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Electrophoretic Clarification of Water

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ABSTRACT

There is an ever growing need for new and superior water treatment methods which will remove the alarming growth and variety of pollutants present in our waters. Suspended particulate matter such as clay, algae, and bacteria are troublesome pollutants in almost all waters and particularly in surface water.

This research project was primarily an investigation into the direct use of electric currents and electric fields to bring about water clarification and purification. Electrophoretic and electrochemical systems appear well adapted for removal of these electrically charged pollutants and in addition have the potential of being entirely automated. Numerous design concepts are advanced and tested through the use of laboratory models. These include a parallel plate model which was designed entirely for electrophoretic removal, and porous filter and electrode grid models which incorporated both electrophoretic and electrochemical capabilities.

Successful water clarification was attained with the parallel plate model only when the influent water was of very low electrical conductivity. Electrolysis products at high conductivities caused sufficient turbulence to completely disrupt electrophoretic transport. In addition, primary and secondary chemical reactions took place due to electrolysis, which altered the characteristics of

the suspension. It was also concluded that the ease of automation with an electrophoretic system does not justify the high cost of water treatment by this method.

Water purification by electrochemical means overcame this problem and was successful both operationally and economically. As a result of the experimental testing, an example design of a small semiautomated electrochemical water system is offered which incorporates electrochemical flocculation, settling and electrochemical disinfection operations.

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CHAPTER I

INTRODUCTION

The preservation and restoration of adequate, well distributed quantities of pure water, free from industrial, municipal, and natural contaminants, is an absolute necessity for the survival of mankind on this planet. The population explosion along with man's tendency to pollute his environment will make this much more difficult to achieve in future years. There is, therefore, a critical need to investigate new and potentially superior water treatment methods and systems which will remove the alarming growth and variety of industrial and municipal pollutants as well as natural contaminants.

One important aspect of water purification and clarification is turbidity removal or removal of suspended particulate matter. Clay, algae, and bacteria are the primary contributors to turbidity in natural waters particularly if the source is surface water. Turbidity removal is especially troublesome when encountered in small, domestic water supplies.

The Potential Use of Electrophoretic and Electrochemical
Methods in Water Purification

Electrophoresis along with other electrokinetic and electrochemical phenomena appears to be a very promising method of water

The literature citations in this manuscript follow the style of the Transactions of the American Society of Agricultural Engineers.

purification. Very simply, electrophoresis is defined as the migration of charged suspended particles in a d.c. electric field. This process of water purification is made possible by the fact that most naturally occurring substances carry a negative charge.

Clays derive their negative charge primarily by substitution of ions of lower valence within the crystal lattice. This involves primarily the substitution of Mg^{++} and Fe^{++} for Al^{3+} in the octahedral layer of the clay crystal and Al^{3+} for Si^{4+} in the tetrahedral layer. Bacteria and algae are normally negative because of dissociable polar groups on the cell surfaces which ionize in water. These groups are primarily carboxyl and amino groups and the degree of ionization depends largely on the pH of the system.

Only in the last decade has electrophoresis been experimentally applied to the purification of water, primarily for the removal of suspended particulate matter. It has also been found to decrease other colloidal and solute impurities (4, 9, 11, 15, 16).*

Bier (4), in a laboratory study of forced-flow electrophoresis, completely eliminated suspended particulate matter of the types previously mentioned and radically decreased colloidal impurities such as iron, humic acid, proteins, and detergents. In addition he reported substantial desalting of the purified effluent.

*Numbers in parentheses refer to reference numbers.

Significant color removal may also be achieved by electrophoresis. Most authors believe color in water is colloidal in nature and thus should possess a negative charge (7). Black and Christianson (6) have shown that the materials responsible for color in water exist primarily in colloidal suspension in water. They suggest that the color in water is due to light scattering and fluorescence rather than molecular absorption of light energy.

Electrophoresis in combination with electrochemical methods also offers considerable potential in specialized water treatment problems where other methods are not applicable or effective. Such a situation was reported by Boyd (9) in the cleanup of heavy water at the Hanford Plutonium Recycle Test Reactor. Another indication of the tremendous potential of electrophoresis for water purification is the fact that as of 1960, the purest laboratory processed water reported to that date was obtained utilizing electrophoresis. Haller and Duecker (14) using a modified electrophoretic procedure for ion removal obtained an electrical conductivity of 0.039×10^{-6} mhos cm^{-1} at 18° Centigrade. This was three times lower than the lowest previously reported value which was obtained by 36 consecutive vacuum distillations. This ultra low-conductivity water was prepared by recirculating water through an electric field of 1,000 volts cm^{-1} maintained between two ion selective membranes.

Tertiary wastewater treatment is another possible use of this method. Faust and Manger (12) determined the electrophoretic mobility of particulate matter in domestic wastewater from 707 observations.

They reported values ranging from 0.55 to 3.75 microns sec^{-1} volts $^{-1}$ cm with an average of 1.73. This mobility is on the same order of magnitude as most clays.

There exists a very definite potential for application of electrophoretic and electrochemical methods to the purification of waters. The design, development, and testing of economically feasible systems yet remain.

Purpose of Research

The purpose of this reported research was to investigate the potential development of a satisfactory water clarification system, both from an operational and an economical standpoint, which utilized electrophoretic and electrochemical methods.

CHAPTER II

REVIEW OF THE LITERATURE

This research was primarily of an applied nature in that the primary interest was the application of electrokinetic theory rather than its investigation. Reviews of the classical theories and investigations along with present interpretation of the diffuse electrical double layer and its relationship to electrokinetic movement of colloids are given in other references (1, 2, 18, 20, 21, 23, 25, 27).

Mueller (20) has given an excellent review of the contributions of various researchers in the field of electrokinetic phenomena up to 1939. These include the contributions of Helmholtz, Gouy, Smoluchowski, and Millikan, Chapman, Stern, Debye, Huckel, Mooney, Henry, Herman, and others.

Recent important research includes the work of Wiersema (27). He gives an excellent account of the forces acting on a spherical colloidal particle when it is in uniform electrophoretic motion. These include the force exerted by the electric field and the force due to frictional resistance or Stokes' friction. In addition, there are two other retarding forces present caused by the ionic atmosphere surrounding the particle. One is the result of the force of the electric field on the counter ions in the diffuse double layer which is transferred to the solvent. The other is due to the fact that the center of gravity of the ionic atmosphere lags behind the center of the

particle and is referred to by some investigators as the relaxation effect. Wiersema also lists many pertinent additional references.

Excellent reviews of recent research as well as historical contributions and theories are also given by Sennett and Olivier (25). In addition, they give a thorough discussion of the electrical double layer and zeta potential, the measurement of electrokinetic effects, and various applications of electrokinetic phenomena in practical systems. The paper lists 112 references and provides an important review of literature concerning electrokinetic phenomena. A less technical account of research on electrokinetic phenomena up through 1956 has been reported by Oaks (21).

Recent research applied specifically to water purification which concerns or utilizes the zeta potential or negative charge of colloidal or suspended particulate matter is basically of two types. One concerns the coagulation and flocculation of the suspended material by neutralizing the repelling negatively charged particles or by their attraction to positive coagulants. The other process is, of course, electrophoresis or the transport of the suspended material due to the presence of a d.c. electric field.

Important contributions have been made by Black (5, 7, 8) and Riddick (23, 24) in the area of coagulation and flocculation as it relates to zeta potential and its control. Many other investigators have also made substantial contributions in this area. To date this writer is aware of only three principal investigators who have considered electrophoresis primarily as a water clarification method.

Hiler (15, 16) has offered the most theoretical approach to the problem of water purification by electrophoresis. Theoretical equations were developed predicting the movement of colloids between a rod and concentric cylinder and between two parallel flat plates. A single particle analysis approach was used for both configurations. In addition, the parallel plate arrangement was subjected to a more rigorous investigation which considered the entire colloidal system utilizing the theory of stochastic processes. In the latter case very good agreement was obtained between theoretical and experimentally obtained values of colloidal concentrations at strategic points within the system.

Two observations were made from the practical standpoint of colloidal removal. One observation was that very low concentrations were quickly obtained in the region near the negative electrode with the parallel plate configuration. Thus water near the negative electrode may be removed for use, while the remainder is recycled or wasted. The other observation was that for most efficient colloidal removal, it was desirable to have some turbulence present in the initial stages of flow through the electric field. According to Hiler, initially the turbulent fluctuations move more particles to the absorbing positive electrodes.

Boyd (9) reported using electrophoresis successfully to clean up heavy water at the Hanford Plutonium Recycle Test Reactor. Numerous other methods that were tried had failed. Using electrophoresis coupled with standard filtration and ion exchange processes, complete

removal of the following contaminants was achieved:

- (a) Oil in colloidal suspension;
- (b) Oil in emulsion;
- (c) Finely divided particles of dust and rust in suspension; and
- (d) Metallic ions.

Twelve 46-inch parallel aluminum plates spaced 0.25 inches apart and providing 6,070 square inches of surface area comprised the unit. This clarification system utilized electrophoresis to bring about deposition of the charged particles on the anode and cathode surfaces. In addition, some flocculation and precipitation occurred due to the aluminum hydroxide which was released from the aluminum plates as they were slowly consumed. This was due to electrolysis or electrochemical phenomena. Boyd did not give cost figures since necessity rather than economics was the major consideration in this case.

Bier (3, 4, 11, 19) has conducted the most intensive laboratory investigation to date in applying electrophoresis as a treatment for domestic water. The Bier unit is a continuous flow apparatus which consists of several electrophoretic cells connected in parallel. The cells are separated by semipermeable membranes which allow free diffusion of low molecular weight ions, but do not allow free flow of water. Thus electrical conductivity is maintained while limiting water and colloidal material to each individual cell. Each cell is further divided into an inflow cell and an outflow cell by a filter which is held in position by thin spacers. The filters have a much

larger porosity than the membranes and allow free flow of water. Millipore and Gelman controlled porosity filters are examples of the types of filters used. The whole assembly is exposed to the electrical field which is applied by means of electrodes located at the two ends of a cell pack. A separate flow system circulates between the cell pack and the electrodes to carry off electrolysis products.

A schematic view of the Bier forced-flow electrophoresis cell is shown in Figure 1. The direction of the gross flow of liquid is indicated by solid arrows (V_f) and that of electrophoretic transport by broken arrows (V_m). Negatively charged colloidal particles in the water entering the first compartment at (I) will migrate toward the positive electrode due to electrophoretic transport. If this velocity of particle migration is maintained equal to or greater than the velocity of hydraulic flow in the opposite direction, the colloids will not pass through the filter. Turbid effluent is drawn out of the inlet compartment at (W).

As previously stated, this unit was successful in removing practically all suspended pollutants from water as well as reducing the soluble ion concentration. Bier's estimated costs for removing silt and clay from water with a specific conductivity of 250 micro-mhos cm^{-1} , assuming two cents per kilowatt hour, were \$1.24 per 1,000 gallons for a laboratory model and \$0.41 for a scale-up plant.

The application of electrophoretic and electrochemical methods to water purification has not been extensively studied. Hiler's work

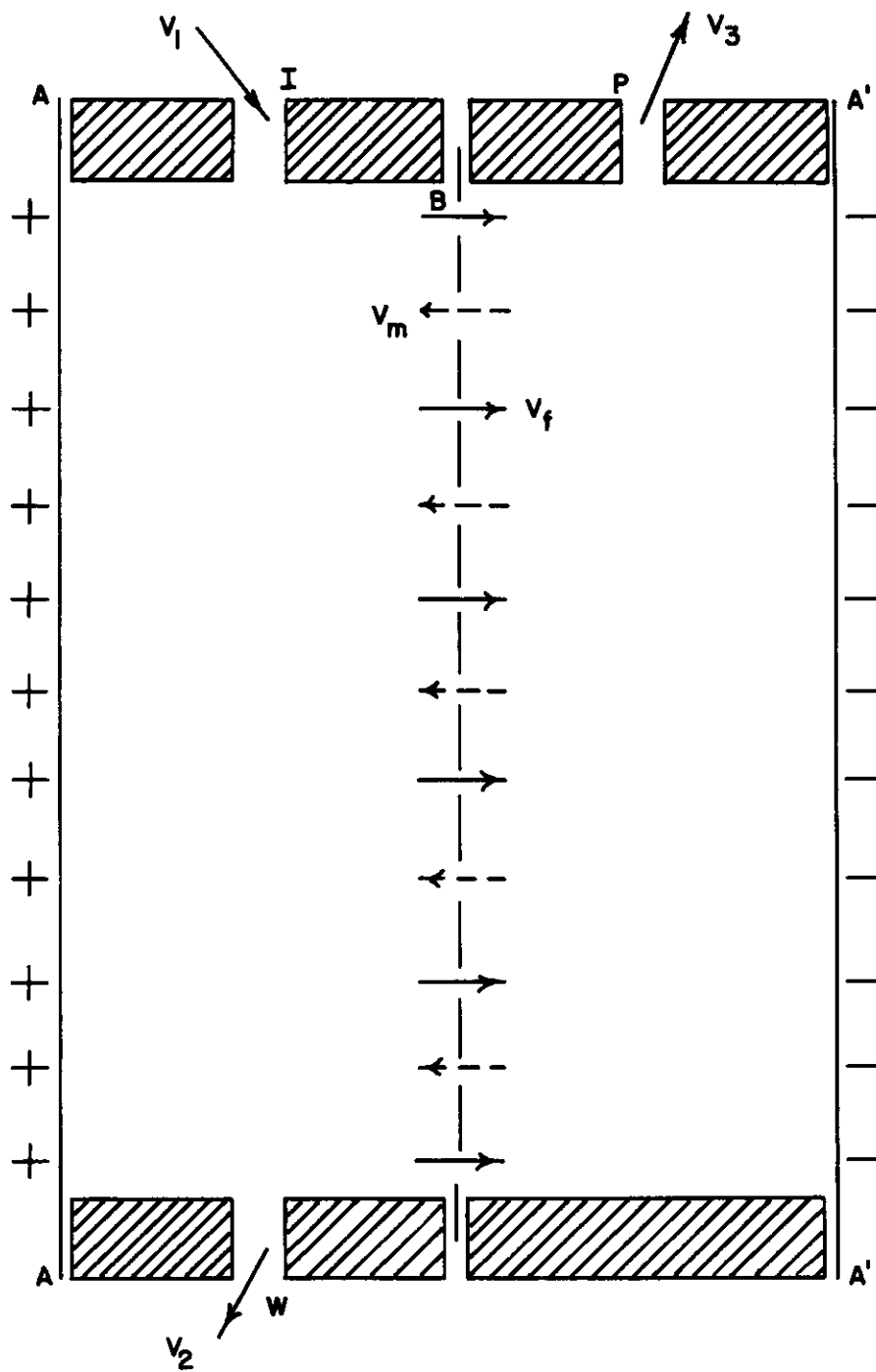


Figure 1. Schematic presentation of Bier forced-flow electrophoresis cell.

was primarily theoretical and its application needs to be investigated. In addition his research was confined to waters of very low electrical conductivities. Boyd indicated a specialized application of electrophoretic and electrochemical phenomena but his results did not separate these results from the additional effects of the filtration and ion exchange operations. He also failed to give an economic analysis of the process. The cost of the operation Bier describes is not competitive with present water treatment methods.

