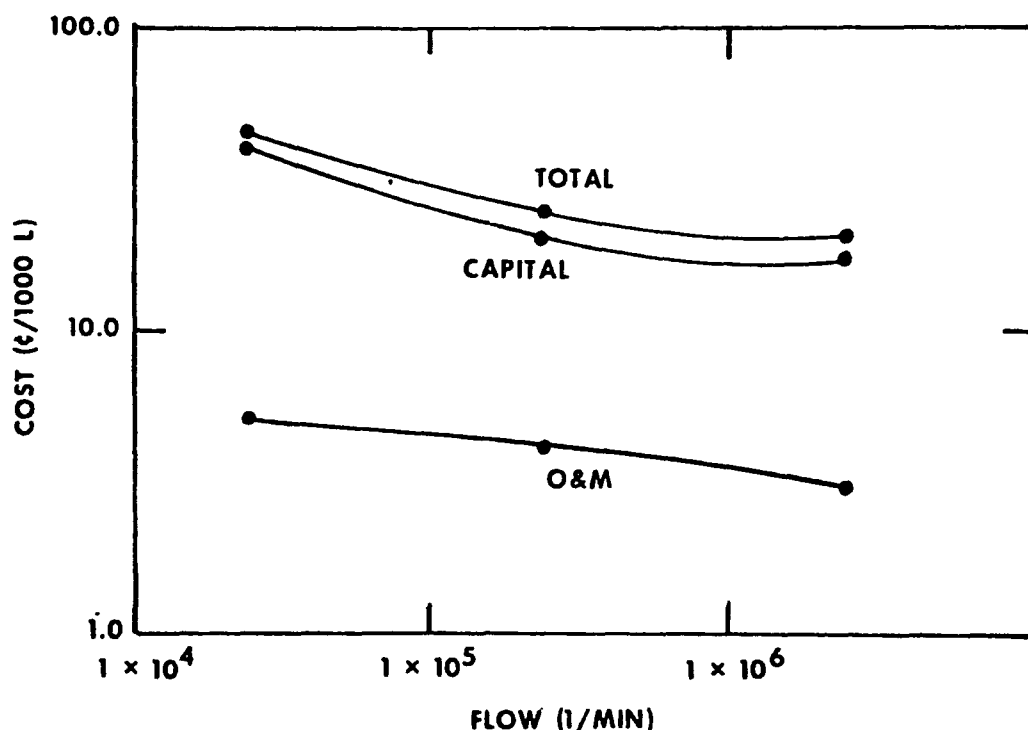
Figure 40. Cost estimates for electrodialysis system.

As in reverse osmosis, the cost of electrodialysis will depend primarily on the quality and quantity of wastewater to be treated. Pretreatment and residual disposal costs are not included. The costs shown are approximations for one stage. More than one stage might be required.

Ion Exchange

Process Description - This process involves the exchange of objectionable ions in the wastewater with non-objectionable ions such as H^+ or OH^- in the resin material (Chen, 1977). Most ion exchange materials are synthetic poly-

mers containing active groups such as HSO_3 and NH_4 to which the exchangeable ions (H^+ and OH^-) are attached. The exchange reaction for removing sodium chromate by a combination of cationic and anionic exchange resins can be represented by:

$$R^-H + Na^+ \rightarrow R^-Na^+ + H^+$$

$$R^{++}(OH-)_2 + CrO^{--}_4 \rightarrow R^{++}CrO^{--}_4 + 2\ OH^{--}$$
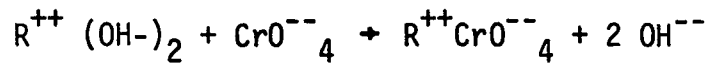
where R^- and R^{++} represent the **anionic and cationic** exchange material.

When the resins are operating on H^+ and OH^- cycles, treatment with ion exchange also results in the production of deionized water which can be used for process water or in other applications requiring a high quality water.

Demineralization by ion exchange is a process for removing inorganic salts and trace metals from wastewaters. In general, salts are composed of positive ions of a base and negative ions of an acid. These ions are removed in two stages: the positive ions by the cation exchanger and the negative ions by the anion exchanger. In the first stage the positive base ions, such as calcium (Ca), sodium (Na), or magnesium (Mg), are exchanged for hydrogen ions (H) in the cation exchange column, thereby converting these positive **cations into their** respective acids. In the second stage the acid negative ions such as silicates (SiO_3), carbonates (CO_3), chloride (Cl), or sulfate (SO_4) are removed and exchanged for hydroxide ions (OH) in the anion exchange column. This completes the two-step removal of the salt. In mixed-bed ion exchangers, as shown in Figure 41, **the two steps are combined into one.**

RAW WATER   DRAIN   ALKALI   AIR OUT

```
MIXED RESIN        ANION           ANION          MIXED RESIN
                   CATION          CATION
```

TREATED WATER   RAW WATER   ACID   AIR

DRAIN

SERVICE   BACKWASH   REGENERATION   RESIN MIXING
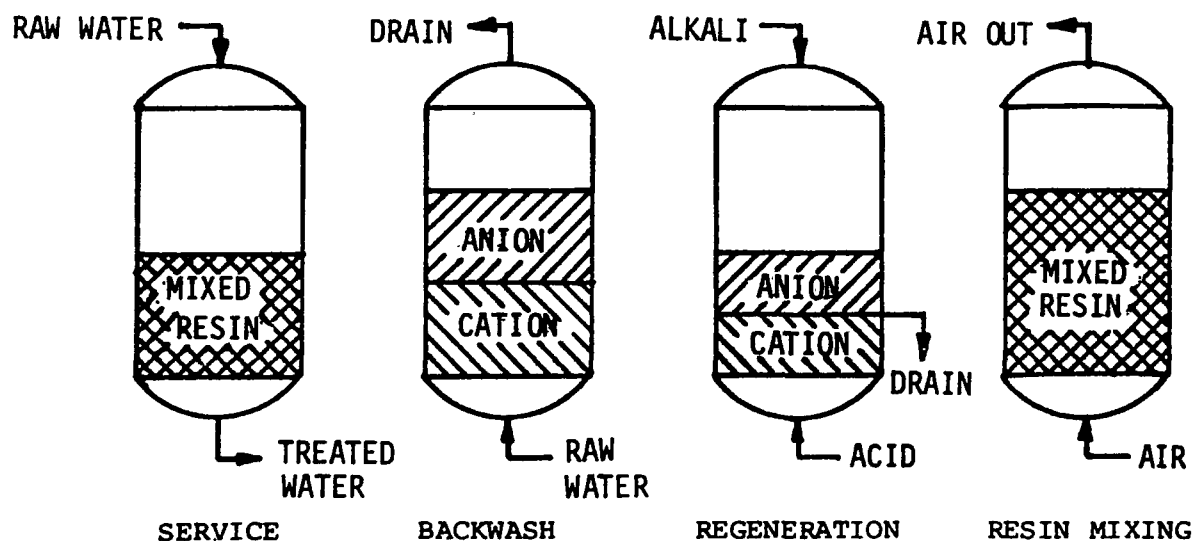
Figure 41.   Mixed-bed ion exchange process.

Once the demineralized ion exchangers are saturated or excessive leakage occurs they have to be regenerated to allow reuse of the resins. Cation exchangers are regenerated by strong acids (H_2SO_4 or HCl) and anion exchangers by caustic soda ($NaOH$). For continuity of operation during bed regeneration, two trains of ion exchange columns are needed.

Ideally, ion exchange columns can reduce a given pollutant concentration to essentially zero. In practical applications, depending on the type of resins used, removal efficiencies for total dissolved solids (TDS) have been reported in the range of 80 to 90 percent (Chen, 1977). Studies using weak electrolyte ion-exchange resins for the removal of ammonia and phenolics from foul-water condensates of refineries have shown promise.

Costs - The basis of the cost curves in Figure 42 for ion exchange systems is from Van Note's publication (Van Note, 1975). The assumptions and flow variable cost elements are presented in Tables 13 and 14. Chemical costs for regeneration are part of the O&M costs. The actual cost for ion exchange systems is dependent on the exchange resin, the characteristics of the wastewater, and the effluent quality required. Pretreatment costs or the cost for disposal of backwash brine have not been included.
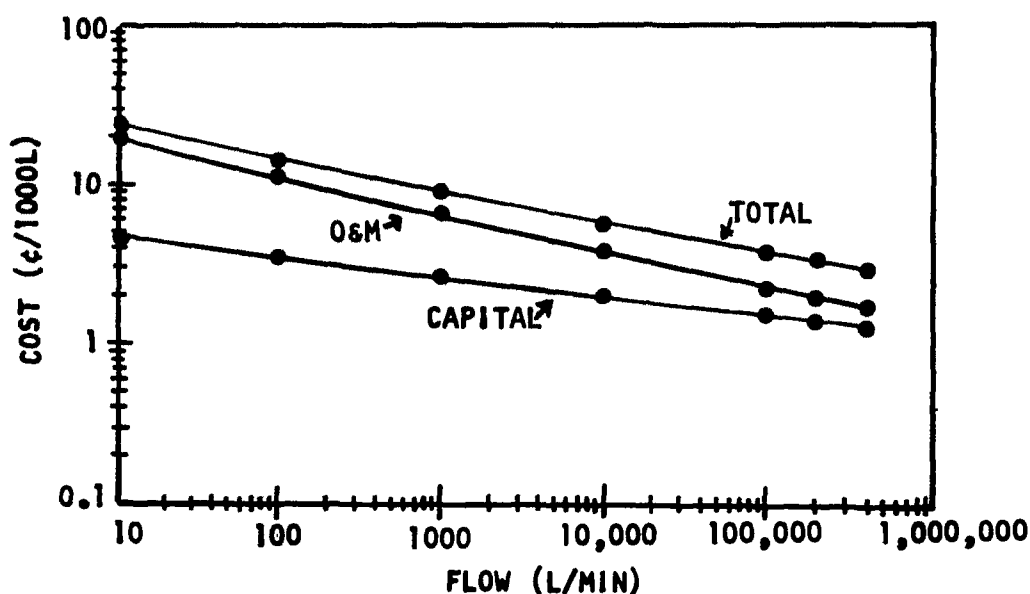
Figure 42. Cost estimates for ion exchange system.

Evaporation - Distillation

Process Description - In evaporation processes components of a liquid are separated by vaporization and condensation. Single- and multiple-effect evaporators are frequently used in the chemical industry to extract water from aqueous solutions.

Evaporators generally use steam as the heat source. Some evaporators may use several stages (termed "effects") to conserve heat. In multiple-effect evaporators (Figure 43), steam is introduced into the first effect in the series, and succeeding effects are operated at lower pressures so that steam condensed from the preceding stage can be used as the heat source in the next. Vapor condensation occurs on exchanger surfaces cooled by inlet water, which is warmed progressively through each stage. Reduced pressure is usually obtained by exhausting the vapor from the final effect to an aspirator-condenser, such as a jet condenser.
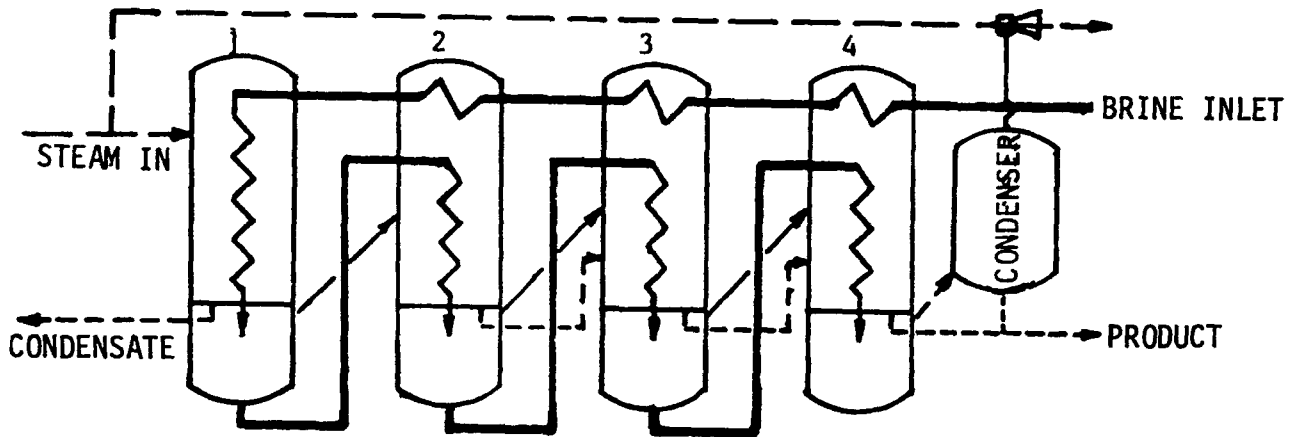
Figure 43. Multiple-effect evaporation.

The multiple-stage flash evaporation scheme places all steam heat exchange outside of the evaporation chambers, in a feed preheater. Distillate is flashed from the brine in each stage at successively lower temperatures and pressures (Figure 44). A test facility, using this technology on geothermal brines, is being operated by the Bureau of Reclamation, at East Mesa, California. Its objective is to produce fresh water for augmenting the Colorado River flow and for irrigation. Multiple flash evaporators are more economical than multi-effect units, and are frequently used in desalination applications.
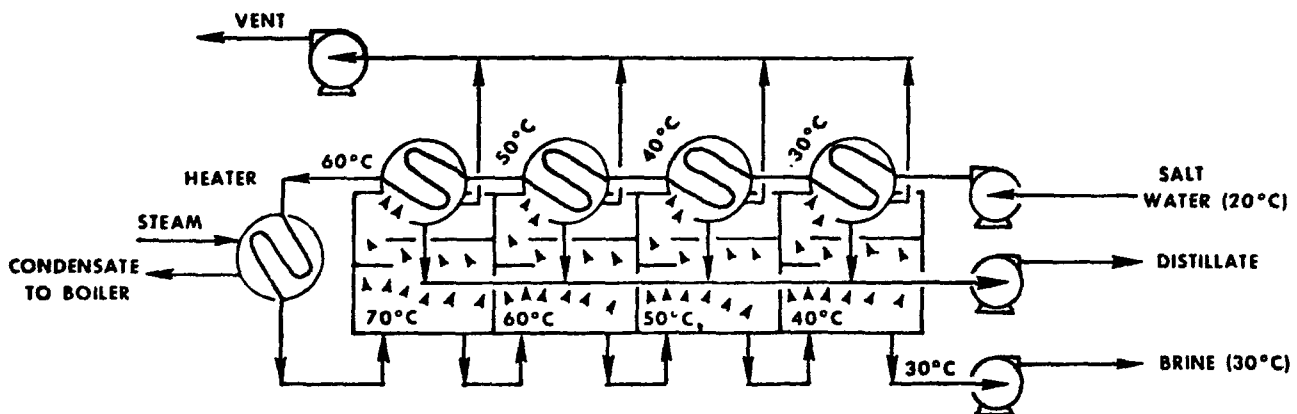
Figure 44. Multiple stage flash evaporation.

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Vapor recompression techniques can also be used to conserve heat (Figure 45). The vapor compression method uses mechanical rather than thermal energy, by compressing overhead vapor and using the compressed vapor as a heat exchange medium before it is discharged and used to preheat incoming feedwater. Compression stills may be economically attractive where cheap electrical power is available to drive the compressor. The effectiveness of this method is about the same as that of evaporation ponds, but it is faster and requires heat input.
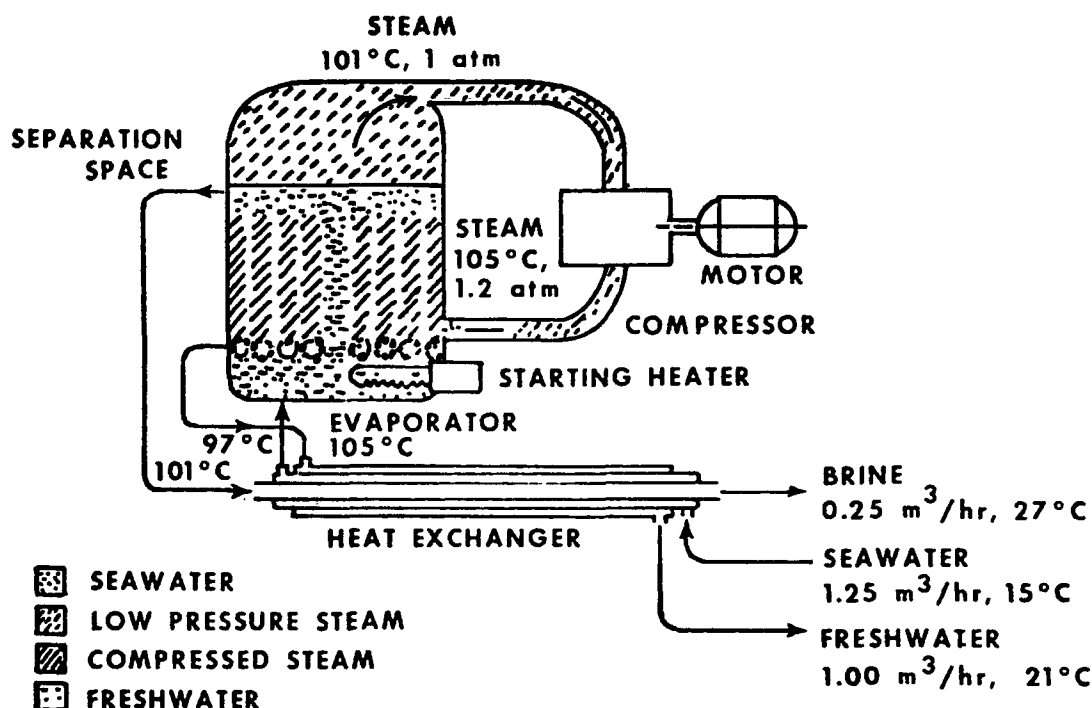
Figure 45. Compression still.