There is a need at the present time for a small domestic water treatment plant which can be automated. This research will be focussed primarily at the possibility of filling this need. These methods, however, may also be applicable to municipal water treatment and many other specialized cases.

CHAPTER III

SYSTEM DESIGN PRINCIPLES

Three basic design theories or principles are advanced here and were incorporated into several electrophoretic and electrochemical clarification system models for evaluation. These design principles concerning essentially different electrode arrangements and flow regimes are obtained primarily from prior theoretical relationships and parameters which have been shown previously to influence electrophoretic or electrochemical removal of pollutants from water. Each design criterion was aimed at obtaining both good operational utility and acceptable economic efficiency.

Parallel Plate Design

Hiler (16), in evaluating electrophoretic clay removal between two parallel plates, was concerned primarily with the deposition of the colloids on the positive plate. This is likewise the concept inherent in Boyd's design (9). The design advanced herein differs somewhat from those which advocate sediment removal almost entirely by deposition on the anode. Instead this design was intended to obtain a downward movement of suspended material toward a turbid effluent outlet with a minimum of deposition on the electrode plates. A given portion of clear effluent was removed from the top portion of the unit. Minimizing deposition is highly desirable in order to

extend the time period between cleaning cycles (polarity reversal and flushing), thereby increasing the efficiency of the system.

As clay moves to, and is deposited on, the anode by electrophoretic transport in a clay-water suspension it becomes stationary. However, it is not free from electrokinetic effects, the electric field being still present. The phenomenon is, however, changed from electrophoresis to electroosmosis. In the latter case, due to the electric field, water moves past the stationary charged clay particle rather than vice versa as occurs with electrophoresis. As this happens, clay layers deposited next to the anode slowly become dryer as the water moves away from the positive plate toward the liquid medium. This drying will continue progressively outward from the plate building up an ever thickening layer of drying deposited clay. As this happens the resistance to current flow is substantially increased and thus a lowering of the electric field strength occurs. When the field strength decreases significantly it becomes necessary to reverse the voltage polarity to remove the clay layer and then flush the system. Thus it is easily seen why it is desirable to keep clay deposition to a minimum.

Figure 2 illustrates the principle upon which the parallel plate model was designed. A downward movement of colloids with minimum clay deposition was achieved by constructing banks of horizontal, parallel plates with a given vertical separation between oppositely charged plates in the same bank. A slight drop from one bank of electrodes to

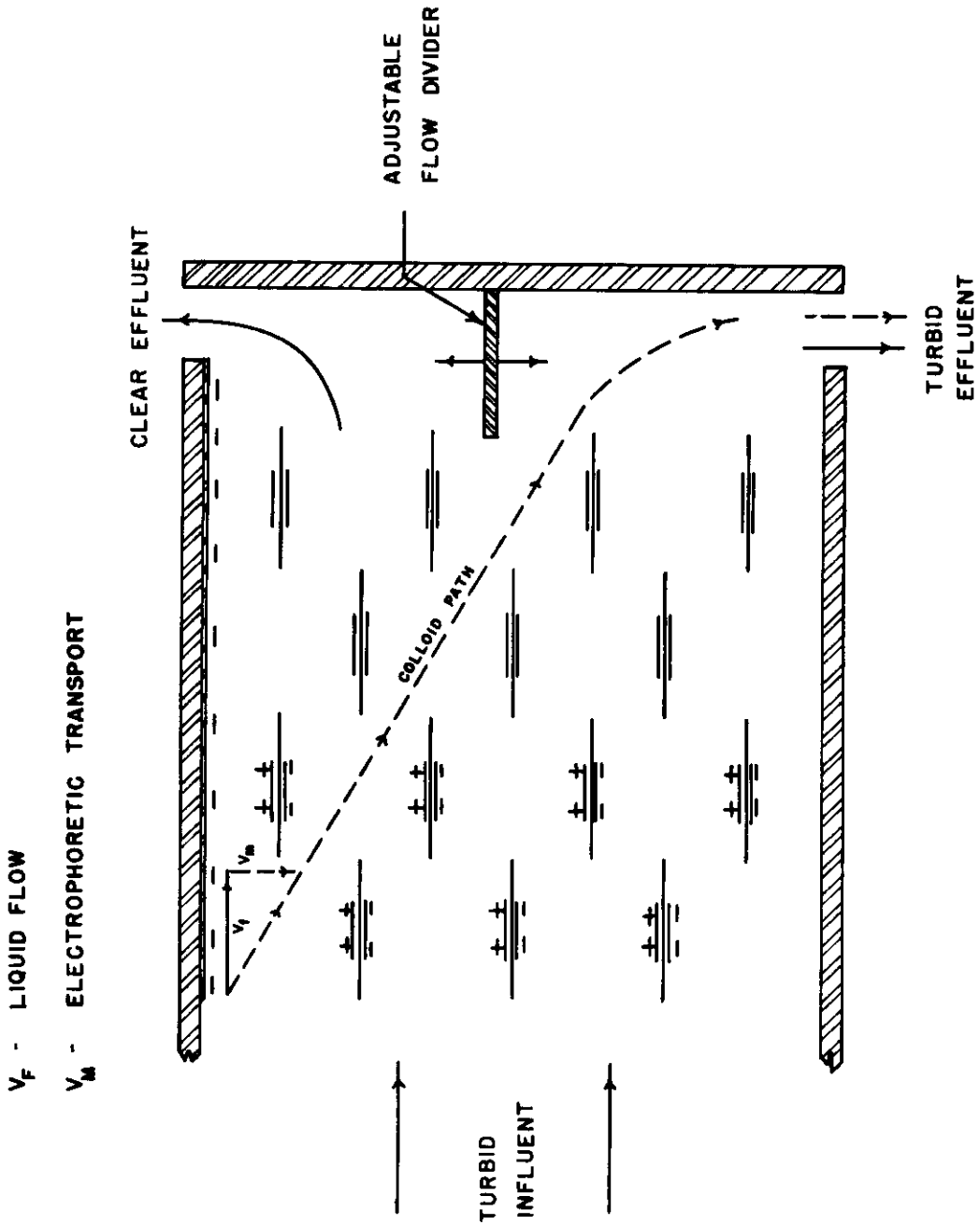


Figure 2. Schematic diagram of parallel plate design principle.

the adjacent bank of electrodes in the direction of flow gives a downward gradient to the plates toward the effluent end of the model.

Electrophoretic transport velocity is directly proportional to the electric field strength between two electrodes as is shown in the Helmholtz-Smoluchowski equation for the electrophoresis of suspended particles:

$$V = \frac{D \chi \zeta}{4 \pi \eta} ,$$

where

V = transport velocity,

D = dielectric constant,

χ = electric field strength,

ζ = zeta potential, and

η = absolute viscosity.

Thus the hydraulic flow rate in the model and the electric field strength between the plates may be adjusted so as to give a downward gradient to the suspended material that is equal to the downward slope between the parallel electrodes, resulting in minimum clay deposition on the plates.

It is obvious that a prior knowledge of the zeta potential or electrophoretic mobility of the colloidal particles is necessary in order to determine the proper electrical field so that the resulting transport velocity when added vectorially to the flow velocity will give a resultant equal to the gradient of the parallel plates. The electrophoretic mobility may be defined as the transport velocity

when the electrical field strength is unity and has the units microns sec^{-1} volts $^{-1}$ cm. These mobilities were determined by the microscope method of electrophoresis and are discussed later.

Porous Filter Design

A porous filter electrophoretic clarification model was designed much along the same line as the Bier unit but with some definite modifications. The Millipore and Gelman type filters were replaced by a porous sintered stainless steel filter which allowed the filter itself to become the negatively charged cathode. By making the filter the negative electrode, the passage of the electric field through the purified water is eliminated. This substantially lowers the electrical potential necessary for a required field strength especially if the conductivity of the treated water has been lowered due to the electrophoretic process. A cost reduction of at least one-half should thus be realized.

Figure 3 gives a schematic diagram of this electrophoretic system design. Turbid influent suspension enters the unit under a constant head and at a constant rate at the top of a model. When operating electrophoretically, at low conductivities, the negatively charged particles, as they pass between the electrodes in the inlet compartment, will migrate toward the positive electrode due to the electrophoretic transport. A portion of the water is drawn through the stainless steel filter at a constant rate and exits the model. If

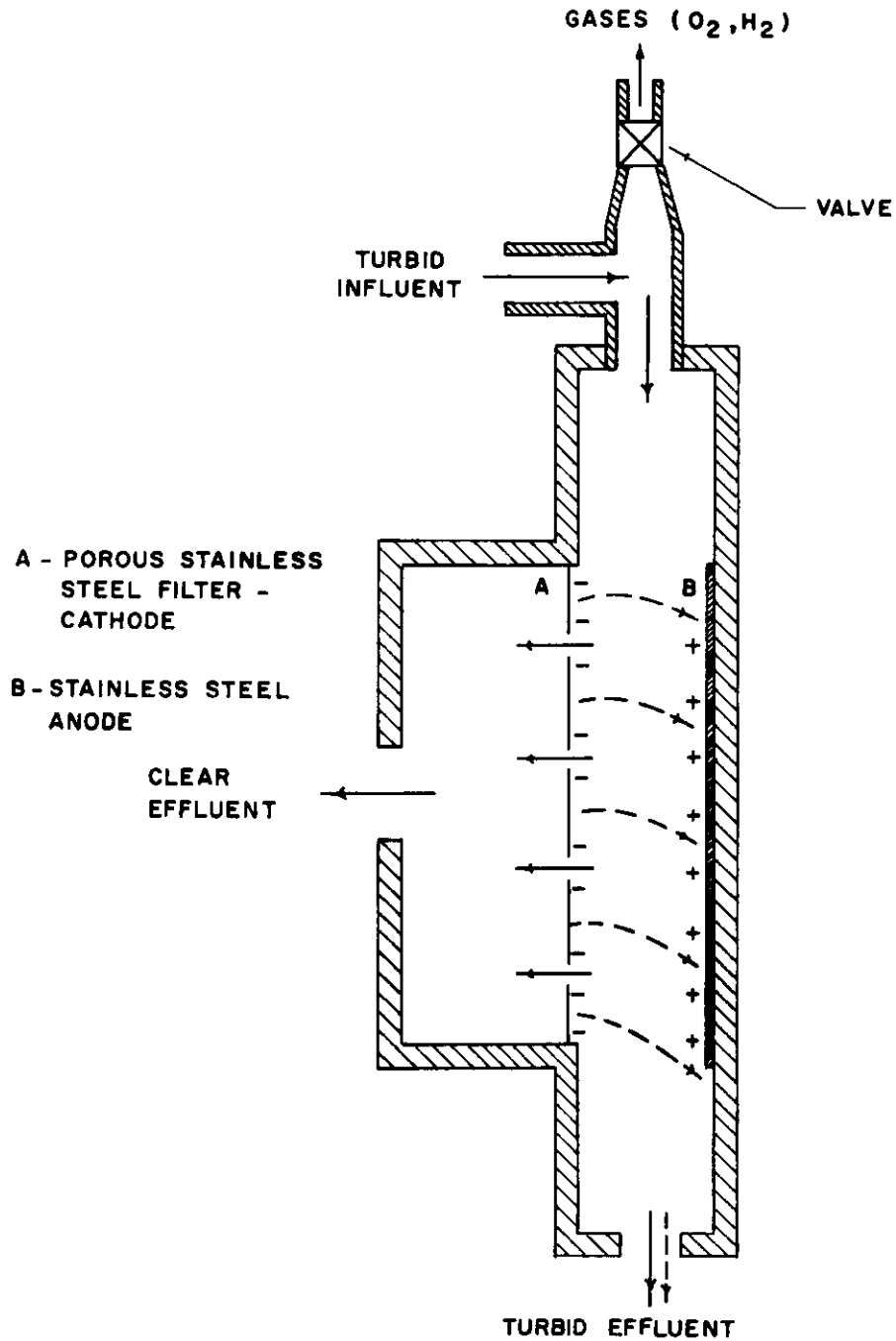


Figure 3. Schematic diagram of the porous filter system design. Solid arrows indicate water flow. Dotted arrows are colloidal transport.

the velocity of particle migration is maintained equal to or greater than the hydraulic flow through the filter in the opposite direction, the colloids should not pass through the filter nor should the filter clog. In effect, the result is a continuously backwashing filter. Turbid effluent is withdrawn from the inlet compartment at the bottom.

When this model is operating electrochemically, at high electrical conductivities, the filter should likewise stop the flocculated particles, allowing them to be carried downward by gravity and swept out the bottom outlet. This model may necessitate periodic backwashing when operating in this manner. The gaseous products of electrolysis escape out the top of the unit.

Electrode Grid Design

Two parallel electrode grids constitute the third electrode arrangement. Water enters the unit from one to four directions between the grids where it divides going toward two effluent outlets. This is illustrated in Figure 4. This design facilitates very close electrode spacing without appreciable flow velocity and turbulence. The electrode grid rather than a porous plate also takes advantage of any coagulating electrochemical reactions without the possibility of a plugging problem.

A computer program utilizing a method of numerical relaxation for solving three dimensional flow nets was developed in order to obtain the optimum dimensions of this type of electrode grid. It was also

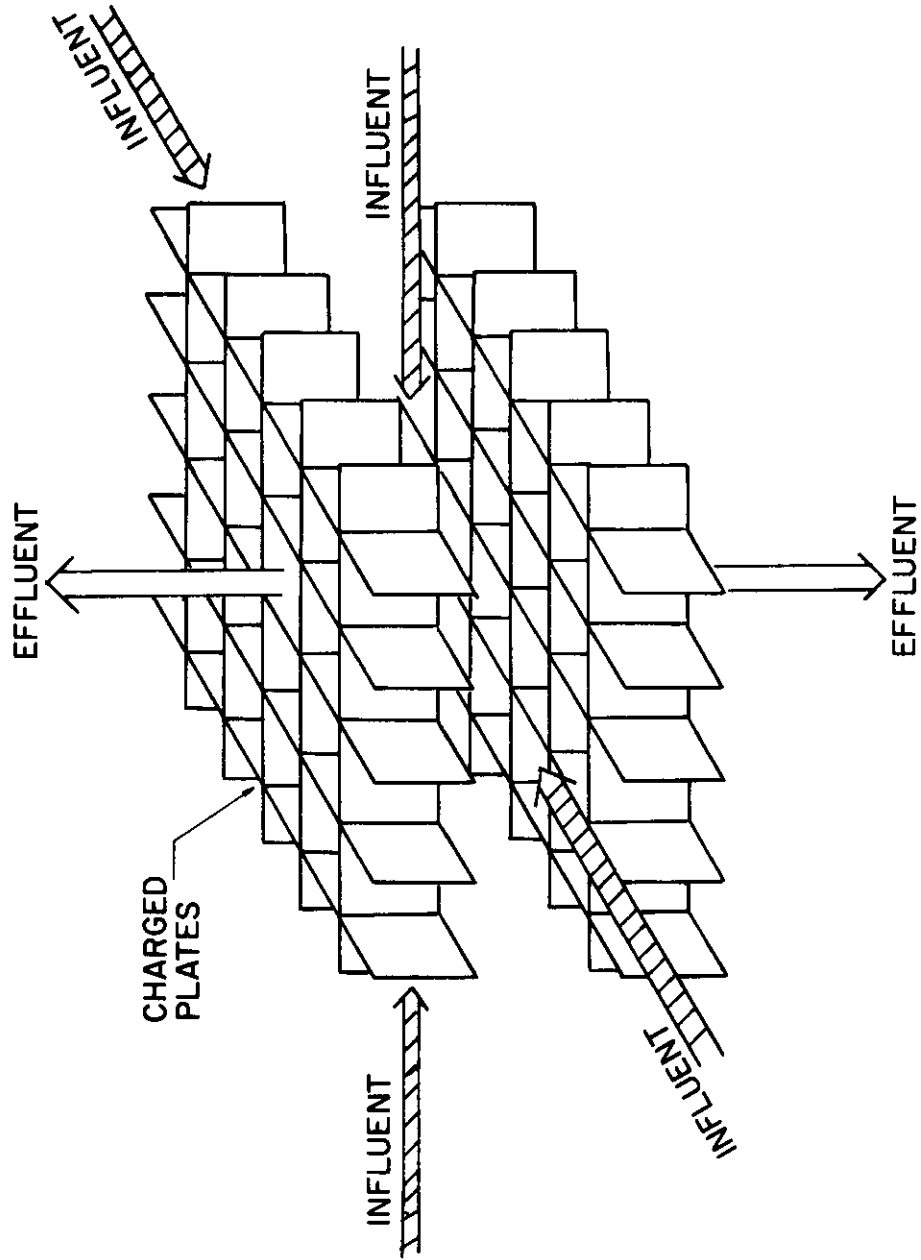


Figure 4. Electrode grid design concept.

used to determine the relationship between grid potential and electric field strength between the grids.

Consider that the steady state electrical flow field between the grid electrodes is composed of a very large but finite number of grids or series of lines running parallel to each of the three axes. Next consider an arbitrary point 0 in the flow field with a potential of ϕ_0 surrounded at distance "a" on each axis by points 1, 2, 3, 4, 5 and 6 with their respective potentials. This is illustrated in Figure 5. Next consider point A halfway between point 0 and point 1 and Point B halfway between point 3 and point 0. This is also shown in Figure 5.

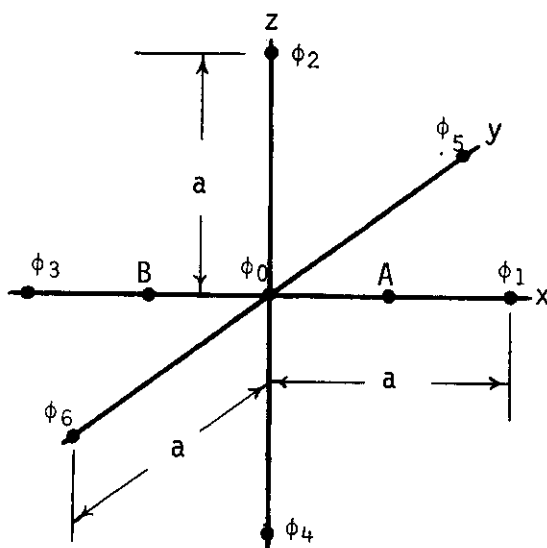


Figure 5. Arbitrary point 0 in hypothetical flow field for computer analysis.

The following partial derivatives may be represented in finite difference form as follows:

$$\left. \frac{\partial \phi}{\partial x} \right|_A = \frac{\phi_1 - \phi_0}{a}, \quad \left. \frac{\partial \phi}{\partial x} \right|_B = \frac{\phi_0 - \phi_3}{a},$$

therefore

$$\frac{\partial^2 \phi}{\partial x^2} = \frac{\frac{\phi_1 - \phi_0}{a} - \frac{\phi_0 - \phi_3}{a}}{a} = \frac{\phi_1 - 2\phi_0 + \phi_3}{a^2}.$$

Likewise in the y and z directions

$$\frac{\partial^2 \phi}{\partial y^2} = \frac{\phi_5 - 2\phi_0 + \phi_6}{a^2} \quad \text{and}$$

$$\frac{\partial^2 \phi}{\partial z^2} = \frac{\phi_2 - 2\phi_0 + \phi_4}{a^2}.$$

Laplace's equation describing three dimensional steady state potential flow is

$$\frac{\partial^2 \phi}{\partial x^2} + \frac{\partial^2 \phi}{\partial y^2} + \frac{\partial^2 \phi}{\partial z^2} = 0.$$

Rewriting in finite difference form gives

$$\frac{\phi_1 - 2\phi_0 + \phi_3}{a^2} + \frac{\phi_5 - 2\phi_0 + \phi_6}{a^2} + \frac{\phi_2 - 2\phi_0 + \phi_4}{a^2} = 0$$

and simplifying we have

$$\phi_0 = \frac{1}{6}(\phi_1 + \phi_2 + \phi_3 + \phi_4 + \phi_5 + \phi_6).$$

The next step is to assume a network of values within the framework of the boundary conditions which exist. This was represented in the form of a three dimensional array in the IBM 360/65 digital computer. From the initially assumed network of values, through the elimination of residual errors by numerous iterations, the eventual

satisfaction of both the differential equation and assigned boundary conditions was achieved.

The computer solution by the relaxation process is shown in Figures 6 through 9. The maximum potential drop between the electrode and locations in a plane which coincides with the bottom of the electrode grid, will occur at a point in the center of each grid dimension. The minimum field strength between the grids will occur between this point and an identical point in the oppositely charged grid. Figures 6 and 7 express this as the minimum potential available in percent of the potential applied to the grid itself. This is plotted versus relative electrode spacing and relative vertical grid depth. The common denominator of these two dimensionless values is the horizontal dimension of each small grid section or is the side dimension of each square void space within the grid. The average potential available between the grids, again expressed as a percentage of grid potential, is presented in Figures 8 and 9 and is plotted versus the same relative values.

The curves in Figures 6 and 8 are of course asymptotic to 100 percent available potential. However, for practical purposes, any electrode spacing to grid width ratios of 2.0 or over will give efficient service. Likewise ratios of less than about 1.2 should be avoided. Figures 7 and 9 indicate that a vertical grid depth to grid width ratio greater than approximately 1.6 is not necessary. In addition ratios of less than about 0.8 will substantially lower the efficiency of the system.

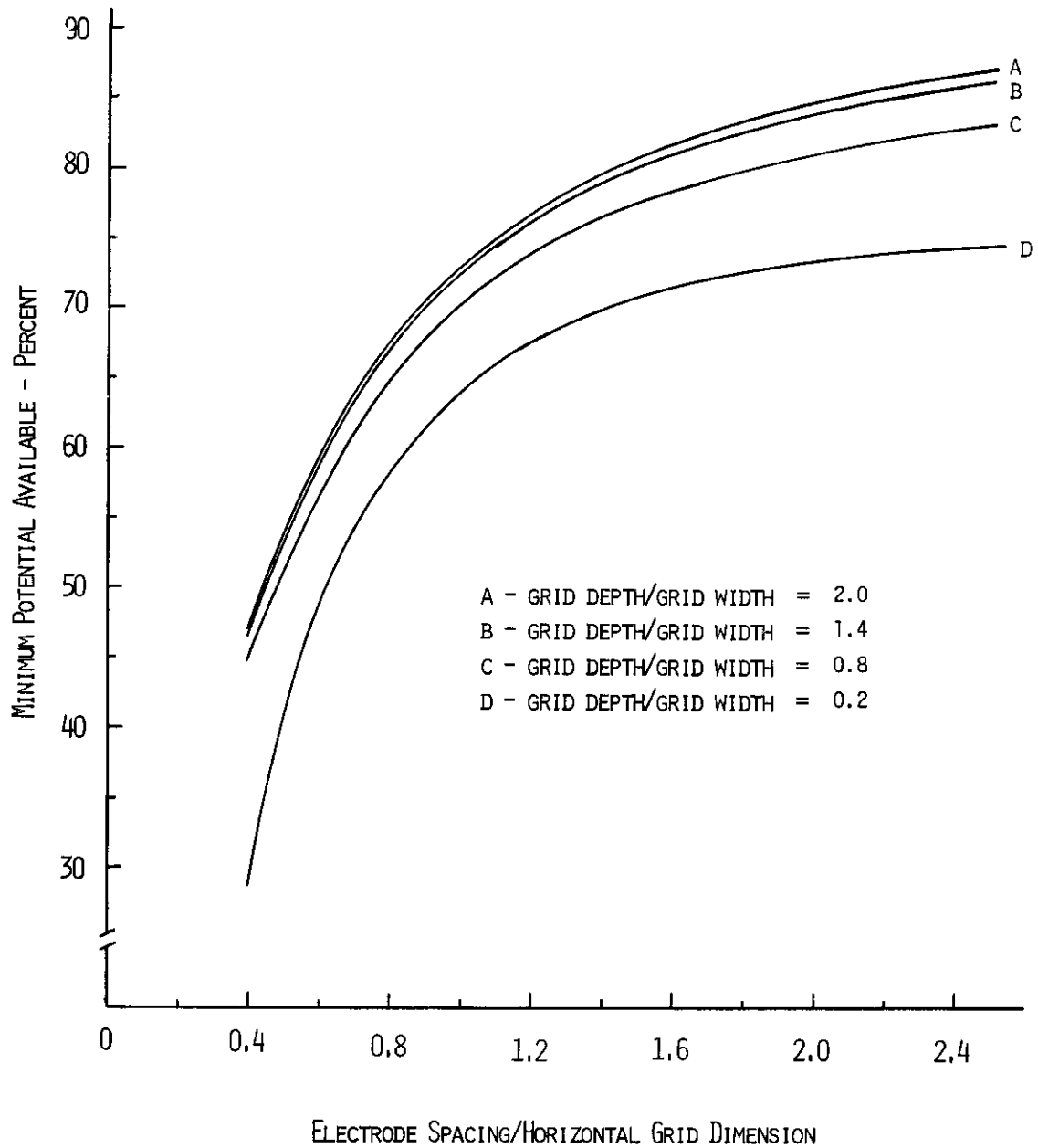


Figure 6. Minimum potential available versus relative electrode spacing.