Evaporation methods are capable of reducing the volume of brine by 70 to 80 percent (Spiegler, 1966). The concentrated salt-brine residue must be properly disposed of by either ocean dumping, deep well injection, or after total evaporation, by landfill.

Costs - Evaporation systems costs are dominated by energy requirements, which are directly proportional to the amount of water to be evaporated. The cost of treatment per unit of flow decreases only slightly with increasing throughput (at a fixed percentage of feed to be evaporated). The total costs shown in Figure 46 are a composite of the operating costs and annualized capital costs using the following assumptions:

Electricity @ 4¢/KWH, steam @ $2/million Btu

8400 operating hours/year, over a 20-year project life (8 percent rate of return).

Capital and operating costs were obtained from experience in chemical and paper industry practice (Rosenblad, 1976; Guthrie, 1974; and Perry, 1973) for multi-effect and vapor recompression evaporators. Most cost data for multistage flash units are available from desalination installations (Howe, 1974; and Spiegler, 1966).
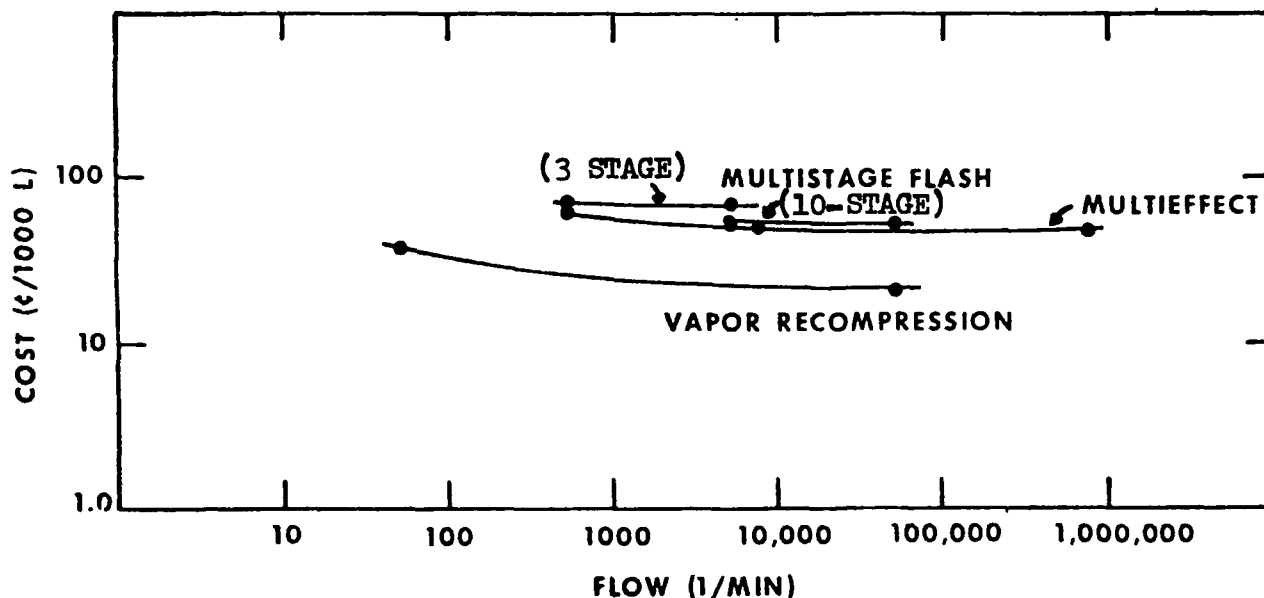
Figure 46. Total costs for evaporation. Basis: 50% of feed evaporated; 40°C feed temperature.

The efficiency of multiple stage evaporators (in terms of water produced per unit quantity of steam) improves with increasing number of stages of evaporation or flashing, and a total cost advantage is obtained from the use of ten stages versus three stages. Even so, the lower cost of vapor re-compression units is clearly evident at power costs of 4¢/KWH, and the cost could be even lower with cheap power available from an associated power plant. A cost breakdown of various evaporative systems for typical flow ranges is summarized in Table 15.

Variables which strongly affect evaporation costs include the percentage of feed to be evaporated; the inlet feed temperature; and (in the use of reduced pressure evaporation) the temperature of cooling water. The cost data shown in these figures are based on 50 percent evaporation of feed-water, with an incoming feed temperature of 40°C. There would be considerably more enthalpy (heat content) available in the incoming feed from the flash down of a geothermal power operation, and this extra enthalpy can be translated into increased amounts of evaporation over the base case. If, for example, an evaporator was designed to evaporate 30 percent of its feed (at 65°C), the same evaporator could yield about 75 percent evaporation at a feed temperature of 150°C, and 90 percent evaporation at 225°C.

TABLE 15. COMPONENT COSTS FOR EVAPORATION.

Basis: 50% of feed evaporated,
40°C feed temperature

System	Feed (liters/min)	O & M Costs, $	Annualized Installed cost, $	Total cost per year, $
Multi-stage flash;				
3 stage	520	175,000	15,360	190,360
	5,200	1,750,000	76,800	1,826,800
5 stage	5,200	1,312,500	134,400	1,446,900
	26,000	6,562,500	384;000	6,946,500
10 stage	520	105,000	57,600	162,600
	5,200	1,050,000	268,800	1,318,800
Multieffect,	7,584·	1,780,800	199,218	1,980,018
6 stage	75,840	17,808,000	300,000	18,108,000
Vapor compression	52	3,473	6,144	9,617
	52,000	3,473,433	1,920,000	5,393,433

The total costs for evaporation (20 to 70¢ per 1000 liters of feed) shown here are far more than for many competitive methods of wastewater treatment, and some amount of a high-salinity waste brine stream will always require disposal.

Other Wastewater Treatment Technologies

In addition to the above-described treatment technologies there are two processes that have been under investigation by the Office of Saline Waters (OSW) for desalination of ocean waters. These include direct freezing/gas hydration and liquid-liquid extraction processes. Direct freezing and the formation of gas hydrates have potential application for separating salt from sea water to produce potable water. However, freezing of high temperature geothermal fluids for the purpose of desalination has technical and economic constraints. Its application to treatment of geothermal wastewater cannot be considered a viable alternative at the present time.

Liquid-liquid extraction involves the use of a solvent (such as di-isopropyl amine-propane or N-butanol) to preferentially extract salt from saline water and subsequent evaporation and recovery of the solvent. The use of liquid-liquid extraction for desalting high temperature geothermal fluids would result in technical problems caused by the instability of solvents at high temperature. Its potential application to geothermal fluid treatment is definitely limited and cannot, currently, be considered feasible.

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Specific Chemical Constituents Abatement Technology

The wastewater control technologies presented in the previous subsection deal primarily with process effectiveness and applicability in the removal of gross constituents. Table 16 summarizes a survey of control technologies for the removal of specific pollutants from wastewaters. Since most literature findings are limited in information on specific pollutant removal, efforts were made to contact knowledgeable persons in the field, such as equipment vendors, engineering consultants, government regulatory agencies, and academia, to seek expert opinions on specific applications of pollutant removal from wastewater.

TABLE 16. REPORTED EFFICIENCIES OF CONTROL TECHNOLOGIES FOR
TREATMENT OF SPECIFIC CONSTITUENTS FROM WASTEWATERS
(percent removal)

	Sedimentation[1]	Chemical Precipitation[2]	Filtration[3]	Electro-dialysis[4]	Ion Exchange[5]	Reverse Osmosis[6]
TS	20-40	40-60	70-95	30-50	80-90	90-99
TDS	10	20-40	10	30-40	80-90	85-95
Fe	10-30	60-100	70-95	30-40	80-90	95-98
Mn	10-30	65.4-99.4	90-98	30-40	80-90	95-98
B	10	20-40	20-40	10	80-90	60-80
Zn	10-30	90-95	60-85	30-40	80-90	85-95
Ba	10	85-99	80-98	99.9	99	95-98
F	10	99	10	10	80-90	88-92
Pb	10-30	95-97	95-98	30-40	80-90	95-98
Cu	10-30	80-85	90-95	30-40	80-90	95-98
As	10-30	80-98	75-95	30-40	80-90	85-95
Hg	10-30	40-60	70-80	30-40	80-90	85-95
Se	10-30	80-90	90-95	30-40	99.7	85-95
Cr	10-30	60-99	60-99	30-40	80-90	85-95
Ag	10-30	90-99	90-99	30-40	85-95	85-95
Cd	10-30	85-95	90-98	30-40	80-90	85-95

Ref: (1) 6,40,63,72; (2) 6,7,23,40,44,60; (3) 40,60,63; (4) 11,40,60,63;
(5) 9,17,40,60,63; (6) 6,7,10,21,48,63

Application of Wastewater Treatment Technologies

Wastewater treatment requirements depend upon the characteristics of the raw wastewater compared to the quality to be maintained in the wastewater disposal area or receiving media. To examine the requirements, three sets of possible raw wastewater constituent characteristics and three sets of possible discharge requirements have been compiled in Table 17 based on EPA's work directive. The values shown are not intended to be actual, but probably include the ranges to be considered in geothermal wastewater treatment. Possible ranges of flows for various uses are shown in Table 18. Based on the information shown in Table 17, the required removal efficiencies were calculated, as shown in Table 19, for the various raw levels vs. discharge levels.

To simplify the regulatory requirements for achieving the removal efficiencies for each of the constituents in Table 19, it is assumed that the removal of total solids (TS) and the soluble metals (SM) with the most stringent

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TABLE 17. ASSUMED GEOTHERMAL WASTE BRINE AND SURFACE
WATER DISCHARGE CONCENTRATIONS (mg/1).

Constituent	Geothermal Waste Brine Concentration Level			Surface Water Discharge Concentration level		
	High	Mid	Low	High	Mid	Low
Total Solids	100,000	10,000	2,000	5,000	1,000	500
Iron	1,000	100	10	5.0	1.0	0.5
Manganese	1,000	10	1	1.0	0.1	0.05
Boron	500	10	1	5.0	2.0	1.0
Zinc	500	10	1	10	5.0	1.0
Barium	500	10	1	5.0	2.0	1.0
Fluoride	100	1	0.1	1.0	0.1	0.05
Lead	100	1	0.1	1.0	0.1	0.05
Copper	50	1	0.1	5.0	2.0	1.0
Arsenic	10	1	0.1	0.5	0.1	0.05
Mercury	10	0.1	0.01	0.01	0.005	0.002
Selenium	0.1	0.05	0.01	0.05	0.02	0.01
Chromium	10	0.1	0.01	0.5	0.1	0.05
Silver	1	0.1	0.01	0.5	0.1	0.05
Cadmium	1	0.1	0.01	0.05	0.02	0.01

TABLE 18. GEOTHERMAL WASTE BRINE FLOW RATES AND
LEVELS FOR VARIOUS USES.

Conversion System	Flow Rate Liters/Min.	Brine Conc. Levels
Direct Steam Power Generation	4,000-30,000	Mid & Low
Flashed Steam, Binary, Total Flow Power Generation	15,000-350,000	High, Mid & Low
Direct Heating Open & Closed	10-1,000	Mid & Low
Desalination	1,000-5,000	High & Mid

removal efficiency for a given level will concurrently meet all the necessary requirements for that level. This assumption is considered valid because the removal of TS to a specified level will also remove a proportional amount of suspended solids (silica and metal silicates) and dissolved solids (soluble metals, fluoride, etc.). Concurrently the removal of SM with the most stringent removal efficiency generally will also remove SM with less stringent requirements. The only exception is boron, which cannot be effectively removed by any current control technology.

TABLE 19. REMOVAL EFFICIENCIES (%) REQUIRED FOR TREATING VARIOUS LEVELS OF RAW GEOTHERMAL FLUIDS.

Constituent	High Level Waste 1	2	3	Mid Level Waste 1	2	3	Low Level Waste 1	2	3
Total Solids (TS)	95	99	99.5	50	90	95	0	50	75
Iron (Fe)	99.5	99.9	99.95	95	99	99.5	50	90	95
Manganese (Mn)	99.9	99.99	99.995	90	99	99.5	0	90	95
Boron (B)	99	99.6	99.8	50	80	90	0	0	0
Zinc (Zn)	98	99	99.8	0	50	90	0	0	0
Barium (Ba)	99	99.6	99.8	50	80	90	0	0	0
Fluoride (F)	99	99.9	99.95	0	90	95	0	0	50
Lead (Pb)	99	99.9	99.95	0	90	95	0	0	50
Copper (Cu)	90	96	98	0	0	0	0	0	0
Arsenic (As)	95	99	99.5	50	90	95	0	0	50
Mercury (Hg)	99.9	99.95	99.98	90	95	98	0	50	80
Selenium (Se)	50	80	90	0	60	80	0	0	0
Chromium (Cr)	95	99	99.5	0	0	50	0	0	0
Silver (Ag)	50	90	95	0	0	50	0	0	0
Cadmium (Cd)	95	98	99	50	80	90	0	0	0

To achieve the three effluent levels, an average value of efficiency was assigned to each of the treatment processes (Table 20). As the efficiencies of most treatment systems vary with the nature and flow conditions of the waste and the engineering design of the treatment processes, these arbitrarily assigned efficiencies are not to be interpreted as definitive efficiencies, but rather as an attempt to demonstrate the number of treatments required for achieving each of the specified effluent levels.

TABLE 20. ASSIGNED EFFICIENCIES OF VARIOUS TREATMENT SYSTEMS FOR REMOVING GROSS CONSTITUENTS.

	Efficiencies Total Solids	Soluble Metals
Sedimentation	30%	5%
Chemical Precipitation	50%	80%
Filtration	85%	85%
Electrodialysis	40%	35%
Ion Exchange	85%	90%
Reverse Osmosis	90%	90%
Evaporation	99.9%	95%

Applications of control technologies for achieving the three effluent level requirements from three levels of raw geothermal fluid are illustrated in Figures 47, 48 and 49. These figures depict the treatment units required that will attain each of the specified effluent levels. Implicit in these illustrations are the following assumptions: (1) pretreatment systems such as sedimentation, chemical precipitation, and filtration do not remove pollutants (TS or SM) more than the assigned efficiencies regardless of the number of identical process units utilized; (2) treatment such as reverse osmosis, ion exchange, or electrodialysis can remove pollutants at greater than the assigned efficiencies if a combination of unit stages is used, since the removal efficiencies are cumulative; (3) the sequence of treatment processes is arranged in a way so that residual pollutants are readily removed to their assigned efficiencies by succeeding unit processes; reversing the order of the treatment process sequence will invalidate the assigned efficiencies; and (4) alternative treatment systems may be developed to produce similar removal efficiencies.

As an example, Figure 47 presents block diagrams of the various treatment systems necessary for achieving the various assumed effluent quality levels from a high level waste. For level 1, the required removal efficiencies for both TS and SM are shown immediately below the level 1 requirement. Removal of 95 percent of the TS requires sedimentation, chemical precipitation, filtration and reverse osmosis. The percentage of TS removed from the system is depicted by the arrow pointing downward from the specific unit process. The percentage of TS remaining is shown by the arrow pointing to the right. Thus, 30 percent of TS is removed by sedimentation with 70 percent remaining in the treated waste. Of the 70 percent TS remaining, an additional 35 percent is removed by chemical precipitation. Effluent from the chemical treatment thus contains 33 percent TS. Filtration then removes another 29.75 percent TS, and reverse osmosis removes an additional 4.725 percent TS. At the end of this sequence of treatment, 99.47 percent TS removal has been achieved and only 0.525 percent TS remains in the treated effluent. A similar procedure can be followed for SM. These flow diagrams show that the treatment requirements for SM removal are always higher than or equal to those designed for TS removal. It appears logical, therefore, to assume that the effluent water quality requirements for each of the three levels should be governed by SM removal rather than TS removal.

WASTEWATER DISPOSAL TECHNOLOGIES

Wastewater from geothermal conversion operations will require disposal regardless of its quality or prior treatment. In general, the cleaner the wastewater, the easier and less expensive the disposal method. For example, effluents that meet water quality standards can simply be discharged to surface drainage. On the other hand, it is more expensive and more difficult to dispose of wastewater that does not meet such standards; it is these disposal methods with which this discussion is most concerned. It should be borne in mind, however, that these methods may also be used for reasons other than simply disposal; for example, injection may be practiced for geothermal reservoir conservation and subsidence prevention.

HIGH LEVEL WASTE

(A) Level 1 Requirement

T.S. = 95% removal; S.M. = 99.9% removal
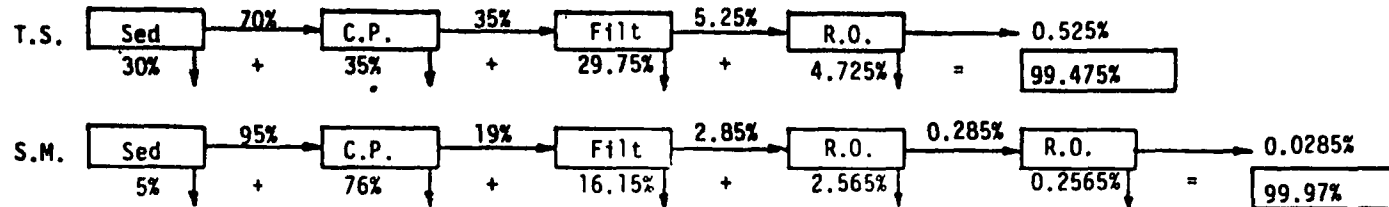
(B) Level 2 Requirement

T.S. = 99% removal; S.M. = 99.99%
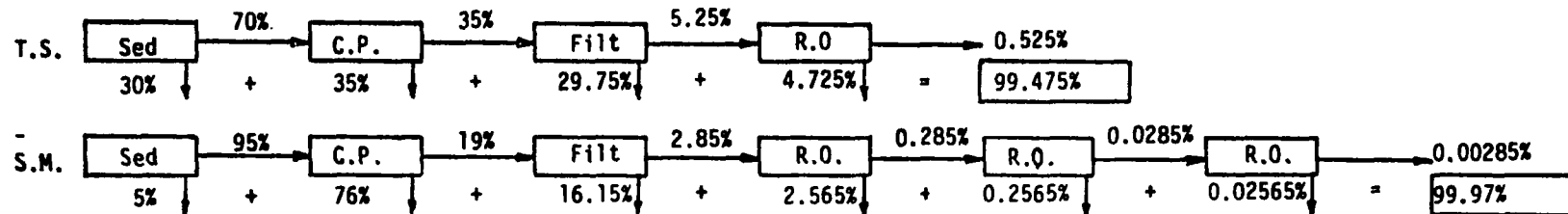
(C) Level 3 Requirement

T.S. = 99.5% removal; S.M. = 99.995% removal
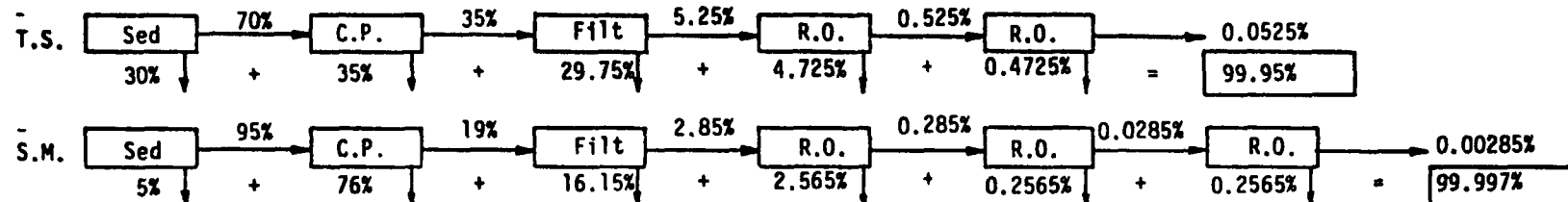
Legend: T.S.   = Total solids
         S.M.   = Soluble metal with most stringent requirement
         Sed   = Sedimentation
         C.P.   = Chemical precipitation
         Filt   = Filtration
         R.O.   = Reverse osmosis

Figure 47. Application of treatment technologies for achieving
three effluent quality levels from high level waste.

MID-LEVEL WASTE

(A) Level 1 Requirement

T.S. = 50% removal; S.M. = 95% removal

T.S. | Sed | --70%--> | C.P. | ----> 35%
30% + 35% = 65%

S.M. | Sed | --95%--> | C.P. | --19%--> | Filt | ----> 2.85%
5% + 76% + 16.15% = 97.15%
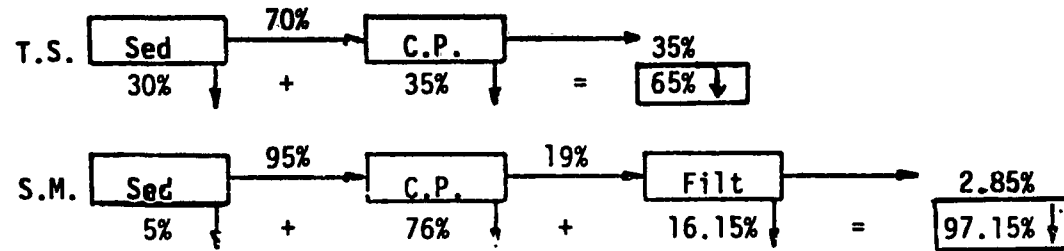
(B) Level 2 Requirement

T.S. = 90% removal; S.M. = 99% removal

T.S. | Sed | --70%--> | C.P. | --35%--> | Filt | ----> 5.25%
30% + 35% + 29.75% = 94.75%

S.M. | Sed | --95%--> | C.P. | --19%--> | Filt | --2.85%--> | R.O. | ----> 0.285%
5% + 76% + 16.15% + 2.565% = 99.715%

(C) Level 3 Requirement

T.S. = 95% removal; S.M. = 99.5%

T.S. | Sed | --70%--> | C.P. | --35%--> | Filt | --5.25%--> | R.O. | ----> 0.525%
30% + 35% + 29.75% + 4.725% = 99.475%
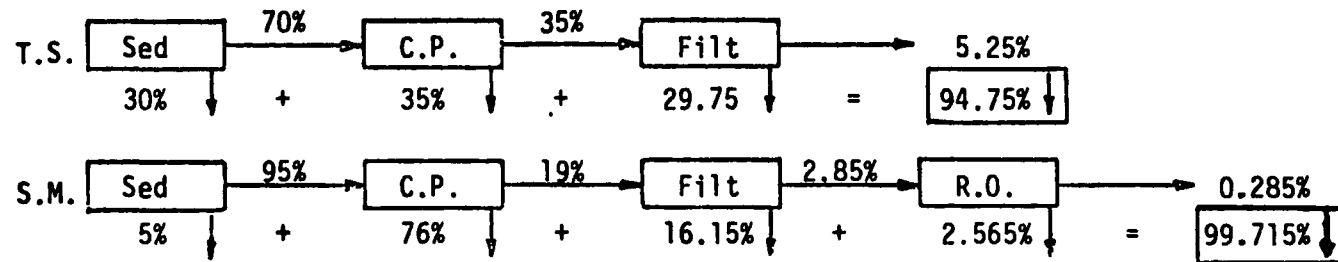
S.M. | Sed | --95%--> | C.P. | --19%--> | Filt | --2.85%--> | R.O. | ----> 0.285%
5% + 76% + 16.15% + 2.565% = 99.715%
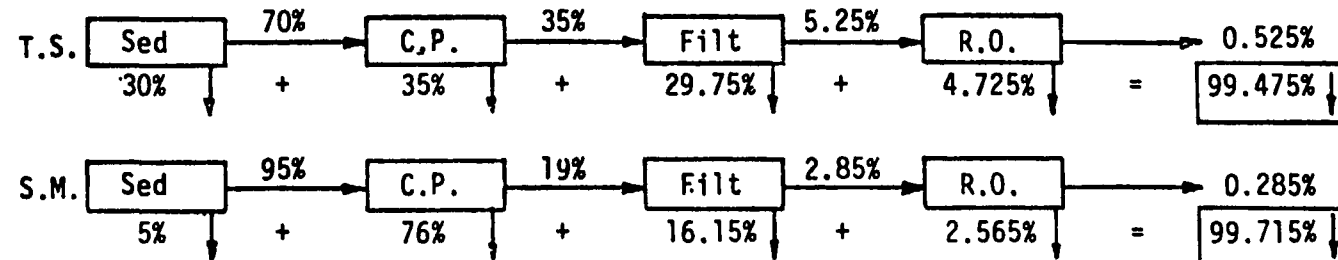
Figure 48. Application of treatment technologies for achieving
three effluent quality levels from mid level waste.

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LOW LEVEL WASTE

(A) Level 1 Requirement

T.S. = 0 removal; S.M. = 50% removal

```
          95%
┌─────┐          ┌─────┐              19%
│ Sed ├─────────►│ C.P.├──────────►      ┌──────┐
└─────┘          └─────┘              │ 81% ↓│
  5%    ↓    +     76%   ↓     =      └──────┘
```
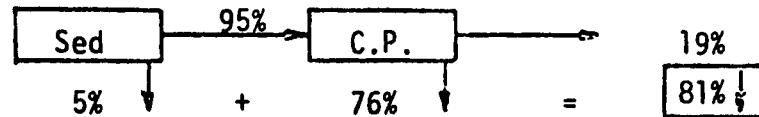
(B) Level 2 Requirement

T.S. = 50% removal; S.M. = 90% removal

```
          70%
┌─────┐          ┌──────┐             35%
│ Sed ├─────────►│ C.P. ├──────────►      ┌──────┐
└─────┘          └──────┘             │ 65% ↓│
  30%   ↓    +     35%   ↓     =      └──────┘
```

```
          95%            19%
┌─────┐          ┌─────┐          ┌─────┐          2.85%
│ Sed ├─────────►│ C.P.├─────────►│Filt ├────────►   ┌─────────┐
└─────┘          └─────┘          └─────┘          │ 97.15% ↓│
  5%    ↓    +     76%   ↓    +    16.15% ↓   =    └─────────┘
```
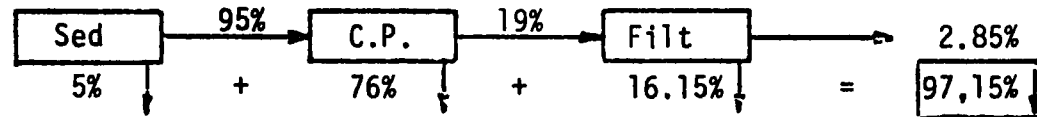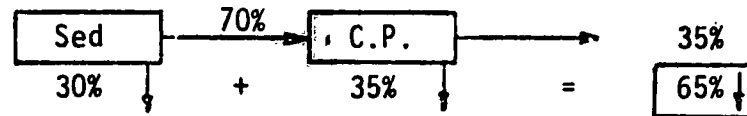
(C) Level 3 Requirement

T.S. = 75% removal; S.M. = 95% removal

```
          70%            35%
┌─────┐          ┌─────┐          ┌─────┐          5.25%
│ Sed ├─────────►│ C.P.├─────────►│Filt ├────────►   ┌─────────┐
└─────┘          └─────┘          └─────┘          │ 94.75% ↓│
  30%   ↓    +     35%   ↓    +    29.75% ↓   =    └─────────┘
```

```
          95%            19%
┌─────┐          ┌─────┐          ┌─────┐          2.85%
│ Sed ├─────────►│ C.P.├─────────►│Filt ├────────►   ┌─────────┐
└─────┘          └─────┘          └─────┘          │ 97.15% ↓│
  5%    ↓    +     76%   ↓    +    16.15% ↓   =    └─────────┘
```
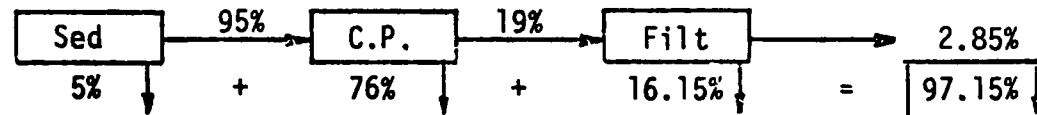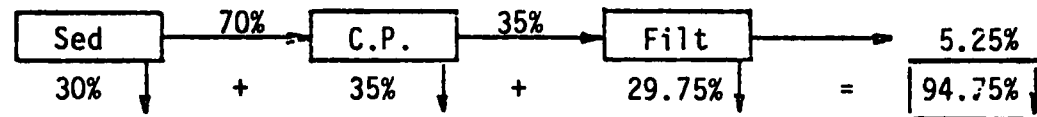
Figure 49. Application of treatment technologies for achieving three effluent quality levels from low level waste.

Subsurface Injection

Technology Description - Successful subsurface injection tests have been
performed in a number of geothermal fields in the United States and abroad:
for example, The Geysers, East Mesa, Niland, and Heber fields in California;
The Valles Caldera field in New Mexico; the Matsukawa and Otake fields in
Japan; The Wairakei field in New Zealand; the Ahuachapan field in El Salvador,
etc. In The Geysers field, return of steam condensate to the geothermal reser-
voir by injection was started in 1969; about eight billion gallons of conden-
sate have been injected to date. The current daily rate of injection is about
5 million gallons. Besides geothermal, many other industries have adopted sub-
surface injection of liquid wastes to prevent or control water pollution. The
practice is widespread in oil production fields. There are several reasons
for choosing subsurface injection as a disposal method. Some of these follow:

- Alternatives to injection are isolating the waste from the
  surface environment and releasing the waste into surface water
  bodies. Surface isolation of large quantities of liquid waste
  generated by geothermal operation is difficult. In most cases,
  before the liquid waste can be released into surface water bodies,
  it will require costly treatment. Treatment will create secondary
  wastes, also requiring disposal.

- Failure to replace reservoir fluid may allow ground subsidence.
  Subsidence has been observed in the geothermal fields at Cerro
  Prieto, Mexico, and Wairakai, New Zealand, where fluid injection
  has not been practiced.

- If reservoir fluid is not replaced, the reservoir pressure may
  decline, unless there is rapid and complete natural recharge.
  Evidence of complete natural recharge is rare. Any decline in
  reservoir pressure causes a decline in the productivity of the
  production wells.

- Injected, cooled geothermal wastewater scavenges heat from the
  reservoir rock matrix and may be withdrawn again at the production
  wells. Injected steam condensate may be reproduced as steam.
  Injection of geothermal waste into the producing formation allows
  a higher recovery of heat stored in the reservoir.

- Injection into geothermal reservoirs is an effective means of
  preventing not only chemical, but also thermal, pollution of
  surface water bodies.

Subsurface injection, if the geothermal fluid is utilized in an open sys-
tem, will generally be preceded by settling in ponds or tanks to remove sus-
pended solids. Sometimes filters may be used for this purpose. The wastewater
may also require chemical or physical deaeration to reduce its corrosiveness.
Finally, it is injected into the geothermal reservoir through the injection
well. Injection may sometimes be accomplished by gravity alone, without the

need for pumping the waste down the well, because of the higher gravity head of the cooler, denser geothermal waste.