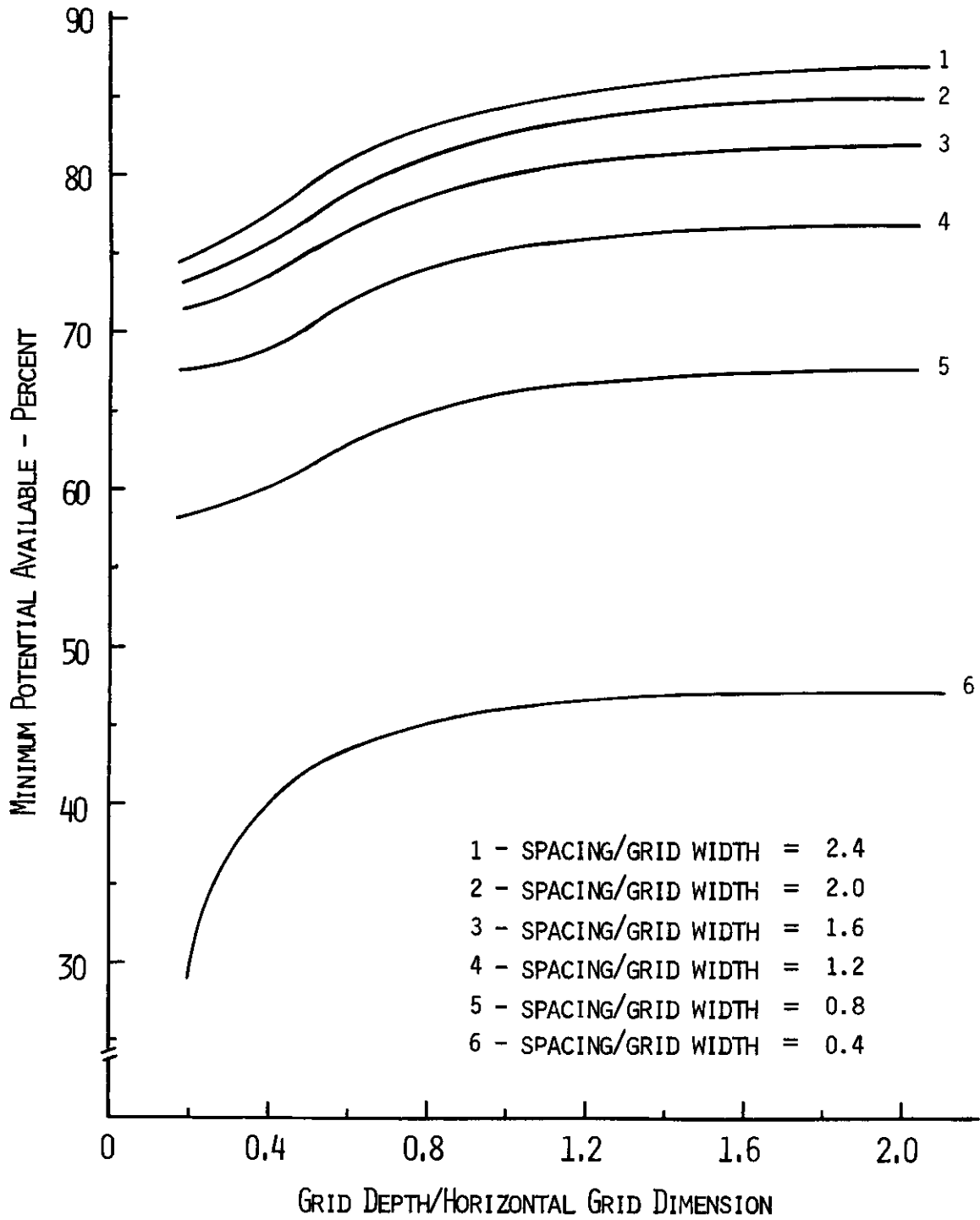
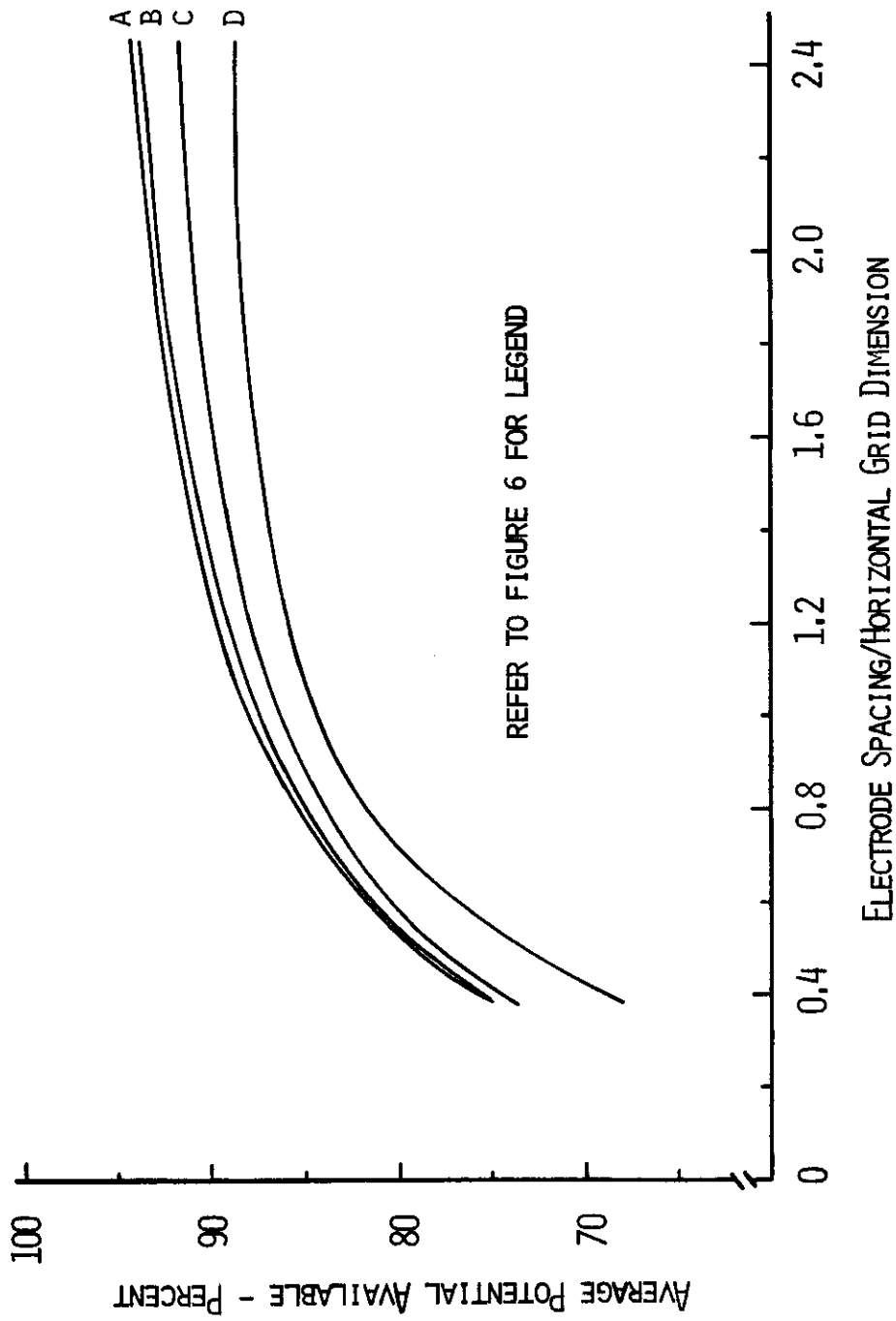


Figure 7. Minimum potential available versus relative grid depth.



REFER TO FIGURE 6 FOR LEGEND

Figure 8. Average potential available versus relative electrode spacing.

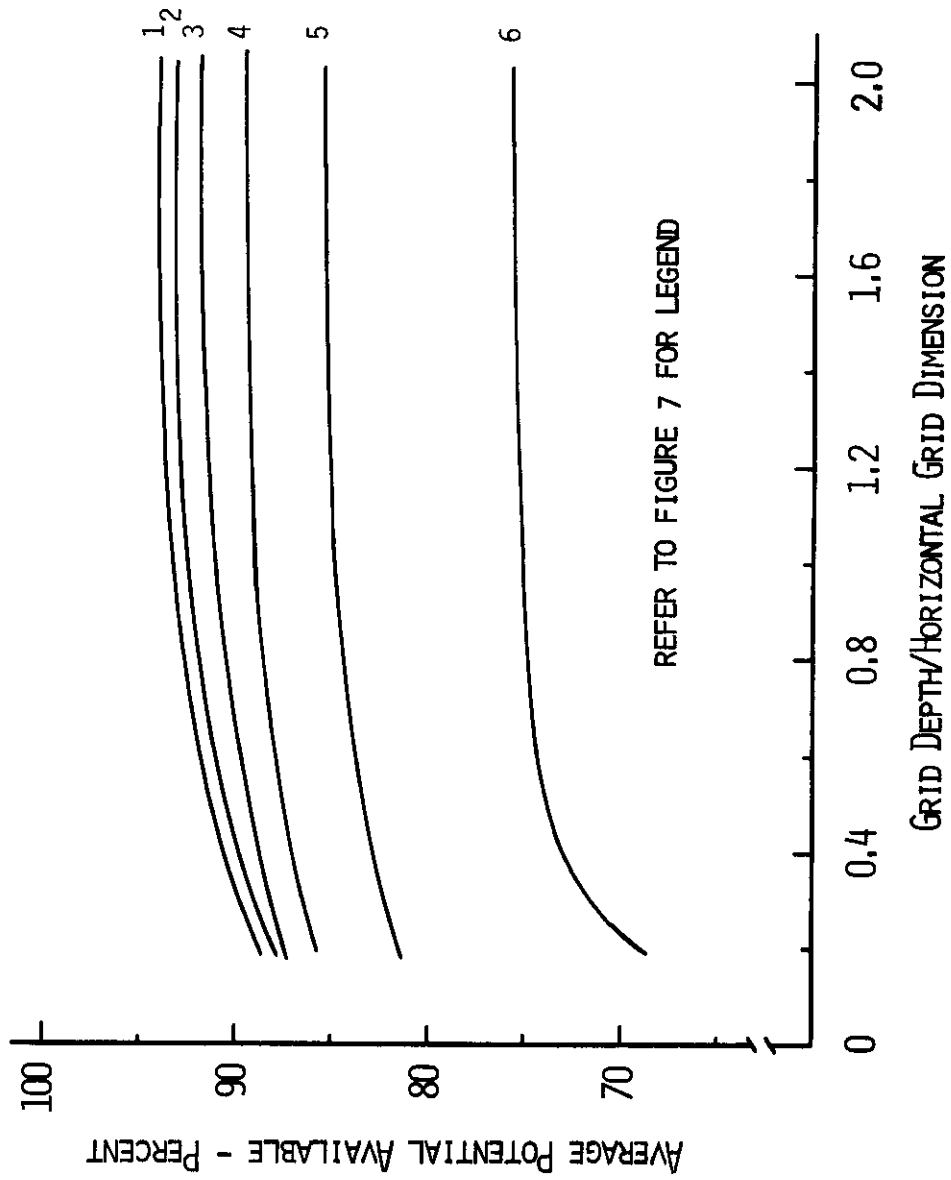


Figure 9. Average potential available versus relative grid depth.

CHAPTER IV

SYSTEMS EVALUATION

Evaluation of the electrophoretic water clarification models was conducted at the Agricultural Engineering Research Laboratories of Texas A&M University.

Equipment and Materials

Experimental Models

Several experimental models were constructed in order to evaluate the design concepts advanced in Chapter III. The following is a brief description of the constructed models.

Parallel plate model. The parallel plate model consisted of 20 banks of horizontal, parallel electrodes with 10 electrode pairs per bank. The negative and positive plates of the electrode pair were separated by a 0.1-inch lucite insulating plate. Each electrode plate was 10 cm. wide and 4 cm. long (in direction of flow). There was a 2 cm. separation between plates in the same bank and a one cm. drop from one bank of plates to the adjacent bank of plates in the direction of flow. A negative plate extended along the entire top length of the model and a positive plate extended along the bottom. The electrode system was housed in a lucite shell 10 cm. wide, 22 cm. tall, and 183 cm. long. The cross sectional area of the model was 182 cm².

The electrode material chosen for this model was type 304 stainless steel for its superior corrosion resistance.

Each bank of electrodes could be activated by means of toggle switches located on the control board. In addition, the top half or bottom half of each bank could be controlled by means of a switch. The plate along the entire top of the unit and the plate along the bottom were also controlled in the same way. A view of the parallel plate model is presented in Figure 10 and is shown in operation in Figure 11.

Porous filter model. A schematic view of this model is shown in Figure 3. The outside dimensions are given in Figure 12 and the unit is shown in operation in Figure 13. As indicated earlier, the filter indicated by the dotted line in Figure 3 is made of porous sintered stainless steel. These filters are available commercially with porosities ranging from 5 microns up to 165 microns. A wide variation in porosities was desirable for testing; therefore, filters of 5 micron and 165 micron porosities were both tested. The effective filter area was 686 cm² or about 0.75 square feet. The construction of this model was also of transparent lucite with the exception of the electrodes.

Electrode grid model. A schematic view of the electrode grid model as it was originally designed and constructed is shown in Figure 14. Figure 15 shows it in operation. The turbid influent enters the unit through 4 inlets and passes between electrode grids

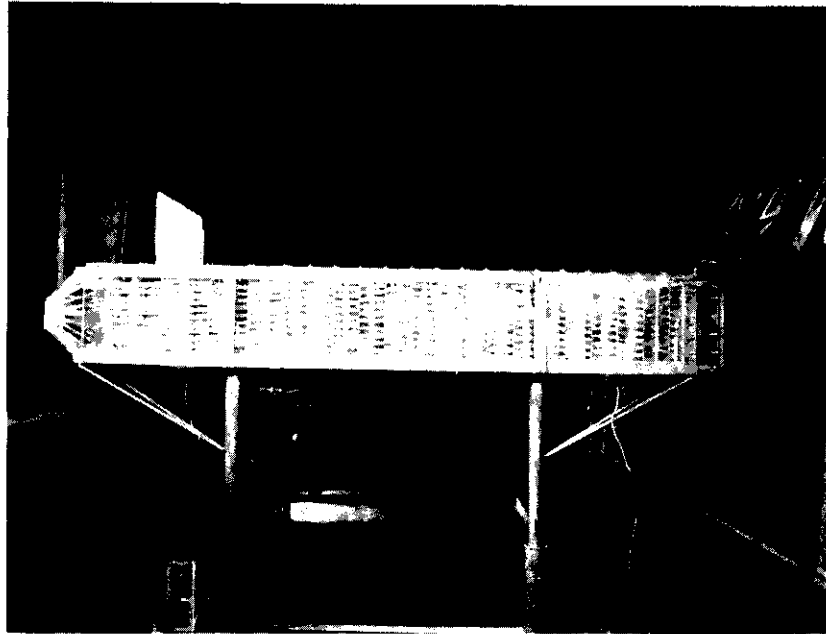


Figure 10. Parallel plate model.

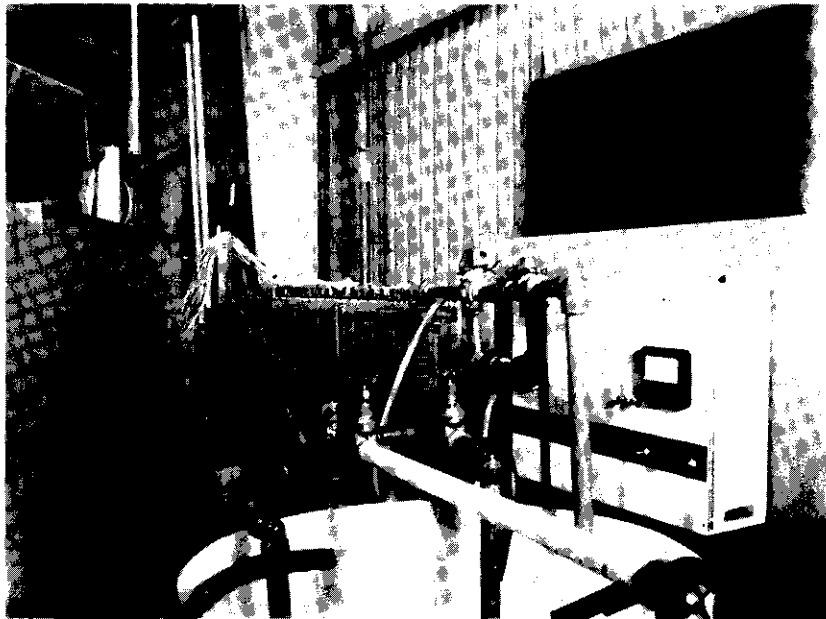


Figure 11. Parallel plate model in operation.

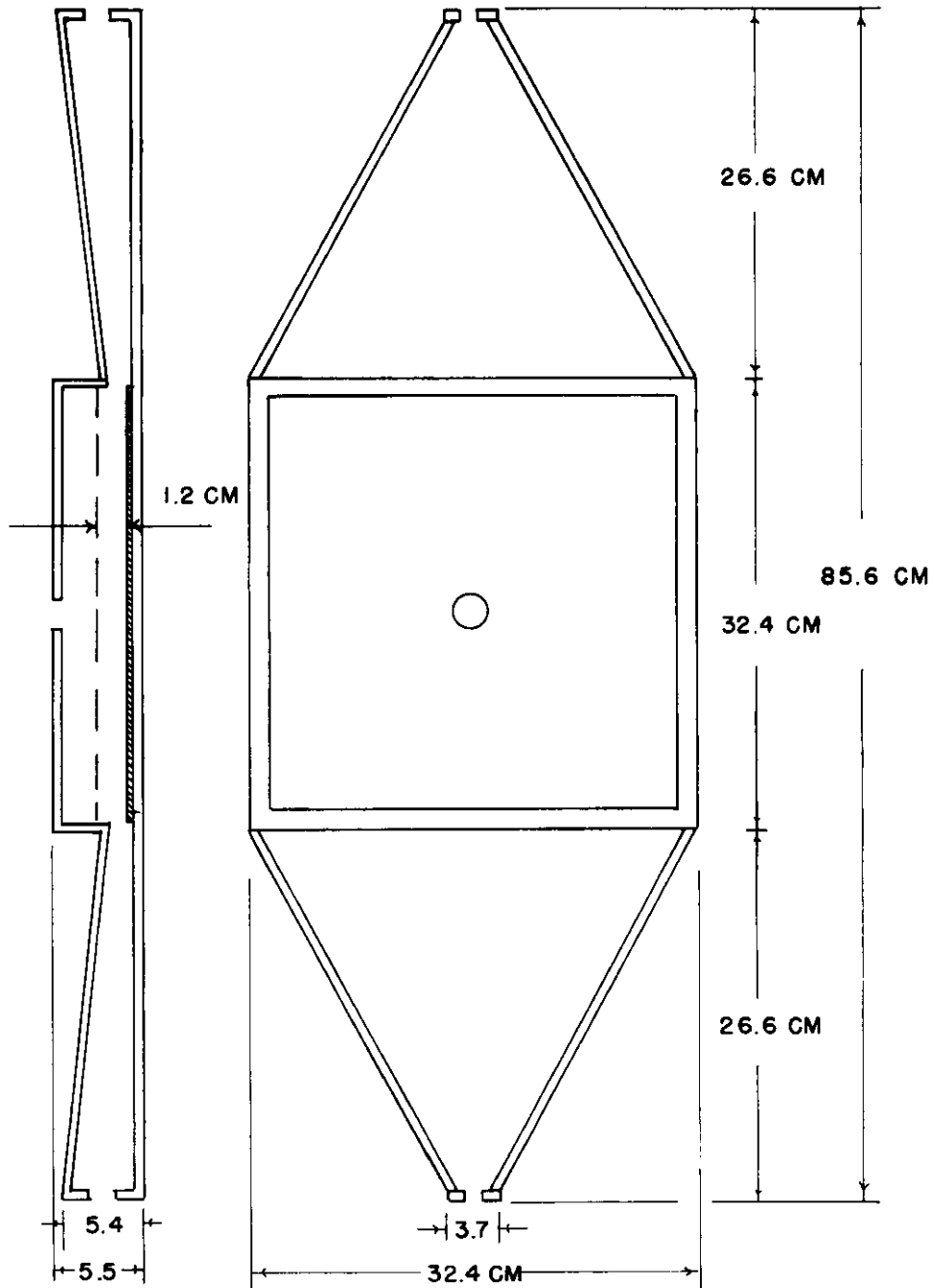


Figure 12. Outside dimensions of porous filter model.

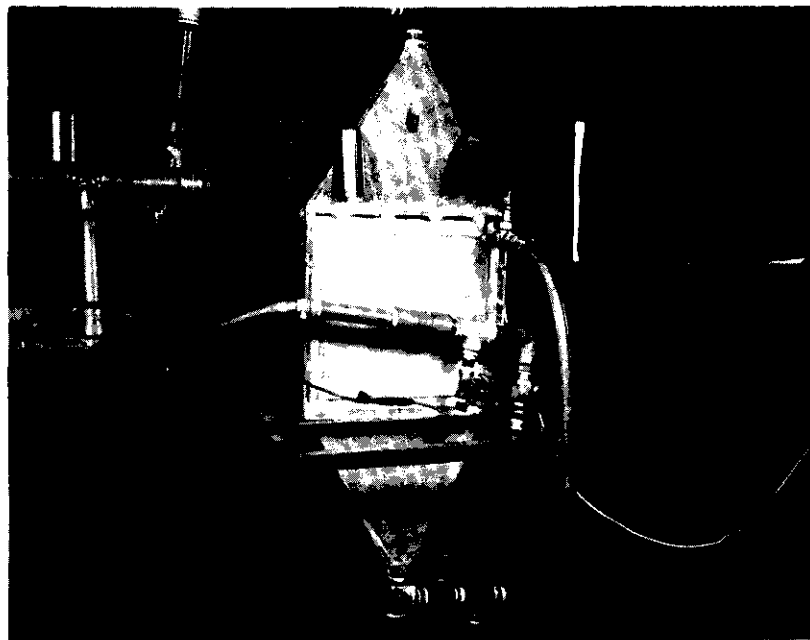


Figure 13. Porous filter model in operation.

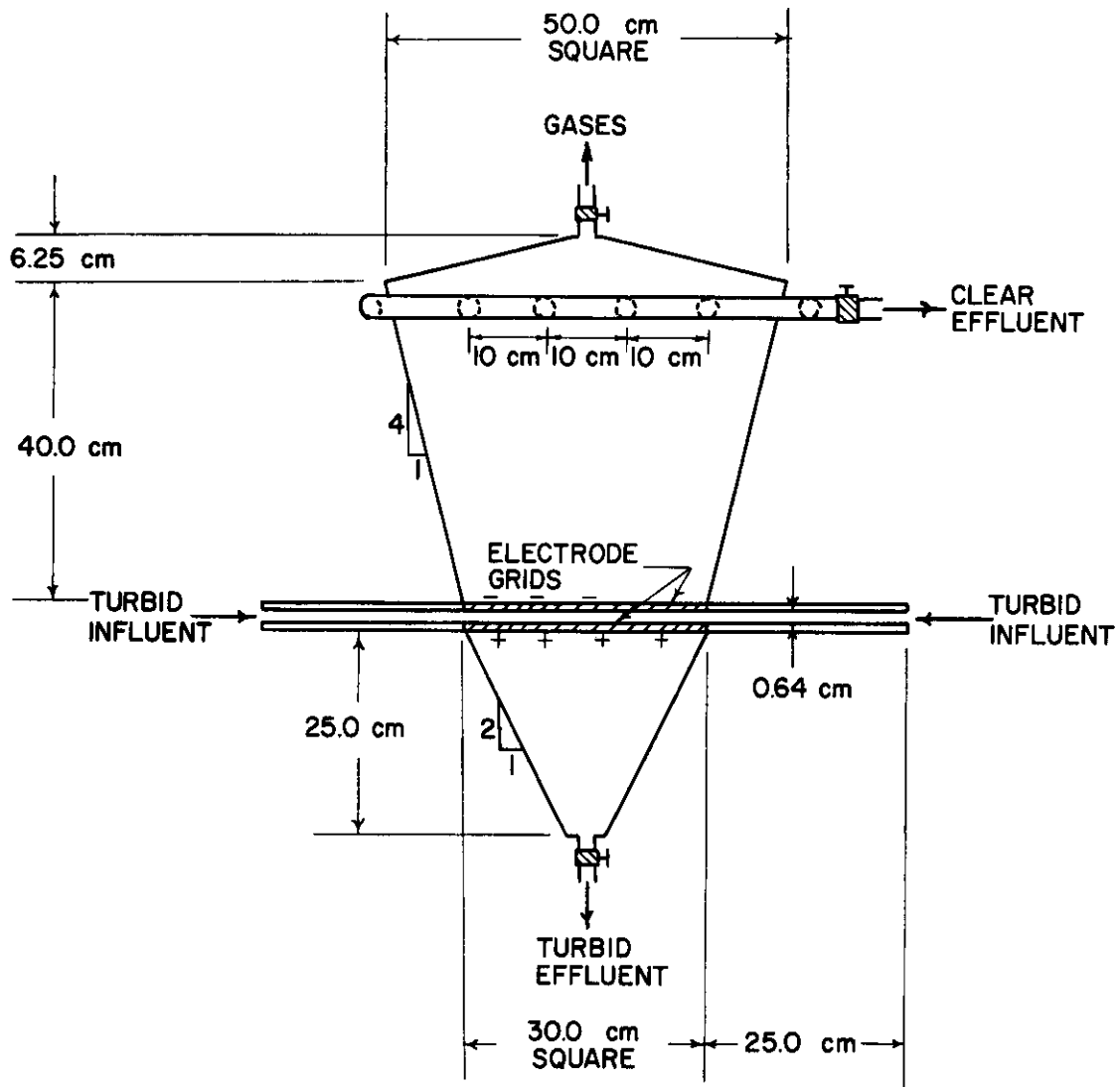


Figure 14. Original design of electrode grid model.

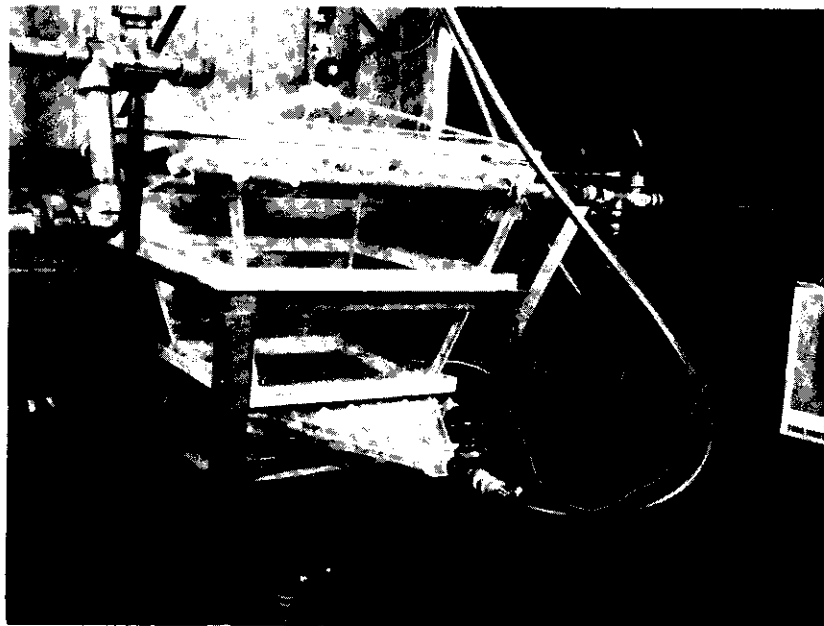


Figure 15. Original electrode grid model in operation.

which are 0.64 cm. apart. The top grid is negative and the bottom grid positive. The inlet is 30 cm. square at the grid or has a cross sectional area there of 900 cm². The electrophoretic transport being downward, the turbid effluent is taken out the bottom of the model.

For waters with high electrical conductivities and corresponding high rates of electrolysis, much of the clay will be carried upward by the rising bubbles of gas where flocculation will take place. However, before reaching the clear effluent outlets, particles silt size or larger should settle out when operating at the design clear flow rate of 2.5 gal. hr.⁻¹. The gases are allowed to escape out the top of the model. The clear effluent leaves through 16, 1.25-cm. ports (4 per side) and is transported away through 1.25-cm. PVC plastic pipe. The centerline of the ports is 38 cm. above the top grid of the electrodes. The electrode material was type 304 stainless steel for one set of tests and aluminum for another and the entire model was constructed of 1.25-cm. clear lucite.

The 0.64 cm. separation between the grids called for a void width dimension of approximately 0.3 cm. and a grid depth of about 0.5 cm., according to design parameters developed in Chapter III. However, due to the difficulty of construction a void width of 1 cm. was used along with a grid depth of 1 cm. Again referring to Chapter III, this resulted in an average field strength of 81.5 percent of that obtained by

two solid plates at the same potential. At the center of the voids the minimum field strength was only 59.5 percent of the grid potential.

With most waters, except those with extremely low electrical conductivities, this design appeared to be adapted primarily to electrochemical flocculation as a clarification means. Bubbles rising from the electrodes were sufficient to completely disrupt the electrophoretic transport. In this first electrode grid model the flocculation process was likewise not greatly successful, removing an average of only 45 percent of the influent clay concentration. This was attributed to the large area available above the electrodes for the process to take place (900 cm² at the electrodes and 2,500 cm² near the outlets).

When precipitates are first formed by chemical reactions they are crystals of molecular size. A small initial increase in size of these colloidal crystals is caused by Brownian motion. However, the completion of this process requires gentle turbulent mixing of the suspension. It has been experimentally established that the rate of flocculation at the point is directly proportional to the mean velocity gradient at that point. The rate is also greatest for a high concentration of particles of large size (10, 22).