Old production wells may be converted to injection wells. However, wells may be drilled solely for injection. Unless the geothermal reservoir rock is very competent (structurally self-supporting), a cased hole with slotted liner in the injection zone is used. Figure 50 is a schematic diagram of a typical injection well at The Geysers.
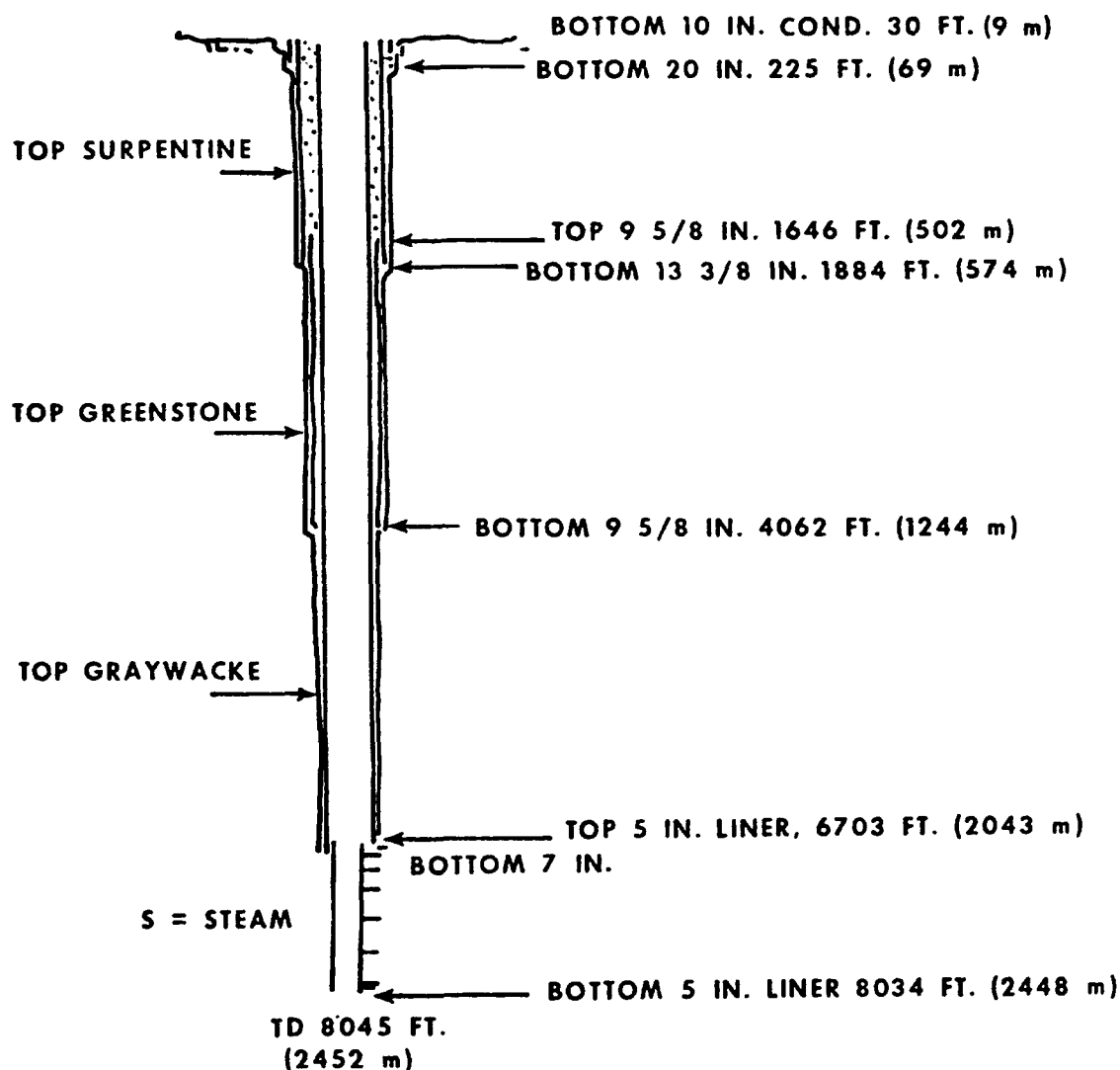
Figure 50. Typical injection well set-up

The injection scheme should be designed to optimize the travel path and time of flow between injection wells and producing wells, thus preventing rapid cooling of the production water. At the same time, the water should be injected sufficiently into the producing reservoir to minimize the decline in reservoir pressure. The key factor in determining the optimum injection plan is the spatial variation of water temperature and permeability in the reservoir.

Cooling and pressure decline around the injection wellbore may cause formation plugging by the deposition of dissolved and suspended solids, and thus increase resistance to injection. In order to maintain the injection rate, pressure must then be increased. Increase in injection pressure increases operating cost and mechanical problems. If the injection system reaches its maximum pressure capacity, more injection wells may need to be drilled, or the old wells stimulated, to maintain the total injection rate, thus escalating costs. There is no simple way yet to estimate loss of injectivity with time. The only sure means of assessing injection potential is to inject continuously for an extended period, at least a few months, and monitor wellhead injection pressure versus flow rate.

Injection wells should be completed carefully to isolate the injection horizon from shallow, fresh water aquifers. Any abandoned well near an injection well may provide a pathway for movement of the waste to shallow fresh water aquifers (Ostroot, 1972). Inadequate cementing behind casings and/or corrosion of liners can result in upward migration of water from geothermal reservoirs.

Surface pretreatment of the wastewater from geothermal operations may be needed to ensure success of a subsurface disposal operation. Generally the pretreatment would involve one or more of the following (Sadow, 1972):

● storage in impervious impoundments to permit, under quiescent conditions, settling and physical separation of the unwanted components;

● corrosion control by proper pH control, deaeration, and use of inhibitors;

● coagulation and clarification to accelerate gravity sedimentation;

● filtration and addition of bactericide to prevent plugging by bacterial growth; and

● pH and/or temperature control to reduce scaling.

Injection Well Cost Estimates

Capital costs for injection include the costs for drilling, casing and cementing, logging, perforation, well head equipment (including pumps and piping), control systems, and engineering supervision. Operation and maintenance costs consist of expenditures for the operation and routine maintenance of wellhead equipment, piping and pumps.

Capital Costs - The capital cost of an injection well may be estimated after determining the following well parameters:
- hole and pipe diameters
- the pumping system required (number, type, rating)
- depth of wells
- number of wells
- hydrology and geology of site

Once these data are known, the system design may be developed, and costs may be estimated based on design specifications for the depth and diameter of the well, the pumping requirements in terms of flow rate and pressure, and the drilling equipment and procedure.

The variation of cost per well with well diameter is shown in Figure 51. Pump cost may vary by over 100% per well, depending on the rating and material requirements. The depth of the well and the number of wells affects pipe costs and the time required for drilling, as will site hydrology and geology.

For a given geologic formation, the cost per unit of depth for drilling an injection well increases with depth. The relationship between drilling cost and well depth has not been clearly established for geothermal applications. Based on one study by Geonomics, Inc. in 1976, the injection well costs for sedimentary lithology vary between $250,000 at 5,000 ft (1,524 m) depth to approximately $750,000 at 10,000 ft (3,048 m) depth. Translated to a cost per unit depth basis, the drilling and completion well cost varies between $150-$300 per meter depth in 1976 dollars (Geonomics, 1977). These costs are affected greatly by the site lithology. For example, the capital cost of drilling and injection per unit of depth in volcanic formations may be 60 to 70 percent higher than in sedimentary formation. Unfortunately, few data are presently available relating drilling costs and lithology.

Because of the wide variations in site-specific geology and hydrology, and lack of complete data characterizing existing wells, injection well cost data have not been usefully parameterized in terms of the various cost determinants. In the absence of such cost data, the capital cost of injection wells was derived by a simplistic approach involving the selection of a representative well cost using empirical cost data for actual wells. This representative cost was then used to develop total costs for multiple well systems capable of injecting various wastewater flow rates generated by the four

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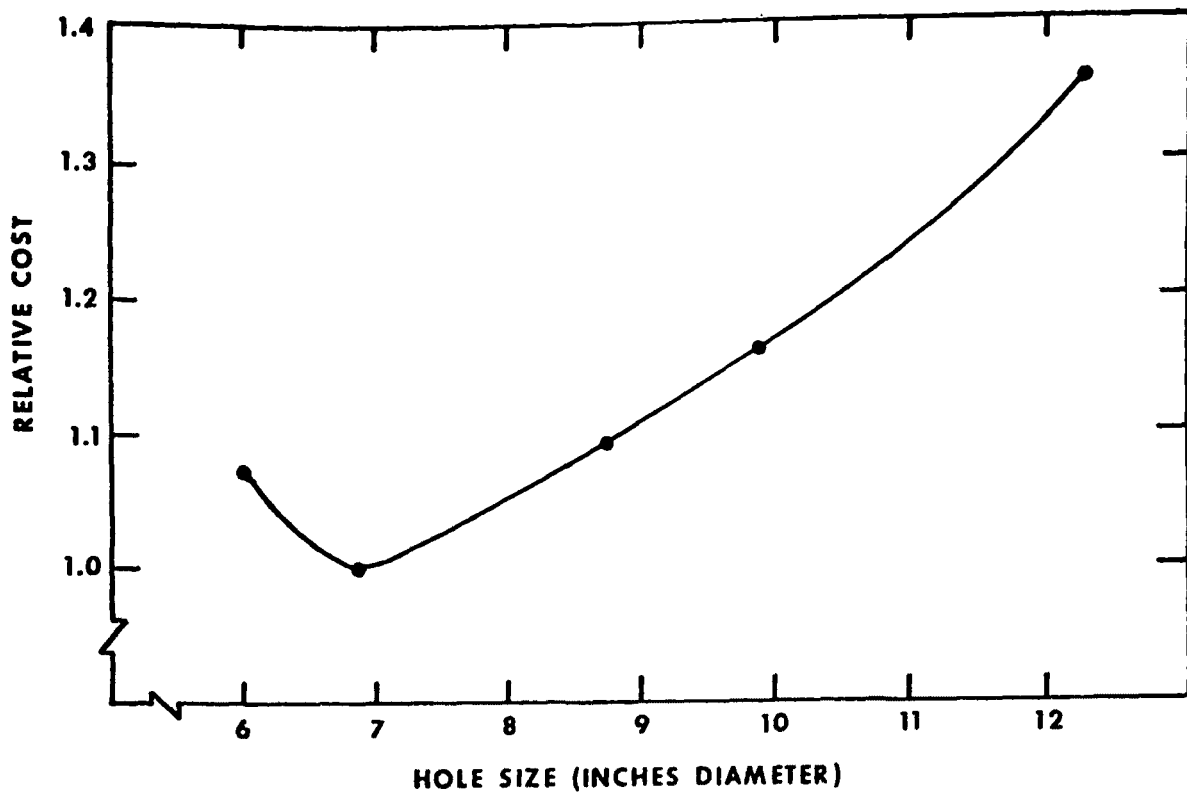
Figure 51. Well hole size cost comparison (capital cost only).

energy conversion processes. The total cost of the well system was also estimated for four selected well capacities representative of existing capacities.

Table 21 summarizes the injection well cost data surveyed as the basis for capital cost estimates developed in this report. The capital investment for an injection well varies from $400,000 to $1,000,000. The individual construction costs also vary widely. Based on inspection of Table 21, the average capital cost of an injection well was taken as $500,000. The cost of individual construction elements were also selected, and are shown in Table 22. The well depth associated with the selected cost data varies between 3000 and 10000 ft (915-3050 m), averaging about 6000 ft (1830 m).

The flow capacity selected for a well is based on inspection of the results of a 1971 survey of facilities using injection well systems to dispose of liquid waste as shown in Table 23. At that time, 82% of the injection wells were at refineries, chemical plants, and steel mills. The survey shows that the potential flow for an aquifer can be quite high, although the median indicates that the bias of the survey data is definitely in favor of the lower flows. The results of a 1970 survey, based on 75 injection facilities, are shown in Table 24. The results are grouped in different ranges of depth, injection rate, and injection pressure, and

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TABLE 21. CAPITAL COST OF INJECTION WELLS, DATA FROM THE LITERATURE

General Information and Cost Parameters	Reference Number									
	59.	4.	32.	32.	62.	47.	47.	47.	47.	47.
Date Published	May 77	1973	July 76	July 76	Dec. 77	c.	c.	c.	c.	c.
Type Well	Steam Production	Steam Production	Injection	Injection	Indection	Acid Waste Disposal	Acid Waste Disposal	Acid Waste Disposal	Acid Waste Disposal	Acid Waste Disposal
Location					Imperial Valley,Ca.	E. Ohio	W. Pa.	W. Pa.	SW. NY	SW. NY
Diameter (in.)										
Total Vertical Depth(ft.)						3,300	4,800	6,000	3,060	4,300
Straight(S) or Directional(D)										
Construction Cost										
• Drilling Contr'or Cost										
• Mud Exp.		$60,000	$128,880	$210,058						
• Casing & Tubing			23,650	36,600						
• Cementing			61,972	77,212						
• Logging			39,300	39,300						
• Perforation			22,215	30,434						
• Well Head Equipment										
• Engineering Supervision			6,820	6,820						
• Control Sys.			6,050	8,800						
Injection Pump									$502,000	$447,000
Injection Pipeline					$500,000	$564,000	$960,000	$770,000		
Total Capital Cost										

c Reference publication date is 1974

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TABLE 21. (Continued)

General Information and Cost Parameters	Reference Number								
	56.	56.	26.	19.	33.	16.	16.	16.	16.
Date Published	Oct. 76	Oct. 76	Dec. 74		May 77	June 77	June 77	June 77	June 77
Type Well	Injection	Injection	Steam Production		Steam Production	Injection	Injection	Injection	Inject
Location	East Mesa, Ca.			None Specific'ly	The Geysers, Ca.	Roosevelt Hot Spr.Ut	Coso Hot Sprg.Ca.	East Mesa Ca.	Long Valley,Ca.
Diameter (in.)			9 5/8	9 7/8					
Total Vertical Depth (ft.)			7,000	10,000	7,000 to 8,000				
Straight(S) or Directional (D)	S	D (48°)	D (30°)						
Construction Cost	$403,077	$535,448							
• Drilling Contractor Cost			$169,000		$658,000				
• Mud Expense			20,000		30,000				
• Casing & Tubing			10,000		124,500				
• Cementing			25,000		50,000				
• Logging			23,000		33,000				
• Perforation			40,000						
• Well Head Equipment					20,000				
• Engineering Supervision					28,000[a]				
• Control Sys.	22,831								
Injection Pump	54,200								
Injection Pipeline	224,000					$700,000	$700,000	$400,000	$700,000
Total Capital Cost					$1,003,500[b]				

(Continued)

[a] Includes overhead

[b] Includes miscellaneous costs amounting to $60,000

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TABLE 22. AVERAGE CAPITAL COST FOR AN INJECTION WELL

Capital Cost Parameters	Cost
Construction Costs	
Drilling Contractor Cost	$170,000
Mud Expense or Air Equipment Rental	20,000
Casing and Tubing (including accessories)	70,000
Cementing	40,000
Logging	20,000
Perforating	50,000
Well Head Equipment	40,000
Engineering Supervision	10,000
Control System	30,000
Injection Pump	50,000
Injection Pipeline	0 [a]
TOTAL CAPITAL COST	$500,000 [b]

[a] No pipeline is required if directional drilling is used. This component could cost several hundred thousand dollars[56] if required.

[b] The total capital cost varies, in actual practice, from approximately $300,000[32] to $1 million[33]. $500,000 is considered to be a reasonable average.

are consistent with the results in Table 23. On the basis of the range of well capacities indicated, four well capacities were selected for cost analysis: 200, 1000, 4000, and 8000 l/min.

The calculation of capital cost of various multiple well systems which achieve the expected geothermal wastewater generation rates (up to 350,000 l/min) is shown in Table 25. The total capital cost is determined from the number of wells required to achieve the required disposal flow rate. Total capital cost is then amortized over a 30-year period based on an 8% interest rate. Replacement of equipment is considered negligible compared to drilling costs. The demand factor has been assumed to be 80% (i.e., the system is not operating 20% of the time).

The annualized costs for the multiple injection well systems are shown in Figure 52, normalized to each 1000 liters of wastewater flow. Four curves are shown, each representing an injection well system utilizing one of the selected capacities. Clearly, the injection system consisting of larger wells is more economical, since fewer wells must be drilled to accomplish the required disposal rate. However, some caution should be exercised in applying

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TABLE 23. SURVEY OF 124 INJECTION WELLS FOR DISPOSAL OF LIQUID WASTE
(NIPCC, 1971)

	Maximum	Minimum	Median
Depth (m)	3890	90	810
Injection Zone			
Depth to Top (m)	3650	61	625
Thickness (m)	640	1.5	56
Injection Rate (lpm)	16300	0.57	512
Injection Pressure (psi)	4000	0	185

TABLE 24. SURVEY OF 75 INJECTION WELLS FOR DISPOSAL OF LIQUID WASTE
(NIPCC, 1971)

Physical Parameters	Percent of Total Wells
Depths of Well	
0-1000 feet (0-305 meters)	7
1000-2000 feet (305-610 meters)	29
2000-4000 feet (610-1220 meters)	22
4000-6000 feet (1220-1830 meters)	31
6000-12,000 feet (1830-3660 meters)	9
greater than 12,000 feet	2

Injection Rate		
gpm	**lpm**	
0-50	0-190	27
50-100	190-379	17
100-200	379-758	25
200-400	758-1516	26
400-800	1516-3032	4
greater than 800	greater than 3032	1

Injection Pressure, psi	
partial vacuum	14
0-150	29
150-300	27
300-600	9
600-1500	20
greater than 1500	1

TABLE 25. CAPITAL COSTS FOR INJECTION SYSTEMS AT FOUR WELL CAPACITIES

Flow	Well Capacity: 200 lpm/well			Well Capacity: 1000 lpm/well			Well Capacity: 4000 lpm/well			Well Capacity: 8000 lpm/well		
	No. Wells Req'd.	Initial Capital Cost	Annualized Cost Per Unit of Flow[b]	No. Wells Req'd.	Initial Capital Cost	Annualized Cost per Unit of Flow[b]	No. Wells Req'd.	Initial Capital Cost	Annualized Cost Per Unit of Flow[b]	No. Wells Req'd	Initial Capital Cost	Annualized Cost Per Unit of Flow[b]
(liters/min.)		($10^6)	($/1000L)	($/1000L)	($/1000L)	($/1000L)		($/1000L)	($/1000L)		($/1000L)	($/1000L)
10	1	0.5	10.60									
100		0.5	1.06	1	0.5	1.06						
500	2	1.0	0.42	1	0.5	0.212						
1000	5	2.5	0.53	1	0.5	0.106	1	0.5	0.106			
4000	20	10.	0.53	4	2.0	0.106	1	0.5	0.0264	1	0.5	0.0264
5000	25[a]	12.5	0.53	5	2.5	0.106	2	1.0	0.0422	1	0.5	0.0211
10,000				10	5.0	0.106	3	1.5	0.0316	2	1.0	0.0211
15,000				15	7.5	0.106	4	2.0	0.0281	2	1.0	0.0141
30,000				30[a]	15.	0.106	8	4.0	0.0281	4	2.0	0.0141
50,000							13	6.5	0.0274	7	3.5	0.0148
100,000							25[a]	12.5	0.0264	13	6.5	0.0137
350,000										44	22.	0.0133