The velocity gradients generated in this system are primarily from the motion of the rising bubbles of electrolysis products in the medium. Due to the large cross sectional area of the model it is

thought that the velocity gradients were neither great enough nor was the concentration of particles in a unit volume of water large enough for optimum flocculation to take place. As a result this model was modified as shown in the schematic diagram of Figure 16.

The top portion of the unit was first replaced by a 61-cm. lucite outlet section which converged in area toward the outlet. It then was turned on its side so that one of the original inlet sections became the mixing and floc forming area in addition to the space between the electrodes. The maximum cross sectional area available in this case for turbulent mixing was only 19.2 cm^2 . Since flocculation was the primary process involved in this model an alternating current of 4 cycles per hour was employed. This frequency selection was more or less arbitrary. However, the period between polarity reversal should be long enough to allow the yet unflocculated particles to concentrate near to, or adhere to, the anode. The higher concentration near the anode will facilitate the electrochemical flocculation in this region. When the polarity is changed the particles adhering to the anode should be discharged in a flocculated state. This will eliminate appreciable clay deposition on either of the grids for an extended period of time. The usable effluent containing settleable flocs was taken out of both horizontal outlets. Some extremely large flocs settled from between the grids to the bottom portion where they were removed periodically. Turbid influent entered the model between the grids horizontally from two sides. The model is shown in operation in Figure 17.

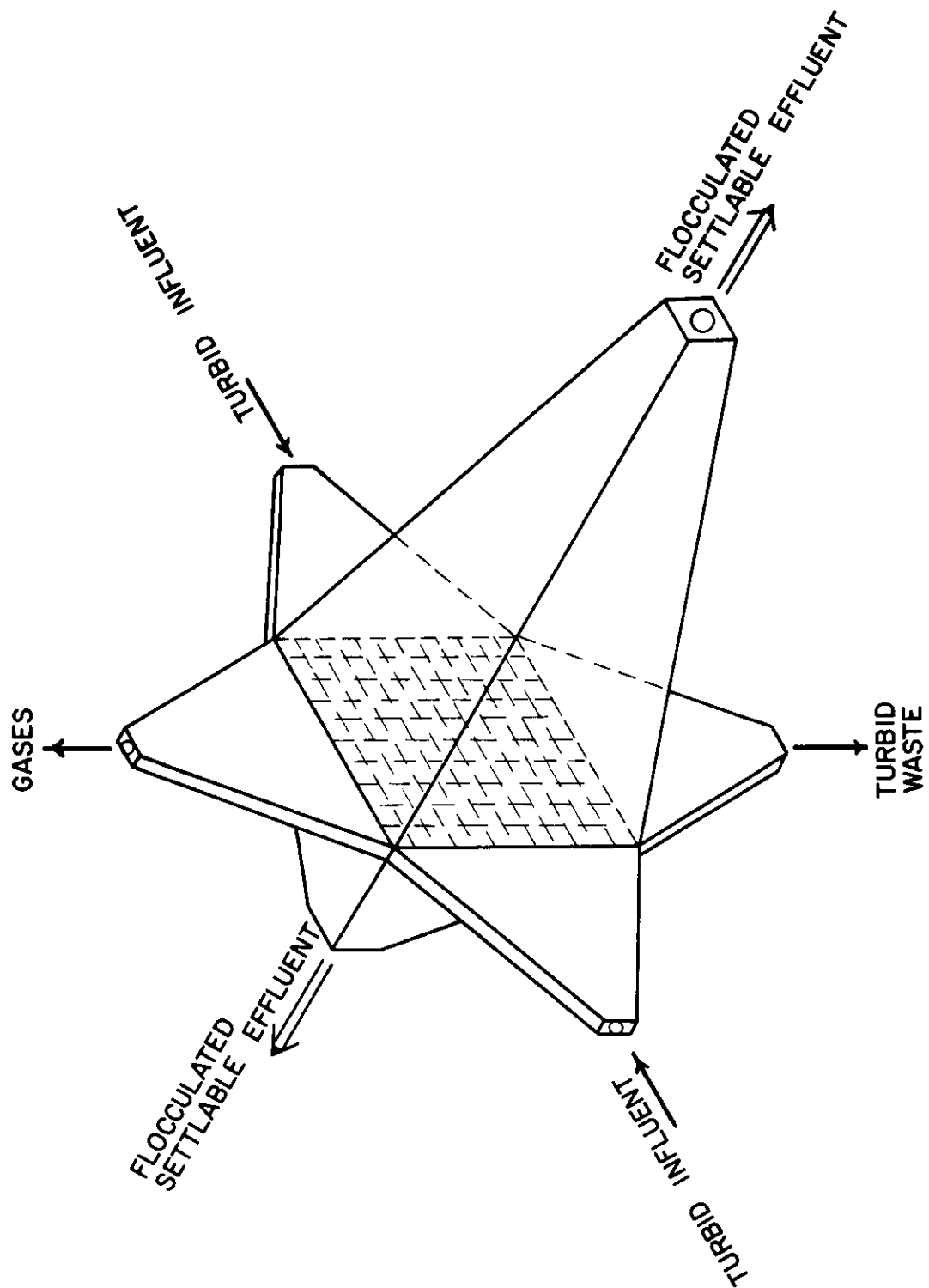


Figure 16. Modified electrode grid model.

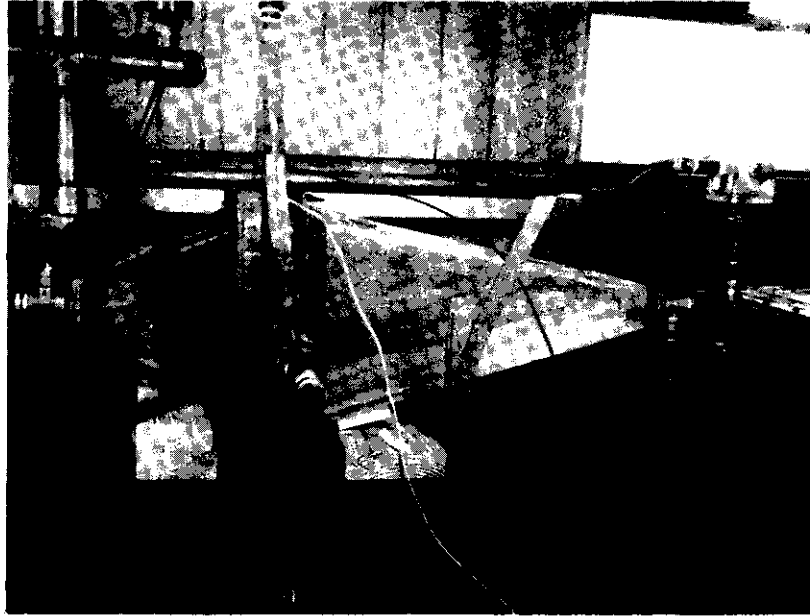


Figure 17. Modified electrode grid model in operation.

It is readily acknowledged that this modified design was an afterthought which utilized portions of the model which were designed for a completely different operation. The results, however, were satisfactory in that the principle of the process was proven sound yet the design as tested dare not be considered the optimum.

Since this unit operation was primarily for electrochemically induced flocculation the need for settling basins was evident. Two settling basins, one for each usable effluent outlet, were designed and constructed. Each settling basin was designed to remove all particles of 4 micron size and larger at a flow rate of 300 ml min^{-1} or about 5 gal hr^{-1} with a detention time of 2 hours. The dimensions of the basins were 30 cm. wide, 75 cm. long, and 20 cm. deep.

Test Equipment

Piping system. Figure 18 illustrates the piping system used to test the models. The influent test suspensions were mixed in a 150 gallon polyethylene tank from which they were pumped by a small centrifugal pump into a constant head device 12 feet above. A regulated bypass from the pump back into the mixing tank provided for pressure relief and also served to agitate the suspension. From the constant head device the water flowed through one-inch PVC pipe into the models. A parallel section was installed in the pipe to facilitate flushing. A micrometer valve was located in one leg and was the controlling device during normal operation. In the other leg was a

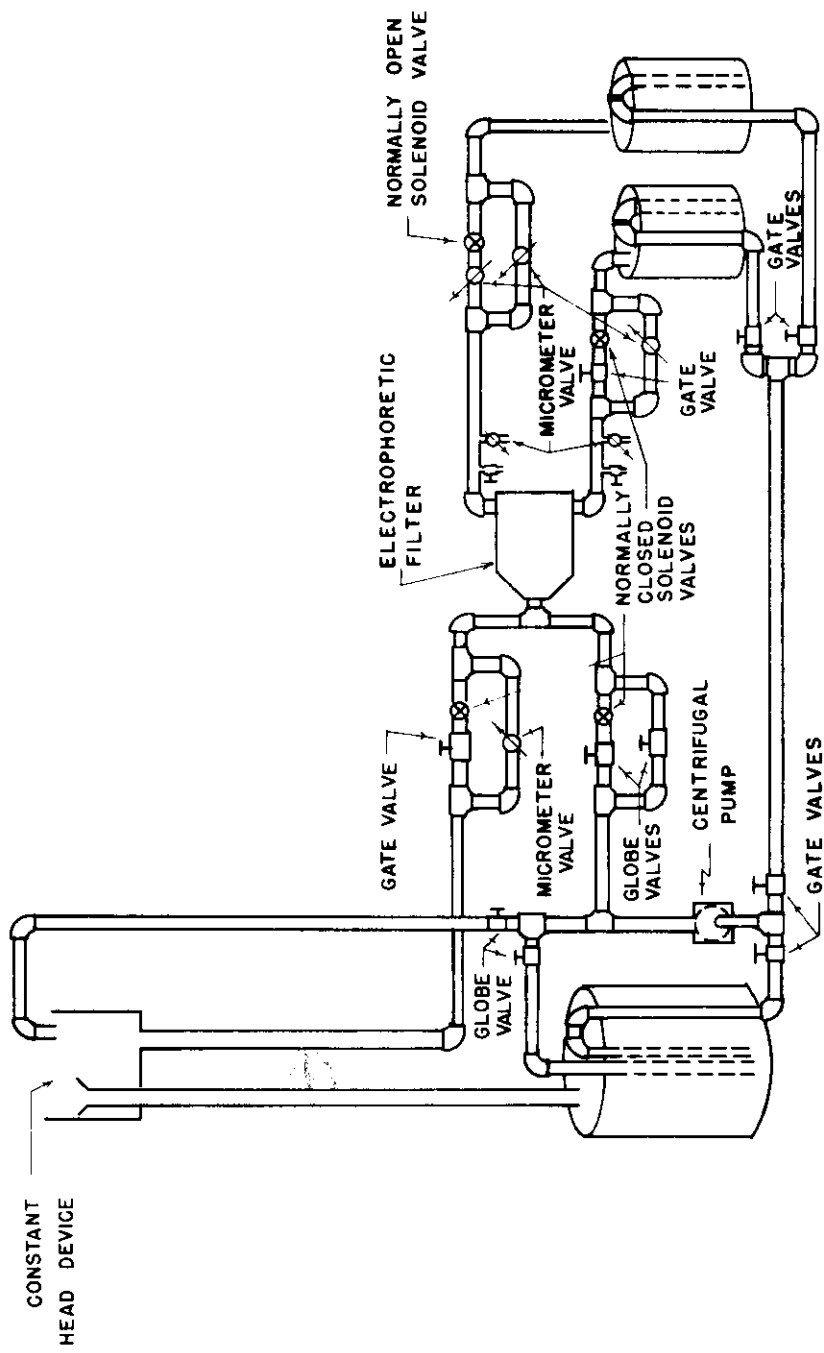


Figure 18. Piping system for testing models.

gate valve and a normally closed solenoid valve. During flushing the solenoid valve was actuated by a relay allowing a higher flow rate for flushing. A similar parallel pipe arrangement connected the pump directly with the model for the highest flow rate possible for flushing but was used only with the parallel plate model.

Clear and turbid effluent left the parallel plate model through one-inch PVC pipe also. Likewise, both outlets were equipped with a section of parallel pipe and solenoid valves to accomplish the flushing cycle. In the other models the effluent was taken off by a combination of PVC pipe and rubber hose. The effluent was caught in separate 100 gallon polyethylene tanks. Taps in the system near the outlets allowed samples to be taken immediately as the water left the model. The next outlet in this pipe was controlled by a micrometer valve and allowed a predetermined amount of effluent to be passed through the recording turbidimeters. Pipe from the effluent tanks back to the pump also allowed recirculation of the effluent if desired.

Electrical system. Power to the electrodes was supplied primarily by two 300-volt, 2.5-amp d.c. power supplies connected in parallel. When higher current densities were desired a 250-volt, 30-amp d.c. generator was used. Current reversal and flushing were achieved by a 4-pole single throw relay which reversed the polarity of the d.c. output and actuated the solenoid valves simultaneously. This relay was energized manually or by an automatic timer. As

mentioned previously, there existed numerous additional switches with which to control the location of the electric field in the first model.

Instrumentation

Effluent clay concentrations were determined by turbidity measurements from two continuously recording Hach surface scatter turbidimeters. The turbid effluent passed through a Hach Model 1861 meter which has a range of from 0.2 to 5,000 mg. $\text{SiO}_2 \text{ l}^{-1}$. The clear effluent was taken through a Hach Model 1889 turbidimeter with a range from one to 1,000 mg. $\text{SiO}_2 \text{ l}^{-1}$. The output signal from the turbidimeters was supplied to a Texas Instruments Servo/Riter II dual channel recorder from which a continuous record of effluent concentrations was obtained. A manually operated Hellige Model 8010 nephelometric type turbidimeter was used for periodic checks of the influent clay concentration and was also used for calibrating the surface scatter meters. A surface scatter turbidimeter is shown in Figure 19.