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a. Arbitrary limit

b. Total cost is annualized based on C = P(CRF), where P - total cost and CRF is capital recovery factor at 8% interest and 30 year period. The demand factor for the well is assumed to be 80%.

the data in Figure 52. The plots are predicated on the assumption that capital cost of a well is invariant regardless of its capacity. Actually, wells of different capacities, all other site parameters being equal, may use different pumps, different pump injection pressure, or a different hole size. If a different hole size is used to accomplish additional flow capacity, such as from 4000 to 8000 1/min, the relative cost of the larger diameter well would be about 1.2 times that of the smaller well (see Figure 51). Still, the cost information developed in Figure 52 is useful to establish preliminary cost estimates and to judge feasibility of multiple well injection systems. For example, it can be seen that the total annualized capital cost of injecting high flow levels of geothermal wastewater from flashed steam plants is relatively low compared to other environmentally acceptable disposal methods. The annualized capital cost is determined to be only $2 million per year at 350,000 lpm injection, using 44 wells, each with a capacity of 8,000 lpm.
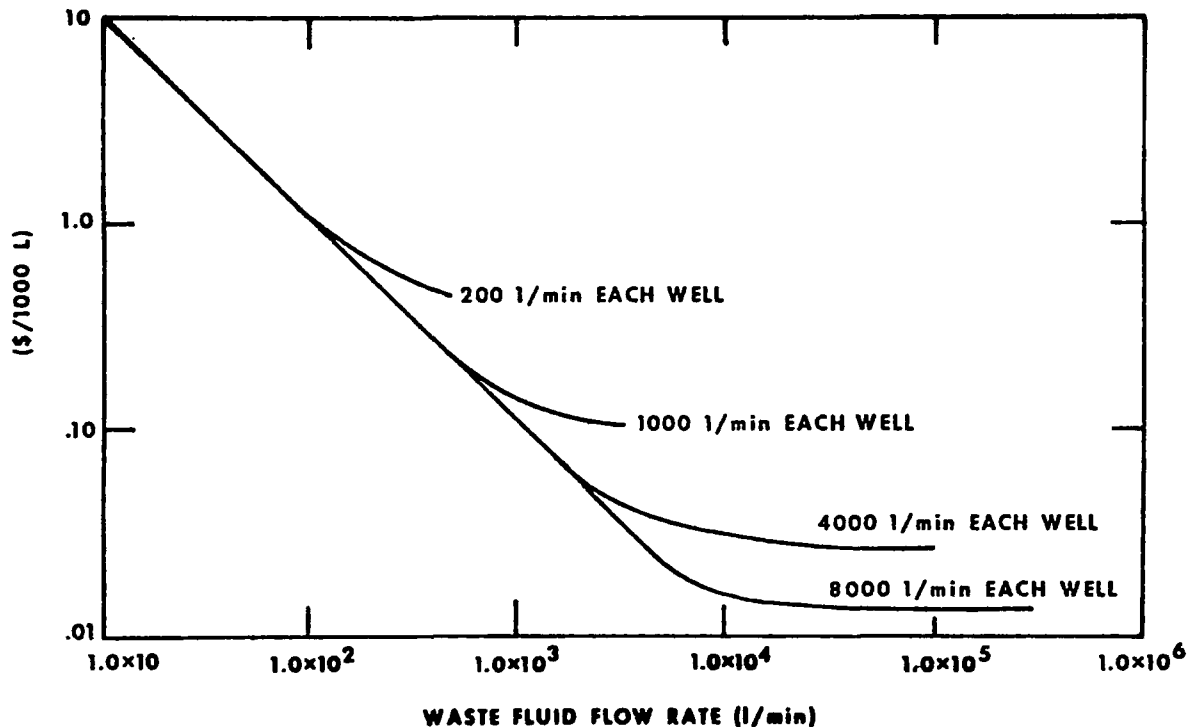
Figure 52. Annualized capital cost for injection of geothermal wastewaters.

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Operation and Maintenance (O&M) Costs - The operating cost for an injection system will consist mostly of the energy cost for pumping. Routine labor costs will be negligible, and maintenance costs over a thirty-year period depend primarily on the application and service required. In many cases, repair costs will be almost zero, while in others, anticipated maintenance or repair (due to corrosion, plugging, or wear) will prohibit the use of injection entirely. For the purpose of costing, it is assumed that 4.0% of the capital investment is annual maintenance costs. This corresponds to $20,000 per well. Depending on the flow temperature and the operating pressure of the injection well, this maintenance may vary somewhat.

Energy costs for pumping, in cents per thousand liters, are independent of flow rate, but are instead, a function of the pressure requirement for the particular injection system. This pressure requirement depends on frictional losses in the tubing, elevation changes for the pumped fluid, and the hydrologic pressure requirement (that is, the pressure required to push the waste liquid into the injection aquifer).

Frictional losses in the well tubing area are usually negligible. For a flow rate of 8000 liters per minute, the losses are 1.3 psi per 100 feet (30.5 m) of tubing. On the other hand, the pressure gain due to the elevation head of the waste is 42 psi per 100 feet (30.5 m) of vertical depth (assuming the waste brine has a density equal to that of water at 100°C). The hydrologic pressure requirement (that is, the pressure required to push the waste liquid into the injection aquifer) is considered the strongest determinant for pumping energy because of high variabilities in pressure differences.

Table 26 shows the expected energy cost for pumping at various values for the pressure requirement. These pressure values are representative of anticipated requirements, based on surveys of existing injection facilities. (See Tables 21 and 22 in the discussion of capital costs.) In some cases, the initial pressure requirement may be zero because of injection aquifer conditions and/or the pressure gain in the well tubing. However, a pump should be included in the design to allow for eventual increases in pressure requirements. Pressure requirements can change because of pressure built up from injection with time and because permeability of the stratum can change as solids are filtered from the injected waste.

Ocean Disposal

Methodology - The disposal of spent geothermal fluids to ocean waters may be an acceptable alternative in some cases since the most common constituent in geothermal brine is sodium chloride. However, if the geothermal waste significantly increases the salinity or toxicity in the area of the outfall, it will not be acceptable for direct disposal without appropriate prior treatment.

Ocean disposal of spent geothermal fluids would, in principle, be an uncomplicated operation. The process involves the conveyance of the liquid, probably by a pipeline, from the geothermal operation to the shore and thence through a pipe laid on or in the ocean bottom to some distance offshore. At

TABLE 26. OPERATING ENERGY COST FOR PUMPS

Pressure requirement (psi)	Cost (cents/1000l)
50	0.714
100	1.43
200	2.86
500	7.14
1000	14.3
1500	21.4
2000	28.6
4000	57.2

the outfall the wastewater may be released in a simple stream or jetted through a manifold or multiple port diffuser. The diffuser facilitates the mixing of wastewater with sea water, both vertical and laterally, thus causing rapid dilution and dispersion..

Because of the large volumes of geothermal waters that will generally be used per unit of energy extracted, pipelines would be large - perhaps one meter or larger in diameter.

Disposal Costs - The technical and economic advantages associated with ocean disposal of wastewaters have been diminished greatly in recent years as a result of new and more stringent pollution standards.

In addition to costly pretreatment requirements, the cost for conveyance and ocean disposal of geothermal plant wastewaters can be exorbitant. Approximate costs for conveyance and ocean disposal of wastewaters may be obtained from compilations of existing cost data such as that prepared for the San Francisco Bay and Sacramento-San Joaquin Delta Area Wastewater Management Survey Report (U.S. Army, 1972). Cost data from this study is presented in Figures 53, 54, 55, and 56.

Figure 53 provides annualized capital costs for conveyance lines at various wastewater flow rates. The curve is based on data for precast pipe installations in open country routing and for precast pipe-foundations laid on stable ground. Assumed land costs for conveyance line right-of-way in rural areas was $3,000/acre ($0.74/m^2). The design life of conveyance lines was assumed to be 50 years.
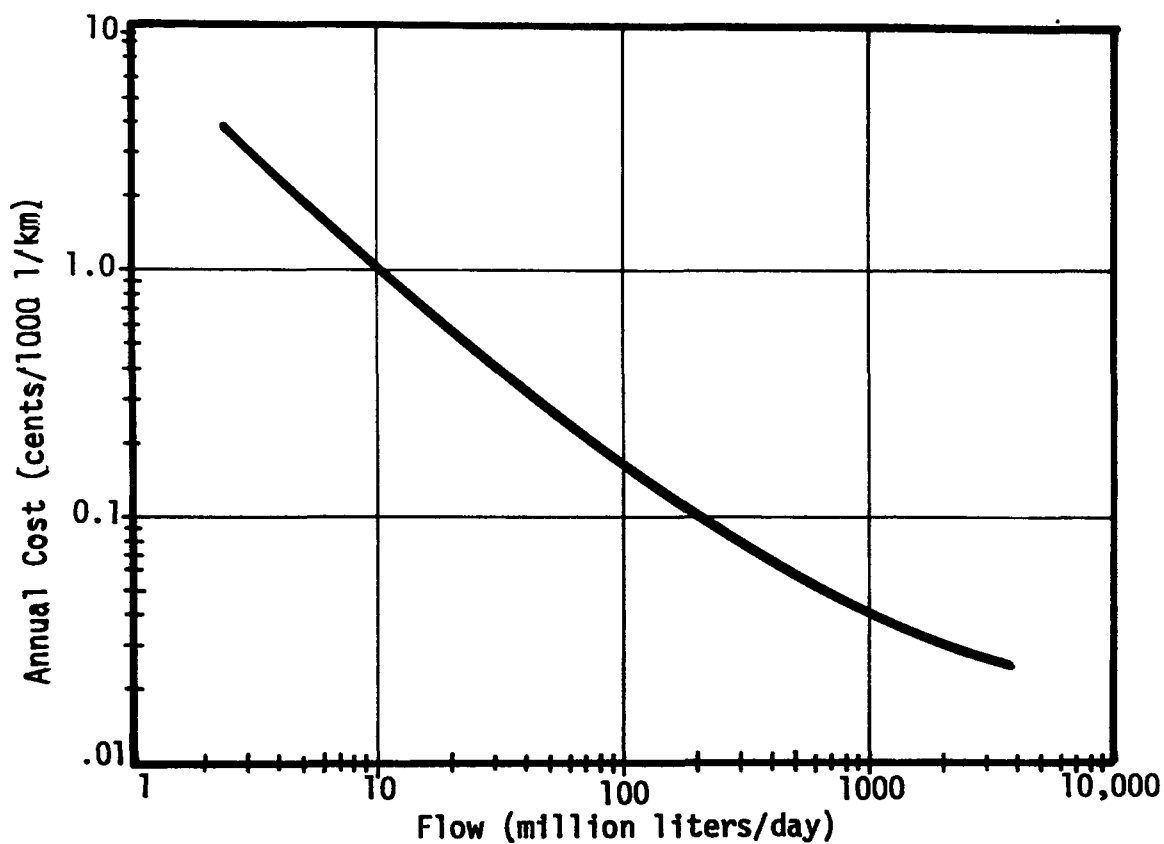
Figure 53. Annualized cost of installation of wastewater conveyance lines for open country routing
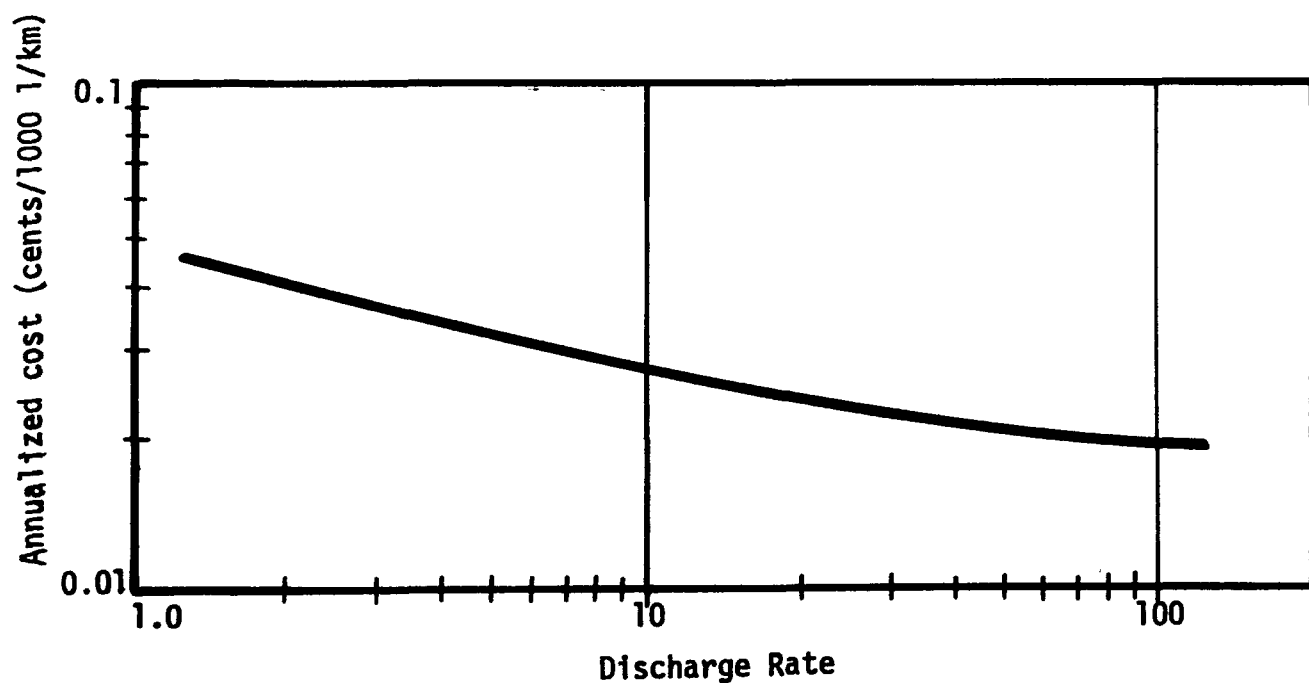
Figure 54. Annual cost of pumping wastewater for head of 40 ft (12 m) to 100 ft (30 m)

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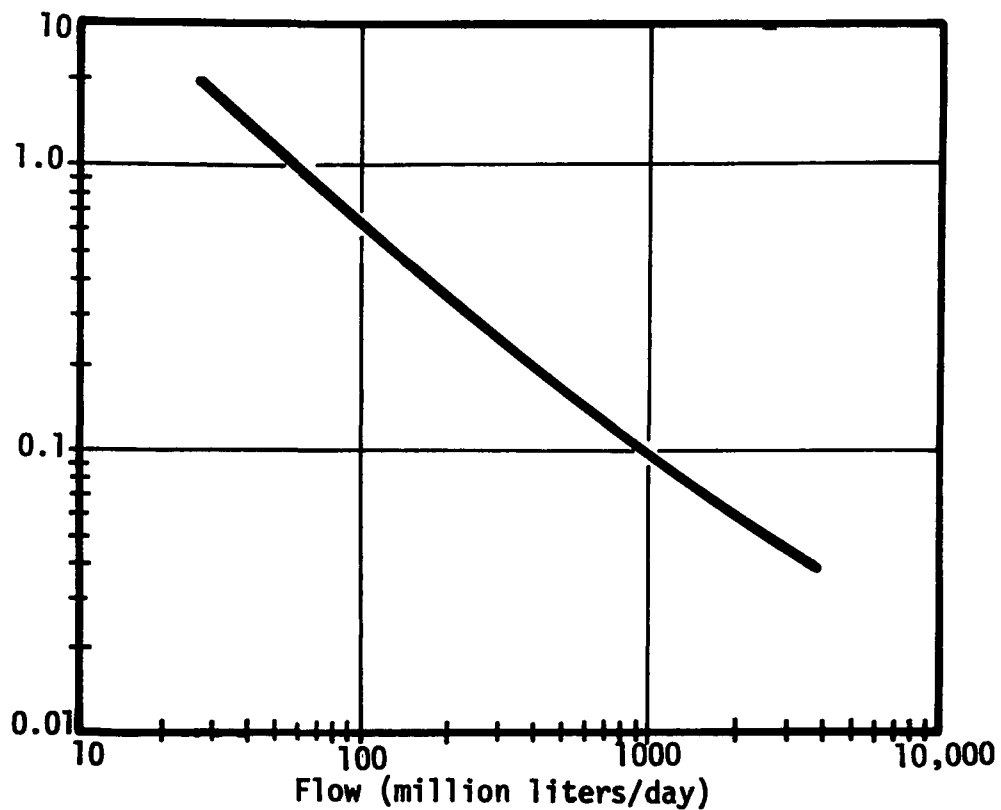
Figure 55. Cost of conveyance lines for ocean outfalls offshore
to depths of 200 feet (61 m)
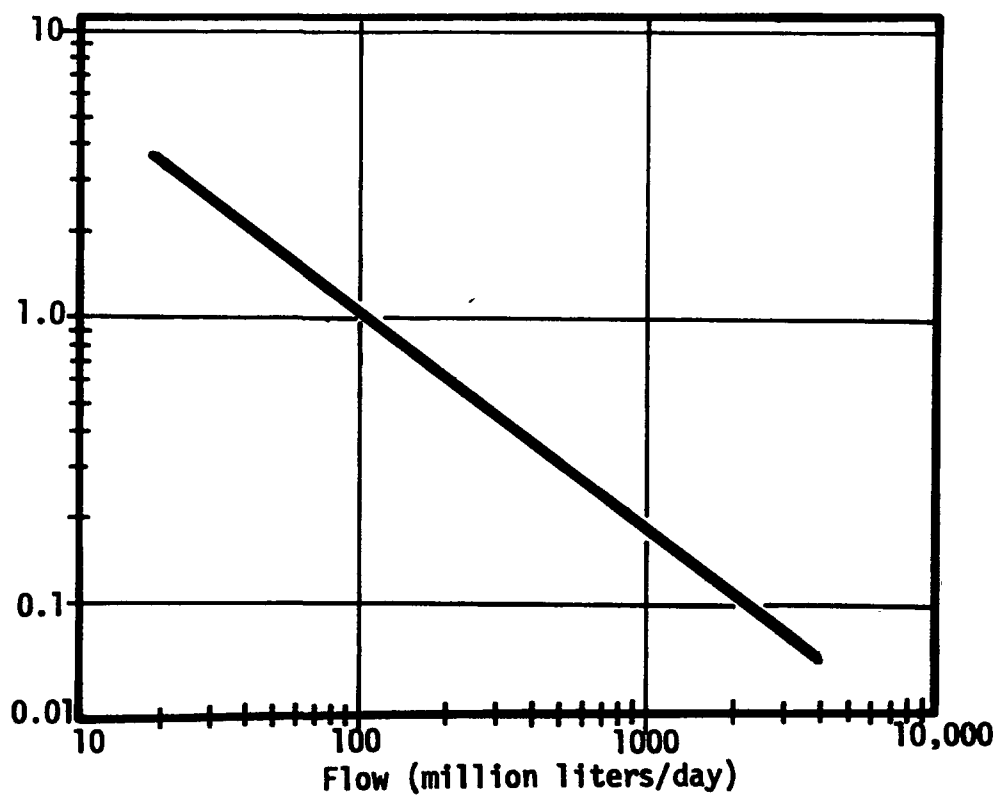
Figure 56. Cost of ocean outfall diffuser

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Figure 54 is a plot of annual cost of pumping wastewater when the elevation head of the pipeline is between 40 and 100 feet (12 m and 30 m). A head loss of 1.5 m per 1000 m was assumed for conveyance lines and pipes were sized accordingly. A peak flow factor varying between 2.9 at 1 MGD (0.044 m^3/s) and 1.5 at 300 MGD (13.2 m^3/s) was used for pumping station and power costs. Overall efficiency of pumps was assumed to be 72%. The design life of the pumping stations was assumed to be 30 years.

Figure 55 shows the cost of conveyance lines for ocean outfall construction offshore to depths of 200 feet (61 m). Figure 56 is a plot of the cost of ocean outfall diffusers.

Costs for Figures 53 to 56 were adjusted from the base data (Jan 1972) to current levels (1977) using the Marshall and Stevens Cost Index. All capital costs were amortized over a 50-year life-of-project period to compute annual costs. For facilities having a 30-year lifetime, replacement costs were included in the initial capital investment by calculating present worth of the replacement facilities.

Table 27 illustrates the high cost of disposing wastewaters from geothermal plants in ocean waters. The cost does not include wastewater pretreatment necessary to achieve effluent standards for ocean discharge.

TABLE 27. NORMALIZED COST OF OCEAN DISPOSAL OF GEOTHERMAL PLANT WASTEWATERS ($/1000 1/min)

Wastewater flow 1/min	Annualized cost of conveyance lines*	Annual cost of pumping**	Annualized outfall cost***	Annualized total cost
1,000	21,900	2,660	50,550	75,110
5,000	6,470	2,130	16,200	24,800
10,000	3,640	1,900	14,850	20,390
100,000	600	1,330	2,400	4,330.
350,000	110	1,030	890	2,070

*The cost of conveyance is based on an assumed open country routing of 200 miles (322 km). **It is assumed that wastewater is pumped through an elevation gain of 100 ft (30 m). ***An offshore outfall distance of 1 mile (1.6 km) is assumed. The outfall cost is the sum of the annualized costs for the outfall line and the diffuser.

Evaporation Ponds

Methodology - Where large land areas are available, evaporation ponds could provide a very simple approach to geothermal wastewater disposal. Evaporation ponds are more practical in arid regions where evaporation losses may reach 60 to 100 inches per year (150 to 250 cm/yr).

Construction of evaporation ponds involves excavation and/or diking, depending upon the topography of the area. In some cases, natural depressions may be utilized. In a few instances, it may be possible to enhance natural salt marshes as a wildlife habitat, principally by providing a constant water supply. It is not expected that evaporation ponds would normally have a surface drainage outlet.

Unless the soil is impermeable, evaporation ponds must be lined to prevent ground water pollution. Types of liners include clay, rubber, asphalt, concrete, and plastics (EPA, 1975).

Table 28 shows the expected water surface area required for evaporation ponds accepting wastewaters at median rates from the various geothermal energy conversion processes.

TABLE 28.   ESTIMATED WATER SURFACE AREA REQUIRED
FOR DISPOSAL OF GEOTHERMAL WASTEWATERS

Geothermal Conversion System	Median Wastewater Rate, 1/min	Water Surface Area, Acres* (km^2)
Direct steam power generation	17,000	1,436 (58)
Flashed steam, binary, total flow power generation	80,000	6,757 (273)
Direct heating open and closed systems	500	42 (1.7)
Desalination	3,000	254 (10.3)

*This is the amount of surface area required to maintain level of evaporation ponds at steady state. The required area is estimated by $A = Q/E$, where A = area required (acres), Q = wastewater generation rate (1/min), and E = evaporation rate (in/year). It is assumed that losses through the pond liner are negligible, and the evaporation rate is 60 inches per year.

Costs of Evaporation Ponds - Cost of evaporation ponds are related to various dependent factors in a recent study conducted for the Environmental Protection Agency (Black & Veatch, 1977). The data apply to average situations in the United States, and have been based on actual costs of projects over a wide geographic area, including varied construction conditions. The cost estimates developed in the study are representative of national average price levels as of January 1971.

The total capital investment cost includes the costs of construction, pond liner, embankment protection, engineering, land, and administrative requirements. The total operating and maintenance cost includes the costs of materials, supplies and labor. An estimate of total annual costs versus size of the evaporation pond is provided in Figure 57. Variations in these costs are to be expected with variations in the controlling factors.

Land Spreading

Methodology - Land spreading is a treatment method that relies primarily on biodegradation of the waste constituents. Inorganic wastes, such as those found in geothermal wastewaters, may not be suitable for land application. Significant concentrations of heavy metals would accumulate in the soil, posing threats to plant and animal life, and surface and ground water uses. The hazards of disposing of non-biodegradable materials on land are causing increasing concern, and regulations are becoming more restrictive.

Spraying on irrigable land, wooded areas, and hillsides has been practiced primarily for the disposal of industrial wastes such as cannery, pulp and paper, dairy and tannery. Treated effluents have often been used for golf course and park watering. The amount of wastewater that can be disposed of by spraying depends largely on the climatic conditions, the infiltration capacity of the soil, the types of crops or grasses grown, and the quality standards imposed where runoff is allowed.

In general, spraying systems may be classified as either low rate or high rate systems. Low rate systems utilize wastewater application rates of approximately 2 to 10 ft/yr, (0.6-3m/yr) whereas high rate systems achieve application rates of 150 to 350 ft/yr (45-107m/yr). Intermediate rates (10-150 ft/yr) are not widely used.

Low rate systems are segmented into two types of application systems, spray irrigation and overland runoff. Spray irriqation.is defined as. the controlled spraying of liquid onto the land at a rate measured in inches per week, with the flow path of the liquid being infiltration and percolation through the soil. Overland runoff is defined as the controlled discharqe (by spraying or other means) of liquid onto the land at a rate measured in inches per week, with the flow path of the liquid being downslope across the land.

High rate systems consist of rapid infiltration and subsequent percolation of wastewater into the soil. The process is defined as the controlled discharge of liquid onto the land at a rate measured in feet per week. Because of its high loading capacity, this process has a low potential for removing residual pollutants from the wastewater.
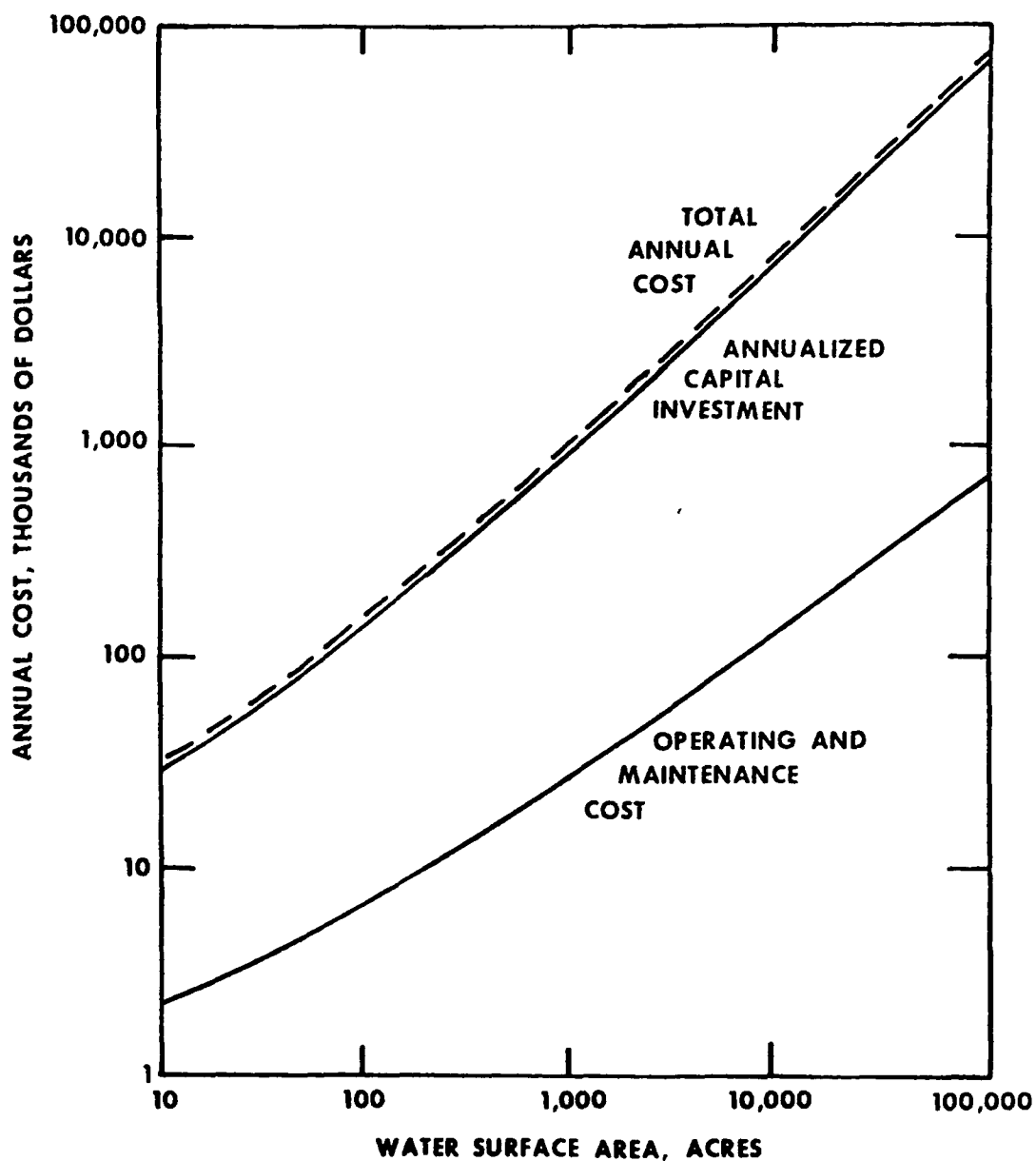
Figure 57. Total annual cost of evaporation ponds versus surface area.

The land area required for wastewater effluent disposal depends on the loading rate used. The loading rate in turn depends on many factors including:

- the soil capacity and permeability for infiltration and percolation;

- hydraulic conductivity (percolation capacity) of the root zone of cover vegetation;

- evapotranspiration capacity of site vegetation; and

- assimilation by soil and vegetation of nitrogen, phosphorus, suspended solids, BOD, heavy metals, and pathogenic organisms.

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The maximum hydraulic loadings of wastewater for various soil textures are shown in Table 29. The aqnual loading rate is substantially lower than might be indicated by the daily loading rate because of the number of rest periods required between applications.

TABLE 29. ESTIMATED MAXIMUM HYDRAULIC LOADING OF WASTEWATER EFFLUENT FOR VARIOUS SOIL TEXTURES (IDEAL CONDITIONS)

| | Movement Through the Soil Root Zone* | |
	cm/day	cm/yr
Fine sand	38.1	762
Sandy loam	19.0	457
Silt loam	8.9	229
Clay loam	3.8	102
Clay	1.3	25.4

*Precipitation plus effluent less evapotranspiration

The infiltration capacity of the soil limits the rate at which water can be applied to an area without runoff. Steeper slopes, previous erosion, and lack of dense vegetative cover also reduce the infiltration capacity and necessitate a corresponding reduction in application rates.

The hydraulic conductivity of the soil in a vertical direction determines the total precipitation and effluent application that can be transmitted to the ground water. Increased precipitation in a wet year reduces the amount of effluent that can be applied to various soil textures under ideal conditions.

Costs - The major advantage with wastewater land spreading is the low cost of the approach. Table 30 shows the total annualized cost of land spraying for the range of geothermal fluid flows anticipated. Capital investment costs and operating costs are based on an overland flow waste treatment system at Paris, Texas (Liptek, 1974). The system reported total construction costs at $1170 per acre ($3.11 per square meter) and operating costs at $.052 per 1000 gallons ($.014/1000 l) of wastewater. Application rate was a relatively high 0.6 inches (1.5 cm) per day. Compared to other land application methods (e.g., evaporation ponds), the cost of geothermal wastewater disposal is relatively low by land spreading. However, a major disadvantage to this approach is the vast amount of land required. For example, a typical size geothermal plant (e.g., 100,000 l/min) would require 3.7 square miles (9.47 km^2) of land designated for waste disposal.

TABLE 30. ANNUAL COST OF DISPOSAL OF GEOTHERMAL
WASTEWATERS BY LAND SPREADING

1/min	Land Surface Area Required,* Acres	Capital Investment,@ $1170/acre	Annualized Capital Cost 30 Years Life**	Operating Cost @$.052/ 1000 gal.*** ($.014/1000ℓ)	Total Annual Cost, $
1,000	23.4	27,400	2,440	7,200	9,640
10,000	234	274,000	24,400	72,000	96,400
100,000	2,340	2,740,000	244,000	720,000	964,000
350,000	8,200	9,600,000	855,000	2,520,000	3,375,000

*Based on application rate of .6 inches/day, (1.5 cm/day).
**At 8 percent interest rate.
***Assumes 8000 hrs of operation per year.

Containment of Unplanned Releases (Spills)

Methodology - Geothermal energy conversion systems will generally include the distribution of large volumes of geothermal fluids through a dispersed well and pipeline system. The possibility of system ruptures should be anticipated, and surface containment should be provided at points of high risk. Containment can include impermeable diking and/or excavation of areas large enough to contain the potential flow until the flow can be stopped.

A commonly used approach for containment involves the routing of spills to a nearby holding basin, similar in design to an evaporation pond. Factors which will affect the design of the holding basin include: the availability of nearby land, the permeability of the soil and the ability of the environment to accept the spill without adverse effect, the presence of other lagoons or ponds already serving the plant, site topography, and geology. Generally, a holding basin will require construction to depths of 10 to 15 feet by forming an embankment with earth moving equipment.