The electrical potential to the electrodes was measured accurately with a Keithley Model 610B electrometer. The current was obtained from ammeters located on the power supplies after they were calibrated by a precision ammeter. The electrical conductivity of the influent suspensions was measured with an Industrial Instruments

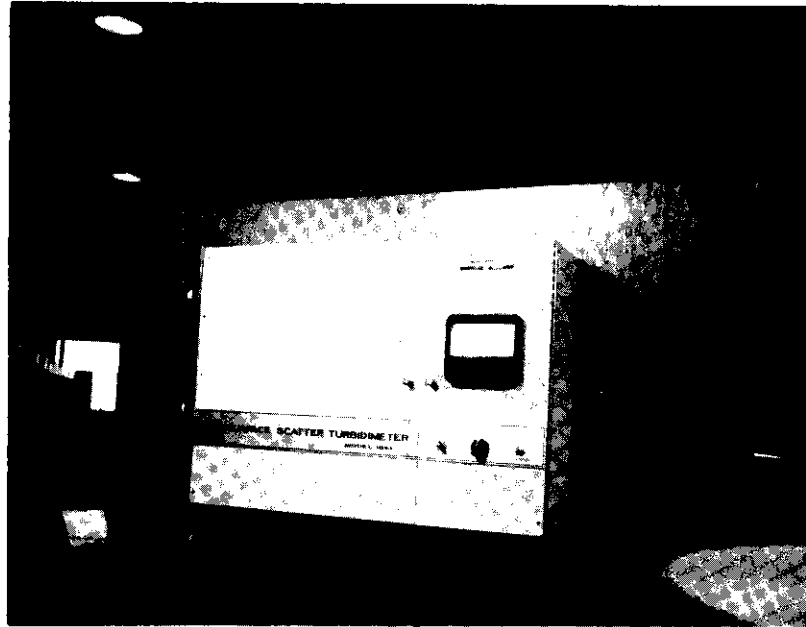


Figure 19. Surface scatter turbidimeter.

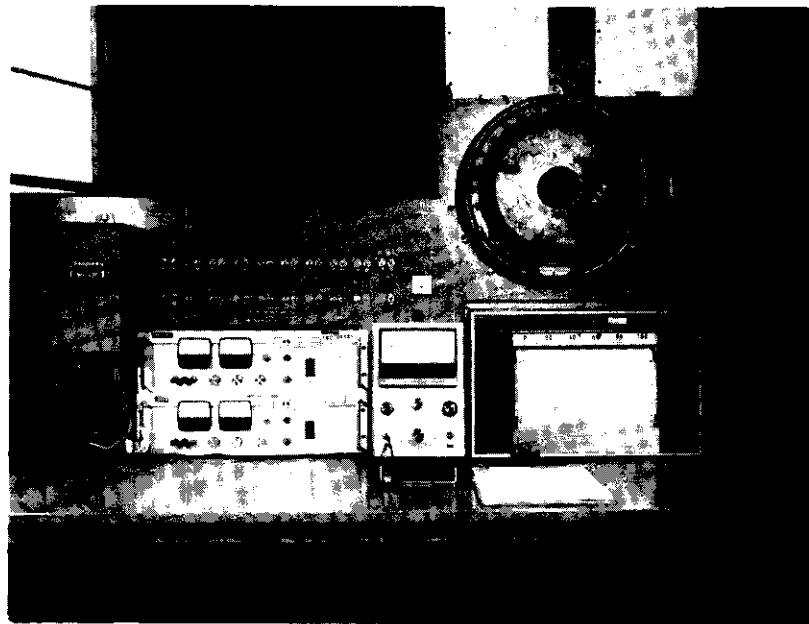


Figure 20. Control panel.

Model RC-12C1-P conductivity bridge with a range of 0.2 to 2.5×10^6 ohms. A view of the control panel is shown in Figure 20 and an overall view of the system is shown in Figure 21.

The electrophoretic mobility of the clay used in these tests and for characterizing numerous other Texas soils was determined by the microscope method of electrophoresis. A Riddick Zeta Meter was used for these determinations as is shown in Figure 22.

Materials

Water. Water for the test suspensions was supplied by an Illco-Way two bed ion exchange unit. This deionized water was mixed in various proportions with the tap water to give a wide range of electrical conductivities. The specific conductivity of the deionized water could be lowered to 1.0 to 2.0 micromhos cm^{-1} .

The chemical analyses of College Station tap water which was used in the test is given in Table I.

Clay. The clay suspensions used for evaluating the removal of charged particulate matter by the various models were prepared with colloidal kaolonite clay obtained from the Georgia Kaolin Company. This material has good brightness and showed excellent reproducibility for concentration determinations by turbidimetric methods. Table II gives the specifications of this clay as presented by the producer.



Figure 21. Overall view of experimental test system.

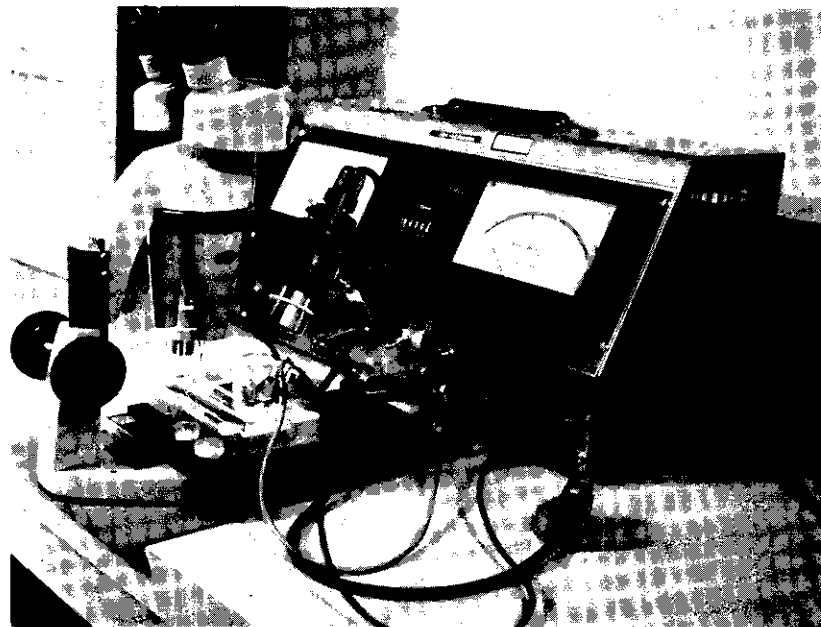


Figure 22. Riddick Zeta Meter.

TABLE I. Chemical Analyses - College Station, Texas Tap Water
(A Combination of Three Separate and Independent Analyses)

<u>Cations</u>	<u>mg l⁻¹</u>	<u>Anions</u>	<u>mg l⁻¹</u>
Calcium	1.6 to 3.25	Carbonate	4.0 to 7.5
Magnesium	0.9 to 1.0	Bicarbonate	403 to 420
Potassium	2.0 to 2.1	Sulfate	4.8 to 12.0
Sodium	190 to 470	Chloride	54 to 160
Manganese	<0.05	Fluoride	0.5
Iron	0.04	Nitrate	<0.4
Electrical Conductivity - 800 to 850 micromhos cm ⁻¹			
pH - 8.3 to 8.4			

The electrophoretic mobility of colloidal suspensions depends both on their concentrations and the type and amount of ions present in the suspending medium. In distilled water the kaolinite clay at 25°C displayed an electrophoretic mobility (EM) of 1.96 microns sec⁻¹ volts⁻¹ cm at a concentration of 10 mg l⁻¹; an EM of 1.70 at 50 mg l⁻¹; an EM of 1.46 at 100 mg l⁻¹; and an EM of 1.30 at 250 mg l⁻¹. The mobility increased about twofold in tap water due to the high proportion of sodium ions in this water. The average EM in tap water at a concentration of 75 mg l⁻¹ was 3.59 microns sec⁻¹ volts⁻¹ cm. It is apparent that a given colloid's environment has a great deal to do with its electrophoretic mobility. Therefore, in practical systems it would be desirable if the system were not extremely sensitive to variations in particle mobility. This was considered in

TABLE II. Clay Specifications

Median Particle Size	0.77 Microns
Brightness (G.E. % of MgO)	85.0 to 86.5 %
pH (20% in distilled water)	5 to 7
Refractive Index	1.56
Specific Gravity	2.61
Chemical Analysis	
Silicon Dioxide (combined)	45.1%
Aluminum Oxide	38.1%
Combined Water (Ignition loss)	13.8%
Titanium Dioxide	1.4%
Ferric Oxide	0.6%
Sodium Oxide	0.3%
Magnesium Oxide	0.2%
Potassium Oxide	0.1%
Calcium Oxide	Trace

the last two designs.

Coliform bacteria. Non-chlorinated municipal sewage effluent was used as a source of coliform bacteria in order to evaluate its removal by the models. The effluent was obtained from the City of Bryan, Texas sewage treatment plant. This non-chlorinated effluent exhibited a most probable number (MPN) of 110,000 coliform bacteria per 100 ml. of

effluent. One gallon of this sewage effluent mixed with 150 gallons of test clay suspension resulted in a test influent MPN between 800 and 1,000 coliform per 100 ml. of suspension.

Procedures

Parallel Plate Model

Flow rate, electric field strength, and number and locations of electrodes were the primary parameters evaluated in the parallel plate tests. The electrical conductivity of the water was maintained very low, ranging from 3 to 20 micromhos cm^{-1} . Influent clay concentrations were held close to 50 mg l^{-1} , the range being from about 40 to 60.

Each test consisted of a given flow rate operating at a given field strength. Flow rates of 880, 660, 440 and 220 ml min^{-1} were evaluated. These flow rates are considered optimum for field strengths of 100, 75, 50 and 25 volts cm^{-1} , respectively, with an average clay electrophoretic mobility of 2 microns sec^{-1} volts $^{-1}$ cm. For example, at a flow rate of 880 ml min^{-1} , an electric field strength of 100 volts cm^{-1} is necessary to establish a transport gradient corresponding to that of the parallel electrode plates. The field strength was actually varied from about 2 times the optimum down to small fractions of the optimum field strength.

As the tests began, the desired potential was applied to the electrode plates and the flow rates, both clear and turbid, were adjusted

to the desired rates. The recorder was zeroed and started. The initial temperature, conductivity, and concentration of the influent suspension were recorded. As the test progressed, the following data were obtained at frequent periodic intervals throughout the test:

- (1) Nephelometric turbidity measurements taken simultaneously at both effluent taps immediately after the effluent left the model and at the same time as the effluent passed through the recording turbidimeters; (This enabled a determination of the time lag between recorded data and what was actually occurring in the model. In addition, the nephelometric determinations at the meter were used to calibrate the surface scatter turbidimeters which were somewhat sensitive to flow rate.)
- (2) Nephelometric turbidity measurements of the influent suspension;
- (3) Influent temperature and effluent temperature, both clear and turbid;
- (4) Influent specific conductivity since this value changed with temperature;
- (5) Current to the electrodes; (This changed with temperature in the model and also as clay accumulated on the anodes.)
- (6) Clear effluent measurements obtained visually from graduations on the sides of the model.

The tests were continued until a steady state condition existed in the model. Turbidity of the clear or turbid effluent and the

percent of clarified effluent were not changing. After the tests, the models were flushed at high flow rates with the voltage polarity being reversed frequently. Then the model along with the turbidimeters were refilled with clear water for the next test.

Porous Filter Model

Parameters evaluated on the porous filter model consisted of filter porosity, flow rate, electric field strength, specific conductivity and influent clay concentrations. Porous stainless steel filters with 5-micron and 165-micron porosities were both evaluated at three flow rates. Flow rates investigated were 112, 224 and 336 ml $\text{min}^{-1} \text{ft}^{-2}$ which are flow rates which should produce optimum conditions, i.e., flow velocity equal and opposite to electrophoretic transport, at field strengths of 10, 20, and 30 volts cm^{-1} , respectively.

Specific conductivity of the influent suspensions was varied from less than 5 micromhos cm^{-1} to around 750 micromhos cm^{-1} at a constant flow rate of 224 ml $\text{min}^{-1} \text{ft}^{-2}$ and a constant influent concentration of around 50 mg l^{-1} . In another set of tests the influent clay concentrations were varied from approximately 3 mg l^{-1} to 93 mg l^{-1} . The flow rate was held constant at 224 ml $\text{min}^{-1} \text{ft}^{-2}$ and the specific conductivity was held between 300 and 400 micromhos cm^{-1} .

The test procedures for this model were much the same as those for the parallel plate model. Identical instrumentation and sampling techniques were used. For each given test condition, i.e., filter

porosity, flow rate, conductivity, and influent concentration, a field strength somewhat greater than optimum was first used. After a stable condition was reached at this field strength it was then lowered slightly and if a change took place conditions again were allowed to stabilize. In this manner it was determined, under each set of conditions, the minimum field strength at which the model could operate and still produce effluent of acceptable quality.

Electrode Grid Model

The primary variables considered in the electrode grid tests were clay concentration, electrical conductivity, and current density. Current density rather than electric field strength was considered since this operation was primarily electrochemical in nature and the rate of formation of electrolysis products is directly proportional to current flow.

The use of the settling basins changed the procedure slightly from that used in the porous filter tests. Samples to determine effluent concentrations were taken both as the water left the model and again when it left the settling basins. The surface scatter turbidimeters and the recorder were not used in these tests. Due to the detention time in the settling basins, a very slow rate of concentration change in water leaving the basins resulted. Thus periodic nephelometric determinations were adequate for analysis. With these exceptions the tests were identical to the porous filter tests.

Electrophoretic Mobility Determinations

Electrophoretic mobility of the test clay material was determined by the microscope method of electrophoresis. The procedure used is explained in detail by Riddick (24). In addition to the kaolinite used for testing, the EM of several different Texas soils was also obtained in order to characterize them in terms of ease of electrophoretic removal. These soils were evaluated from suspensions mixed in distilled water to a concentration of 75 mg l^{-1} .

Coliform Bacteria Determinations

The most probable number (MPN) of coliform bacteria was determined by procedures described in Standard Methods for the Examination of Water and Wastewater (27). The presumptive and confirmed tests were both employed.

Results

In the following analyses the cost figures include operating costs only and are limited to the power used directly for the electrophoretic or electrochemical process in the model. The cost of electricity was assumed to be \$0.02 per kilowatt hour in all cases.