Costs - The cost for construction of holding basins may be estimated using cost data for aerated stabilization ponds similar in design (Black & Veatch, 1972). The costs shown in Figure 58 are derived from these data. Costs are shown for surface containment ponds suitable to manage unplanned releases of

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geothermal fluids at various flows and durations. The total annualized costs include construction cost, engineering design and embankment protection. The surface area requirements for the specified flow ranges (10 to 350,000 l/min.) vary from 12 square feet (1.1 m^2) to 41 acres (166,000 m^2) depending on the duration of the spill.
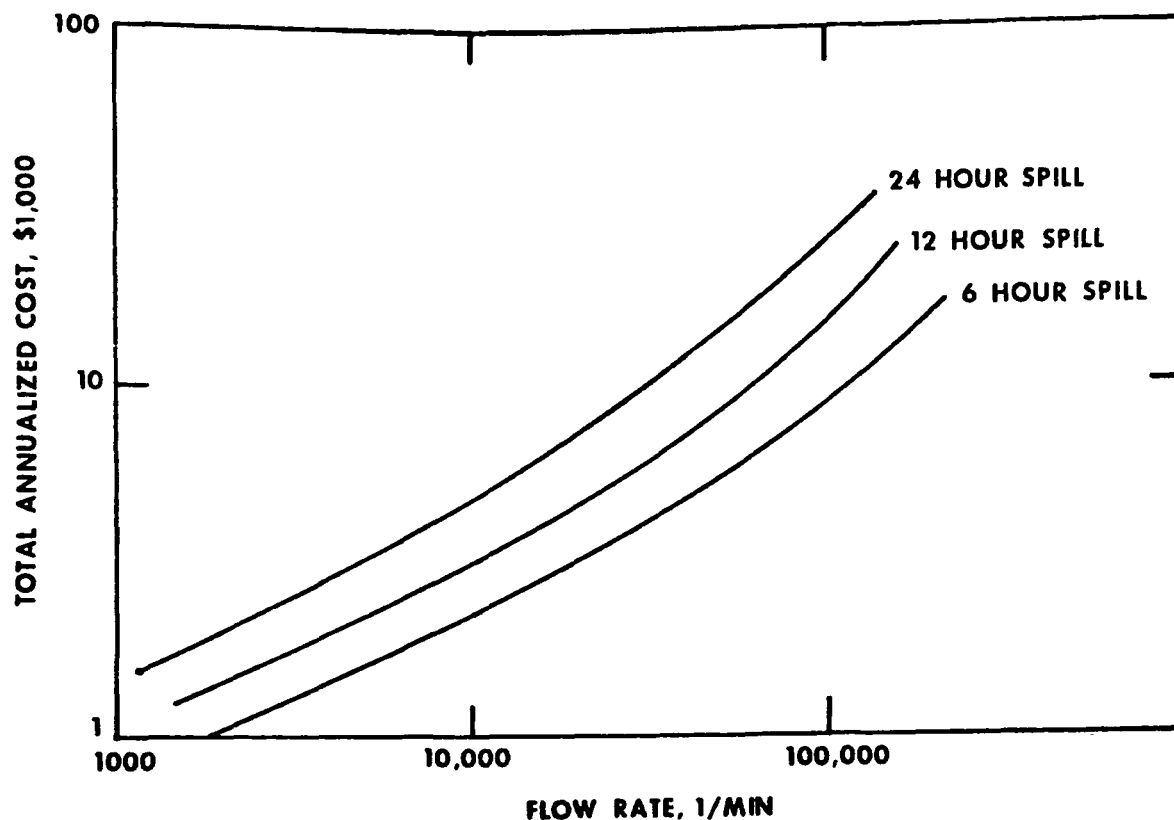

Figure 58. Annualized investment cost of spill containment ponds (10 foot depth).

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SECTION 7

SOLID WASTE GENERATION AND DISPOSAL COSTS

Design and costing of pollution control equipment require knowledge of
the kind and quantity of pollutants to be removed. In addition, any pollution
equipment generates waste sludges. The costs for waste sludge handling and
disposal vary with their quantity of generation. This section discusses the
pollutant loading from geothermal development and waste products generated as
a result of pollution control. In addition, the cost for disposal of residual
waste products is also presented.

POLLUTANT LOADING

As discussed in the section 5, the air pollutant currently of greatest con-
cern from geothermal development is hydrogen sulfide (H_2S). The generation rate
of H_2S is a function of the concentration of H_2S in the geothermal resources and
the steam flow rate. For steam flow in the range of 100,000 to 1,000,000 kg/hr,
the H_2S loading at various concentrations has been determined (Figure 59).

The pollutant loading from water discharges is a function of brine con-
centration and flow. For the purpose of this report four major conversion
processes have been identified for detail analyses. They are: direct steam
power generation system; flashed steam; binary; total flow power generation
system; direct heating, open and closed system; and desalination system.
Arbitrarily assigned concentration levels and flow ranges for analyses of
these geothermal conversion processes have been presented in Tables 17 and 18
in Section 6.

For simplicity, the major wastewater pollutants are grouped as total
solids (TS) and soluble metals (SM). Total solids include both dissolved and
suspended solids. Dissolved solids are mainly sodium, chlorides, sulfates and
carbonates. Soluble metals are quantitatively minor constituents and include
iron, manganese, boron, zinc, barium, lead, copper, arsenic, mercury, selenium,
chromium, silver and cadmium. Suspended solids are primarily silica and metal
silicates. They are often created from dissolved solids within the conversion
system upon reduction of brine temperature and pressure. The pollutant load-
ing from water discharges for these four major conversion processes are depict-
ed in Figures 60, 61, 62 and 63. The loading rates (in kg/hr) are derived
from information presented in Tables 17 and 18.

WASTE PRODUCTS GENERATED BY POLLUTION CONTROL EQUIPMENT

To estimate costs of handling and disposal of residual products, it is
necessary to identify and estimate the amount of the residual matter produced
by each control system process. The following sub-sections describe the resi-
dual generation sources and estimate the residual quantities. The discussion

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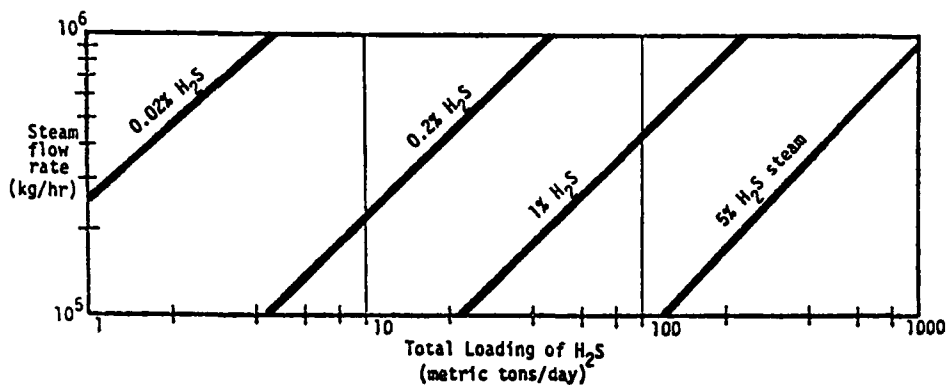
Figure 59. Hydrogen sulfide loading rate to turbine
as a function of steam flow rate
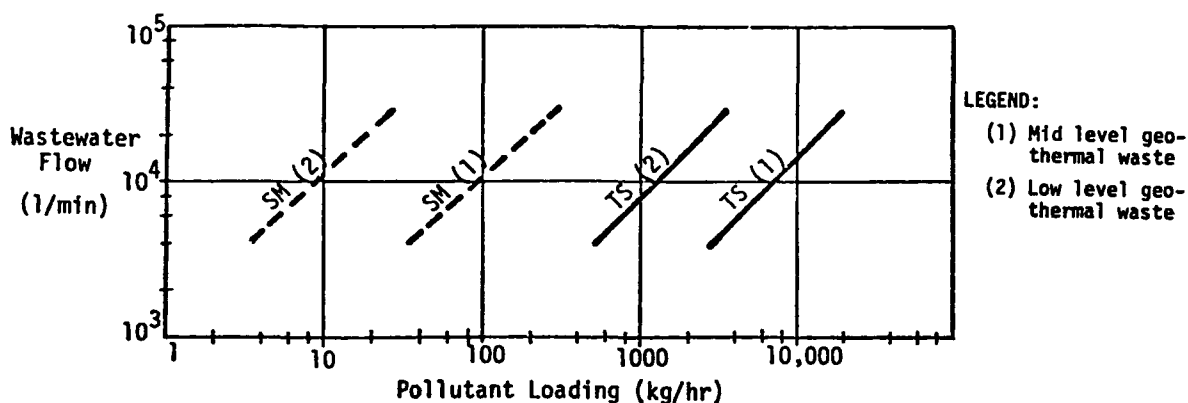
Figure 60. Wastewater pollutant loading for direct
steam power generation system
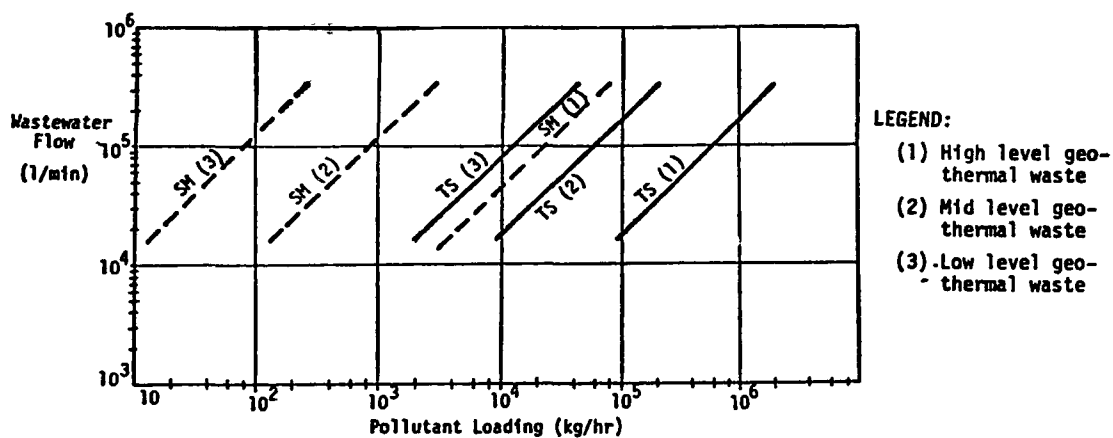
Figure 61. Wastewater pollutant loading for flashed steam,
binary, total flow power generation system

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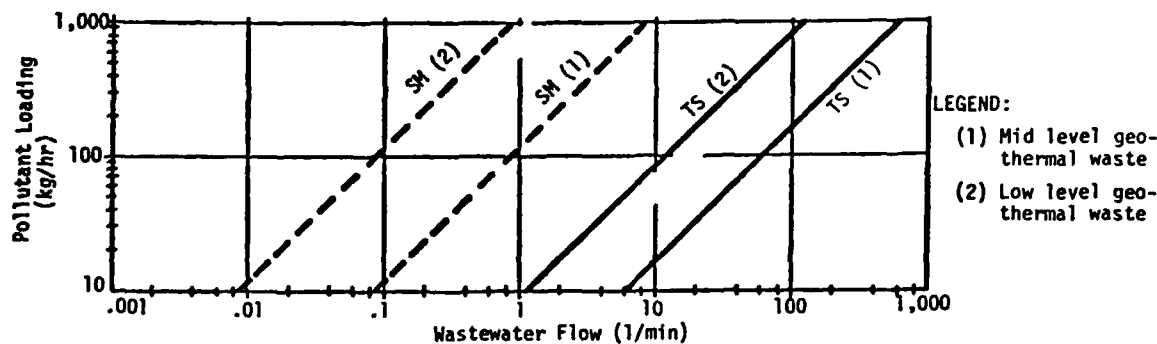
Figure 62.   Wastewater pollutant loading for direct
heating, open and closed systems
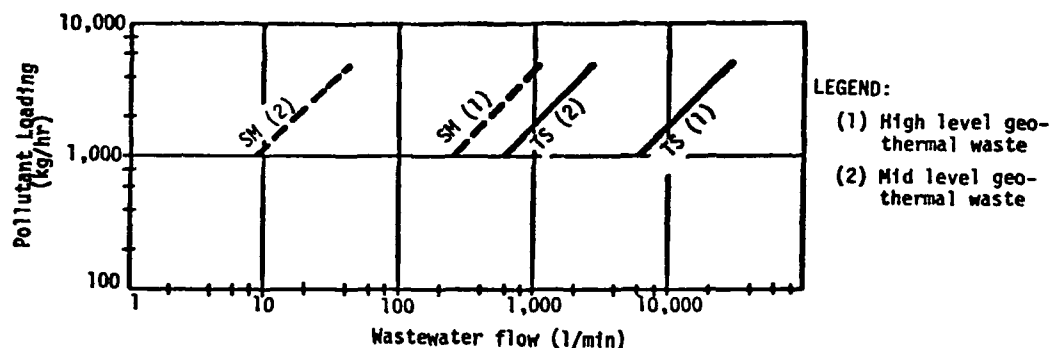
Figure 63.   Wastewater pollutant loading for
desalination system

is separated in two parts: 1) residuals from air pollution controls, and 2) residuals from water pollution controls.

Residuals from Air Pollution Controls

The pollutant of greatest concern is hydrogen sulfide. Candidate control systems for removal of hydrogen sulfide are the Stretford, iron catalyst, EIC, and Dow oxygenation processes. Applicability of these processes depends on the type of energy conversion system and the geothermal fluid properties.

For the various air pollution control systems applicable to steam turbine power generation, the H_2S concentration is assumed to range from .02 percent to 5 percent by weight of the steam input. Steam flow rates in the geothermal systems are assumed to vary from 100,000 to 1,000,000 kg/hr.

Stretford Process -

The Stretford process is self-maintained as the scrubbing chemicals are regenerated in multi-stage reactions and high purity elemental sulfur is the only residual product produced. Thus, the quantity of residuals produced by the Stretford process is estimated directly from the amount of H_2S known to be present in the ejector gases, and the assumption that the Stretford will remove essentially 100 percent of the H_2S from the treated gas stream.

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The quantity of H_2S in the ejector gases is approximately 80 to 90 percent of that existing in the turbine stream with the remaining 10 to 20 percent dissolved in the condensate. Figure 64 illustrates the amount of sulfur produced when the Stretford unit is used to control various concentrations of H_2S in the turbine discharge stream. The potential rate of sulfur production can be expressed as .847 QC kg/hr, where Q is the turbine steam rate (kg/hr) and C is the fraction by weight of H_2S in the steam. The constant .847 is the product of the fraction of H_2S in the ejector stream (90 percent) and the weight fraction of sulfur in H_2S (32/34).

Dow Oxygenation -

In the Dow Oxygenation process, H_2S is removed from geothermal liquid by injecting oxygen into the brine upstream of the power plant. The H_2S in solution is oxidized to sulfate, sulfide, and sulfur. The relative amount of different solids formed depends on the brine temperature and total dissolved solids in the brine. If the brine is already saturated with salts (e.g., calcium sulfate), free sulfur particles may be generated (J. Wilson, 1977). When the brine is not saturated with salts, the additional sulfate formed by oxygenation will remain in solution which may require treatment after the geothermal fluids are spent. The free sulfur fines will also remain suspended until removal is effected. In the most probable scenario, a substantial portion of sulfates, sulfate precipitates, sulfites, and sulfur created by the oxygenation process will remain in the geothermal fluid as it is flashed, condensed, and routed for final disposition as a wastewater. Ideally the residuals would be injected with the spent water and not require prior separation. However, for the purpose of the analysis here, it is assumed that the products of the oxygenation process will be removed from the wastewater as sludges so that the residual disposal problems attributable to this air pollution control system may be assessed.

Bench scale tests of the Dow process have shown that 90 percent to 100 percent of H_2S removal may be expected. Figure 65 shows the amount of solid material that will require disposal if the Dow Oxygenation process is employed for liquid dominated geothermal processes. It is assumed that calcium sulfate ($CaSO_4$) is representative of the precipitates in the waste solids which must be removed, and that the solids are concentrated to 25 percent by weight in settling ponds before disposal as a heavy sludge. The sludge thus produced consists of 1/3 free sulfur and 2/3 calcium sulfate.

EIC Process -

In the EIC process, H_2S is removed from steam upstream of the power plant. An aqueous solution of copper sulfate scrubs the H_2S from the steam as copper sulfide precipitate. A copper sulfide slurry is collected and concentrated by centrifuging. Copper sulfate is then regenerated from the concentrated solution. Sulfur dioxide and sulfur trioxide are produced during regeneration and scrubbed by an ammoniacal solution to produce ammonium sulfate, which is mixed with cooling tower water and injected underground. Hence, in the normal operation, all residual matter produced by the EIC process will be carried by the spent geothermal fluids to the injection wells.

Where injection may not be permitted, or where pre-treatment of the waste-water to facilitate injection is required, a portion of residual matter created by treatment will be attributable to the ammonium sulfate generated by the EIC control process. For the purpose of the analysis here, it is assumed that the total quantity of ammonium sulfate in solution will be removed as sludge and conveyed away for disposal. Reclaiming of the compound for fertilizer would probably be no more cost effective than outright disposal, since additional purification of the fertilizer would be necessary due to presence of other elements such as heavy metals (Brown, 1977). Figure 66 shows the amount of solid material, as ammonium sulfate, which could be removed as sludge and conveyed to landfill for disposal. The sulfate is assumed to be present in a sludge concentrated to 25 percent solids by weight. The efficiency of the EIC process in removing H_2S from the steam is assumed to be 90 percent.

Iron Catalyst -

In the iron catalyst process, ferric sulfate is added to cooling water to oxidize hydrogen sulfide contained in the aqueous phase. The cooling waters are then used additionally to scrub hydrogen sulfide from the condenser ejector gases. Elemental sulfur, water, and ferrous ions are formed, and the sulfur is removed from the cooling water by filtration. The filtration step produces a thick and toxic sludge that must be conveyed to a landfill for disposal.

Tests of the iron catalyst process at The Geysers Geothermal Field have shown approximately 50 percent removal of hydrogen sulfide to be technically achievable. This efficiency includes H_2S removal from the condensed cooling tower fluids and from the ejector gases. Figure 67 illustrates the quantities of waste sludge generated by the iron catalyst control process for various steam H_2S loadings. The sludge is assumed to be 25 percent solids (primarily sulfur) by weight.

Residuals from Treatment of Wastewaters

The residual materials removed by treatment of geothermal wastewaters consist of suspended solids and dissolved materials in the wastewater, as well as chemicals added during treatment. Together, these residual materials, along with water, constitute a sludge which is usually disposed of in a landfill or evaporation lagoon.

The amounts of sludge (Rs) produced by the various physical treatment processes may be approximated by:

$$Rs = \frac{QCE}{f}$$

where Q = wastewater production rate, l/min.

C = concentration of constituent in wastewater, mg/l

E = efficiency of system for removal of constituent

f = fraction by weight of constituent in residual sludge

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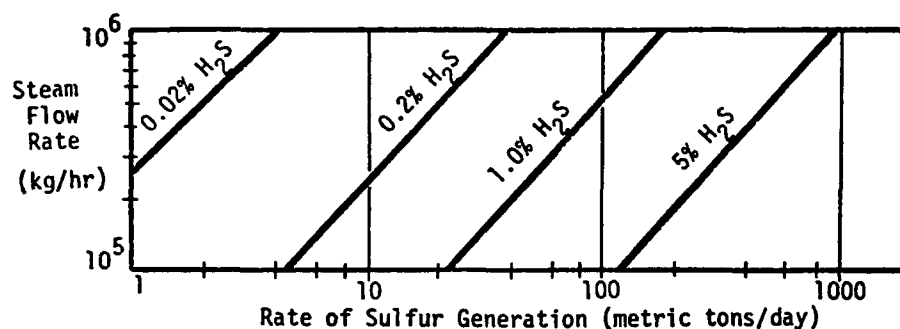Figure 64.   Sulfur generation from Stretford unit for various steam flow rates through turbine
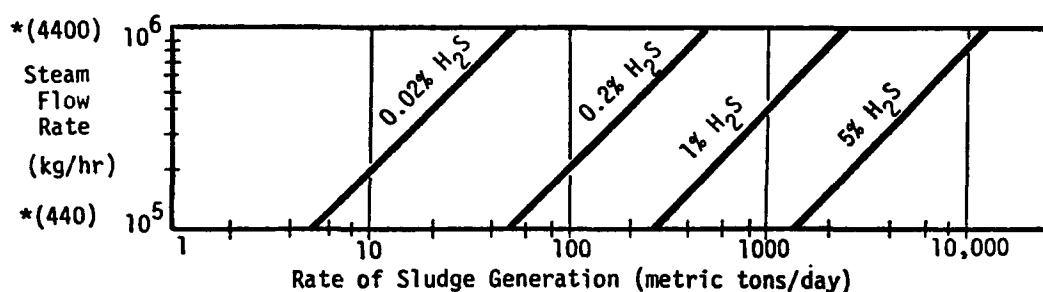

Figure 65.   Potential solid waste as sludge generated by Dow oxygenated system *(Brine flow rate is gpm)
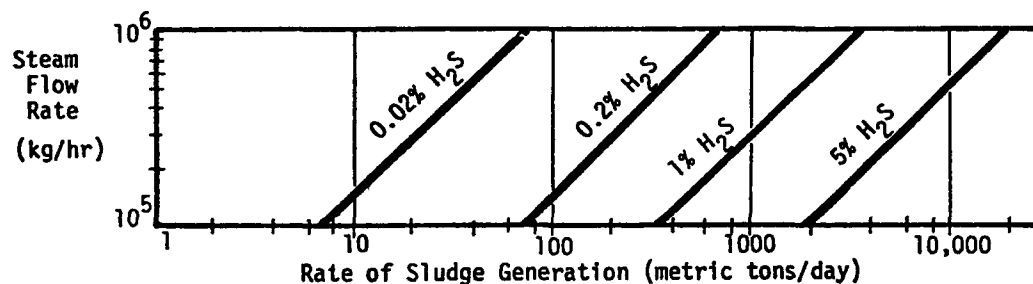

Figure 66.   Sludge generation from EIC process for various steam flow rates through turbine
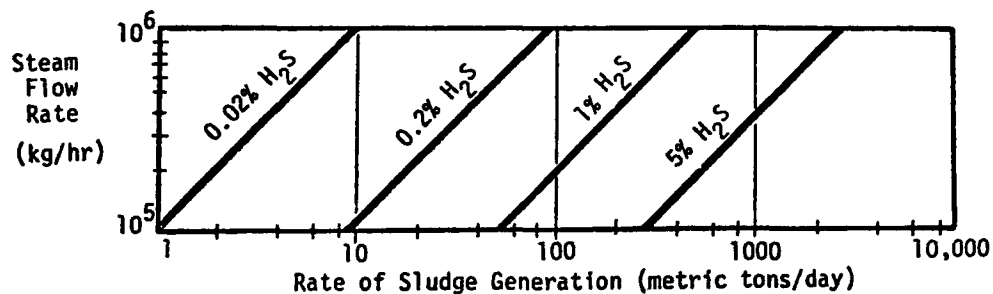
Figure 67.   Sludge generation from iron catalyst process for various steam flow rates through turbine

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The sludge produced by chemical precipitation of constituents is estimated by:

$$Rs = \frac{QCE}{f} \times \frac{mp}{mc}$$

where   mp = molecular weight of precipitate

mc = molecular weight of constituent in wastewater

The above equations were used to calculate the amounts of waste sludge produced by treatment of geothermal wastewaters from three specified levels of pollution to three acceptable discharge levels as delineated in Tables 17 and 18(Section 6). A total of nine scenario abatement schemes have been considered consisting of three levels of wastewater composition and three possible levels of discharge concentrations.

The wastewater treatment systems capable of attaining the various levels of effluent quality are comprised of combinations of individual treatment technologies which were discussed in Section 6. These schemes, and the cleanup scenarios identified above, were utilized as the basis for calculating the quantities of solid waste associated with the various geothermal conversion systems.

Table 31 summarizes estimates of the quantities of solid materials removed from the wastewater for each of the cleanup scenarios. The estimates are presented in terms of solids removal <u>factors</u> (i.e., quantity per unit of wastewater flow). Chemical precipitation was assumed to produce a solid weighing 1.5 times the material in the raw wastewater.

Table 32 summarizes estimates of the quantities of waste sludge generated by the treatments to remove solid materials from the wastewater. The estimates are presented in terms of sludge generation <u>factors</u>. The sludge is assumed to be 98 percent water by weight (a typical dilution rate for municipal sludges).

Based on the factors developed for solids removal and sludge generation rates, Table 33 presents a summary of the probable quantities of solid waste produced by treatment as a function of flow rates from the various conversion processes with varying raw waste and effluent qualities. It should be noted that the amounts of sludge generated are based on 2 percent solids concentration. Before these sludges are disposed of in landfill sites, they are usually concentrated in settling ponds or by evaporation to approximately 50 percent solids by weight. Thus the quantities shown in Table 33 can be reduced by a factor of twenty-five. For high strength waste, the amount of sludge generated could not be determined by the above equations because the raw waste constitutes a sludge itself (10% solids). For this strength waste, the sludge generation rate (SGR) was estimated by using the equation SGR = 1.583 Q where Q is the waste flow rate in lpm.

COST OF SLUDGE DISPOSAL

The amount of sludge requiring disposal depends on the geothermal process

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TABLE 31. SOLIDS REMOVAL RATE ACCOMPLISHED BY WASTEWATER CLEANUP SYSTEMS.

Concentration level of raw wastewater constituents, metric tons/day per 1/min of raw wastewater

Level of cleanup	High			Mid			Low		
	1	2	3	1	2	3	1	2	3
Total solids(TS) removal rate	.169	.169	.169	.0162	.0168	.0168	.00238	.00323	.00323
Soluble metals(SM) removal rate	.00161	.00161	.00161	.000281	.000292	.000292	.0000247	.0000281	.0000281
Total (TS + SM)	.171	.171	.171	.0165	.0171	.0171	.00241	.00326	.00326

TABLE 32. SLUDGE GENERATION RATES ASSOCIATED WITH CLEANUP OF GEOTHERMAL WASTEWATERS.

Concentration level of raw wastewater constituents, metric tons/day per 1/min of raw wastewater

Level of cleanup	High			Mid			Low		
	1	2	3	1	2	3	1	2	3
Sludge (Assuming 98% water by weight)	8.55	8.55	8.55	.825	.855	.855	.121	.163	.163

TABLE 33. SLUDGE GENERATION RATE (METRIC TONS/DAY) AS A FUNCTION OF
FLOW RATES FOR SPECIFIED CONVERSION PROCESSES.

Raw waste and treatment level	Direct steam power generation flow (lpm) 4000 to 30,000			Flashed steam, binary, total flow power gen. flow (lpm) 15,000 to 350,000			Direct heating open and closed systems flow (lpm) 10 to 1000			Desalination process flow (lpm) 1000 to 5000		
High level*												
1		--		23,750	to	554,000		--		1580	to	7910
2		--		23,750	to	554,000		--		1580	to	7910
3		--		23,750	to	554,000		--		1580	to	7910
Mid level												
1	3300	to	24,750	12,375	to	288,750	8.25	to	825	825	to	4125
2	3420	to	25,650	12,825	to	299,250	8.55	to	855	855	to	4275
3	3420	to	25,650	12,825	to	299,250	8.55	to	855	855	to	4275
Low level												
1	484	to	3630	1815	to	42,350	1.21	to	121		--	
2	652	to	4890	2445	to	57,050	1.63	to	163		--	
3	652	to	4890	2445	to	57,050	1.63	to	163		--	

* These estimates do not reflect wastewater treatment because the raw waste constitutes a sludge itself. Therefore the values reported are the binary wastewater generation rates (corrected to a specific gravity of 1.1); for high level waste, $S.G.R. = 1.583Q$.

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and the size of the plant. Based on the wastewater flow ranges assumed for this study, waste sludge will be generated in quantities from one metric ton/day to 554,000 metric tons/day. The sludge may be disposed of either directly by landfill or by a two-step process involving settling lagoons and landfill. If the sludge contains substantial amounts of heavy metals, it will be considered a "hazardous waste" and therefore subject to special disposal regulations. For example, land spreading of such wastes would not be acceptable. Lagoons and landfills must meet specified design standards before hazardous waste may be accepted; these requirements will lead to higher cost of disposal.

Landfill

The cost for disposal of hazardous waste in an appropriate landfill varies from $8 to $12 per ton of waste. Rate reductions of up to 25 percent may be available for the disposal of high volume and/or repetitive wastes (Kinna, 1977). Normally, the greatest cost of waste disposal at a landfill is hauling. Typical hauling rate is $32 per hour for a truck having a capacity of 20-25 tons. Hauling time consists of about 2 hours for loading and unloading, plus actual road-trip travel time.

Table 34 shows the total cost for waste disposal of geothermal wastewater treatment sludge when the disposal site is 200 miles (322 km) from the plant. The sludge is assumed to be 50 percent by weight solids, which implies prior dewatering in an evaporation lagoon before disposal to the landfill. The disposal costs of Table 34 do not reflect the expense of sludge dewatering required to prepare the waste sludge for economical landfill disposal. Dewatering of the sludge has been assumed to be carried out to a relatively high degree (50 percent solids), such that waste conveyance and disposal costs are minimized. The actual degree of dewatering would depend on the economic trade-off between wet transportation and drying of the sludge before transport.

Evaporation Ponds

Normally, landfills are used as the ultimate disposal locations, and evaporation ponds are used as an interim process for drying or concentrating sludges before removal to landfill. Evaporation ponds may also be used for ultimate disposal; however, potential hazards of heavy metal accumulation and subsequent leakage into the soil and groundwater may rule out permanent disposal of geothermal wastewater sludges in evaporation ponds.

The factors affecting cost of an evaporation pond include: proximity to wastewater treatment site, lining requirements, local meteorology, construction costs, cost of land, and administrative costs. Conveyance distance costs for the sludge to the drying ponds is usually minimized by locating the ponds near the wastewater treatment site. Unless the ponds are in impermeable soils, it will be necessary to install a liner material (e.g., plastic sheet) in the pond to prevent movement of leachate into water sources. The depth of the pond is usually about 3 to 5 feet (1-1.5 m) and total surface area and land requirements are determined by the rate of sludge generation and the precipitation and evaporation occurring at the site. Table 35 summarizes estimates for the land requirements and cost for evaporation ponds which will accept

TABLE 34. COST OF LANDFILL FOR GEOTHERMAL WASTEWATER TREATMENT SLUDGES

Geothermal converison system	Median Sludge generation rate (concentrated)*	Cost of hauling**	Cost of disposal**	Total cost of disposal	
	Metric ton/day	$/day	$/day	$/day	$/year
Direct steam power generation	581	6500	5810	12,310	4,493,200
Flashed steam, binary, total flow power generation	6240	70,000	62,400	132,400	48,326,000
Direct heating open and closed systems	17.2	193	172	365	133,200
Desalination	103	1154	1030	2184	797,200

*Sludge is assumed to be concentrated to 50 percent by weight water after dredging from evaporation ponds. The median rate is based on the median level of the expected sludge generation range for the conversion system specified. The quality of the raw wastewater is assumed to be at the mid level anticipated for given conversion system.

**Cost of disposal is taken as $10/metric ton and cost of hauling is estimated at $224 per truckload of 20 metric tons. Hence, the total daily cost of removal and disposal is $21.4/ton of concentrated sludge. Expressed in terms of the diluted sludge (prior to dewatering), the annual cost of disposal and hauling is $312/ton.

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TABLE 35. TOTAL ANNUAL COST OF WASTE SLUDGE TREATMENT, REMOVAL
AND DISPOSAL, AT "MEDIAN" SLUDGE GENERATION RATES.*

Energy conversion system	Cost of evaporation ponds	Hauling cost to landfill	Landfill disposal	Total annual cost
	COST IN TOTAL DOLLARS			
Direct steam power generation	1,000,000	2,372,500	2,120,700	5,493,000
Flashed steam, binary, total flow power generation	13,500,000	25,550,000	22,776,000	61,826,000
Direct heating, open and closed systems	84,000	70,450	62,780	217,230
Desalination	800,000	421,210	376,000	1,597,000

*(Based on information developed in Table 34).

"median" quantities of waste sludge generated by the various geothermal conversion systems. The water surface area requirements are estimated based on the amount of evaporation expected at the geothermal site. A typical net evaporation rate of 60 inches/year (1.5 m/year) in the Southwest United States (Wiessman, 1972) is sufficient to maintain a constant water level in a one acre evaporation pond accepting 16 metric tons/day of liquid sludge (assuming negligible water loss through the liner of the pond). The total land area required to accommodate the evaporation ponds is based on results of a recent study for EPA (Black & Veatch, 1977) which established a correlation of water surface area to land area requirement. The required land includes provision for access roads, dikes, and support equipment. The total cost of the evaporation ponds is dependent on the various factors discussed above, and is determined from Figure 57 in Section 6.

Table 35 shows that substantial amounts of land area will be needed if geothermal waste sludges are concentrated in evaporation ponds before removal to a landfill site. The land requirements would be greatest for flashed steam power generation systems. It is estimated that a median sludge generation rate from a flashed steam system would require nearly 10,640 acres (43 km^2) of land for evaporation ponds. The corresponding costs of the evaporation ponds, plus the cost of conveyance and disposal at landfills is prohibitive, as shown in Table 36.

TABLE 36. LAND REQUIREMENT AND COST OF SLUDGE EVAPORATION PONDS

Geothermal conversion systems	Median sludge generation rate, metric ton/day	Water surface area required,[a] acres*	Land area required acres	Total annualized cost cost of evaporation ponds**
Direct steam power generation	14,535	901	991	$ 41,000,000
Flashed steam, binary, total flow power generation	156,000	9672	10,640	$413,500,000
Direct heating, open and closed systems	431	27	46	$ 84,000
Desalination	2565	159	175	$ 800,000

*Determined by $A = Q/\gamma E$, where A = area required, Q = sludge generation rate, γ = density of sludge, and E = evaporation rate. It is assumed that γ = 60 lb/ft³ and E = 60 inches per year. Thus, $A = .062\ Q$, where A = acres of water surface area required, and Q is the wastewater flow in metric ton/day. (Based on information developed in Table 34).

**Based on evaporation pond cost data presented in Section 6 (Figure 58).

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16. ABSTRACT

This report provides preliminary cost estimates of air and water pollution control technologies for geothermal energy conversion facilities. Costs for solid waste disposal are also estimated. The technologies examined include those for control of hydrogen sulfide emissions and for control of water discharges containing metals and inorganic dissolved solids.

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