Parallel Plate Model

As previously stated, flow rates of 880, 660, 440 and 220 ml

min^{-1} were evaluated. Operating at optimum field strength, 440 ml min^{-1} was found to be the most desirable as indicated in Figure 23. This corresponds to a flow velocity of 0.04 cm sec^{-1} . The percent clay removal could be maintained at a high level for all flow rates but in order to attain this the clear effluent percentage must drop at flow velocities in excess of 0.04 cm sec^{-1} . The production rate of clear effluent at this flow velocity and optimum field strength was approximately six gallons hr^{-1} at a cost of \$1.11 per 1,000 gallons.

An unanticipated phenomenon occurred which makes it possible to obtain good results at field strengths appreciably less than those considered optimum. Some heating occurred in the unit. Therefore, when the cooler turbid influent entered the model it immediately sank to the bottom. As a result, the electric field, instead of moving the clay to the bottom, needed only to hold it there. Lower than predicted optimum field strengths were observed to be adequate for this. A temperature differential of only 0.5 to one degree Centigrade was necessary for this to occur.

Again operating at the optimum flow of 440 ml min^{-1} , but at 1/4 the optimum field strength ($12.5 \text{ volts cm}^{-1}$), the clear effluent percentage dropped from 88% at optimum to 37%, as shown in Figure 24. This lowered the clear effluent production rate to $2.6 \text{ gallons hr}^{-1}$ but also reduced the operating cost from \$1.11 to \$0.25 per 1,000 gallons. Thus, at a constant electrical conductivity, the operating

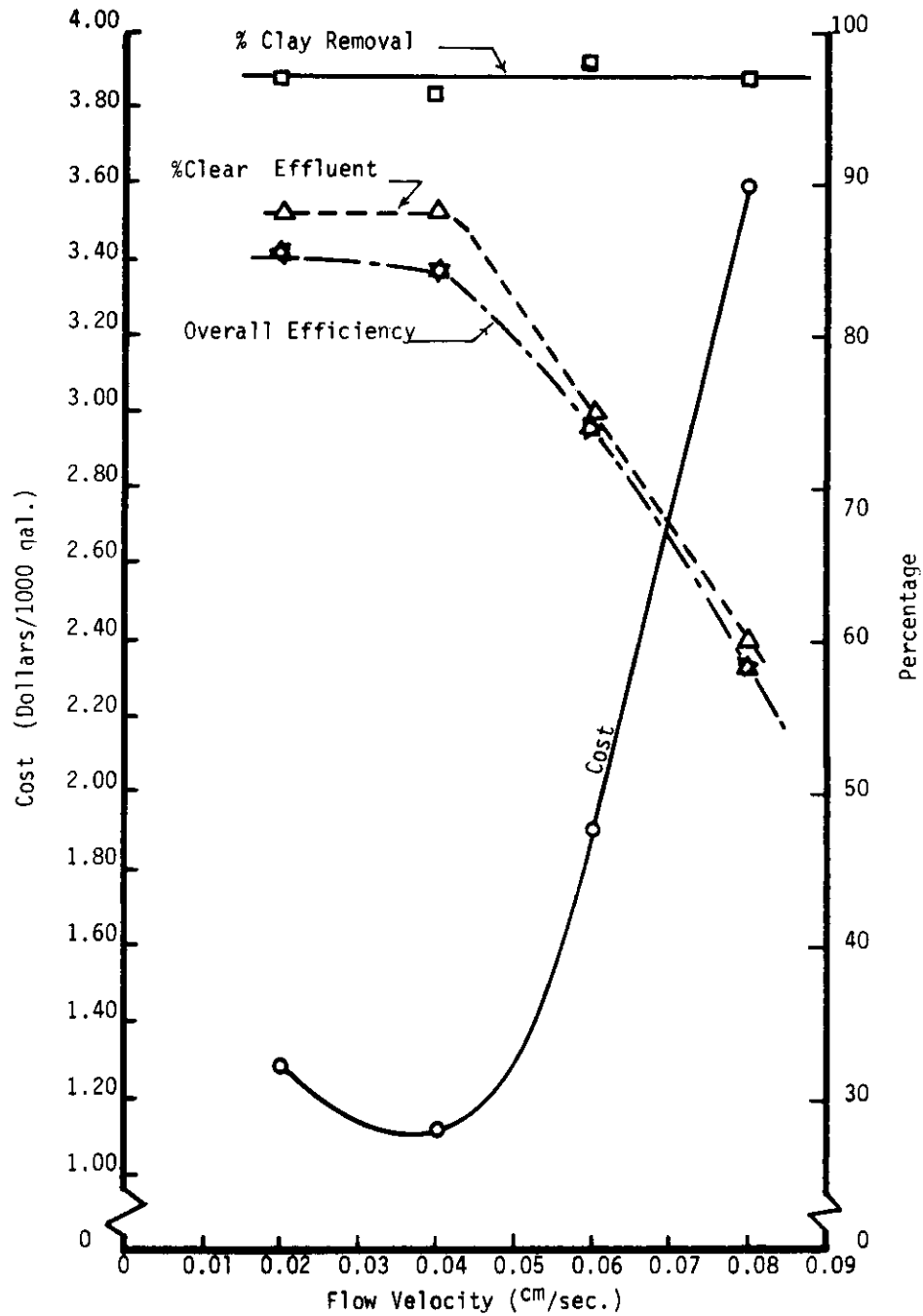


Figure 23. Limiting flow velocity evaluation for parallel plate model at optimum field strength.

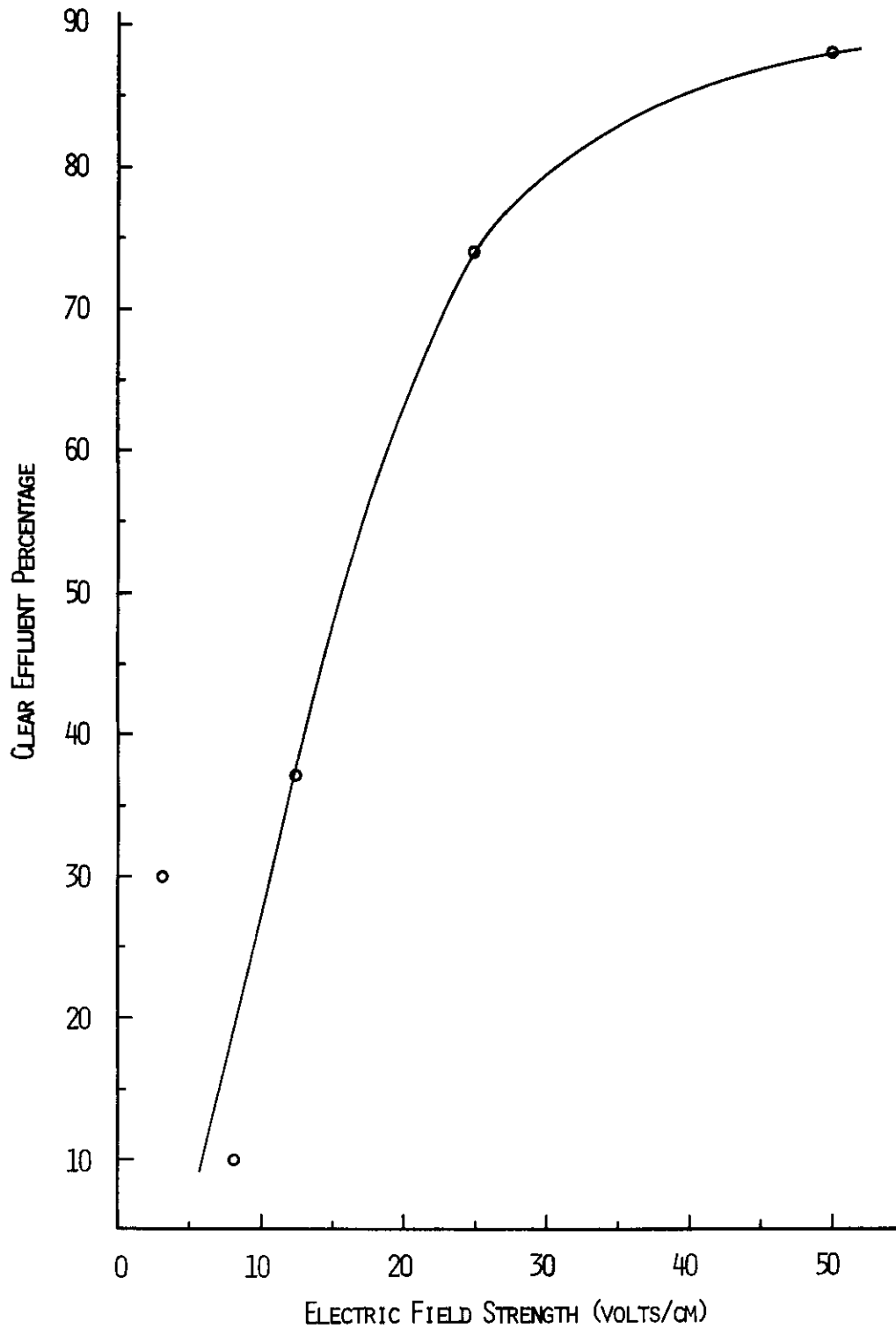


Figure 24. Clear effluent percentage versus field strength at flow rate of 440 ml min^{-1} - parallel plate model.

cost was roughly proportional to the rate at which clear effluent is produced as shown in Figure 25. This includes data from the entire range of flow rates. For example, operation at a flow rate of 220 ml min⁻¹ and at 1/6 the optimum field strength (4.15 volts cm⁻¹) yielded 96 percent clay removal and 50 percent clear effluent. This resulted in a clear effluent production rate of 1.7 gallons hr⁻¹ at a cost of \$0.07 per 1,000 gallons. The scatter in the data is caused primarily by electrical conductivity variations experienced in the test although these variations were small. The curve of Figure 25 was fitted by curvilinear regression. The regression of cost on clear flow rate was highly significant (0.01 level) and may be expressed as

$$\text{COST} = \$0.0445 (\text{GPH})^{1.97}$$

As a result of the temperature gradient an electric field maintained only in the top half of the model was found to be adequate and in fact superior to using all electrodes to keep the clay confined to the lower half of the model. This greatly reduced the power requirement and operating expense as shown in Table III. This comparison was again made using the optimum flow rate of 440 ml min⁻¹ but operating at 1/4 the optimum field strength.

The latter case is the most economical and actually increases the clear effluent production. The author feels that this is the optimum operating condition for the parallel plate type electrophoretic model.

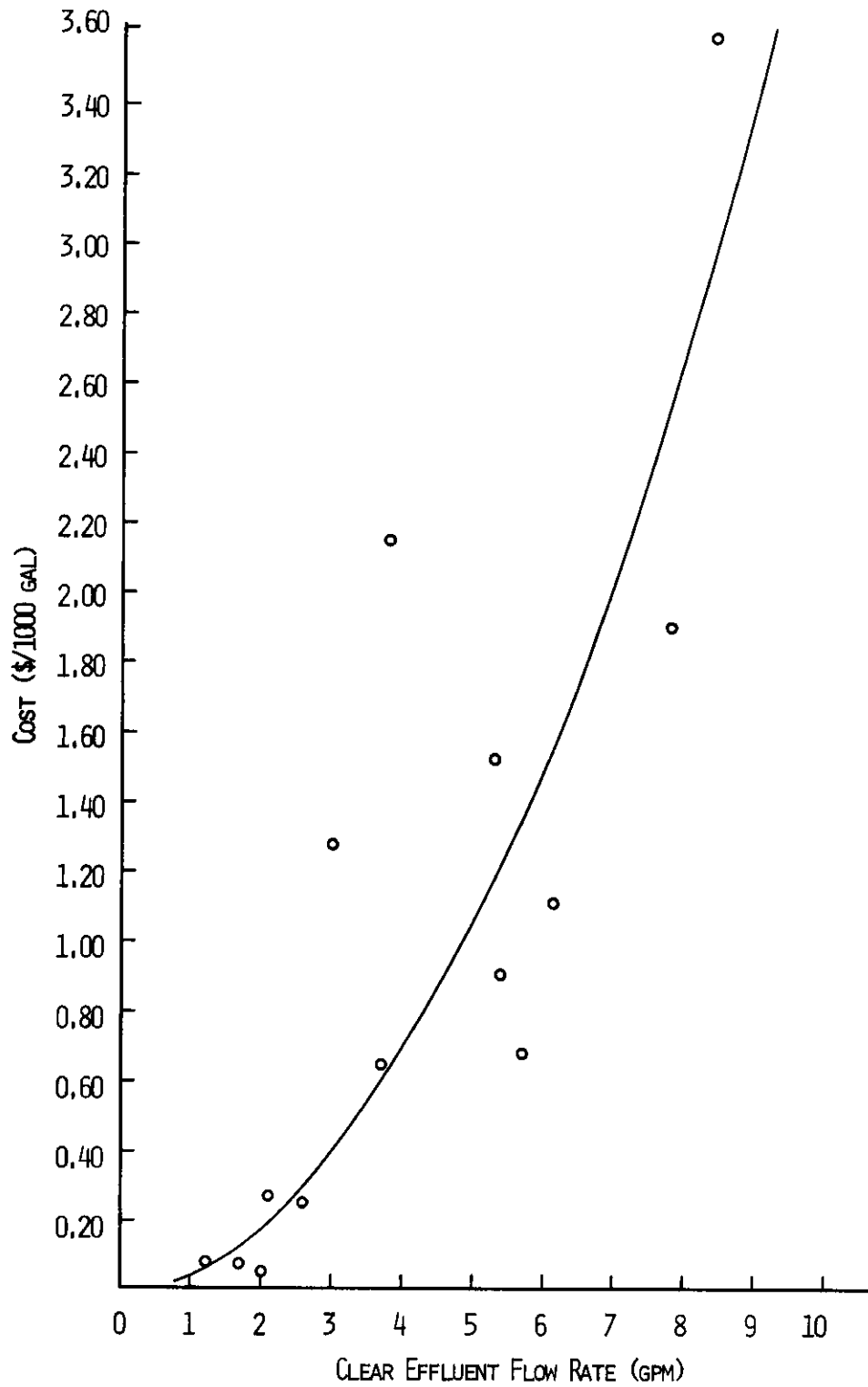


Figure 25. Power cost versus clear effluent production rate - parallel plate model.

TABLE III. Variation of Number and Location of Electrodes, Flow Rate 440 ml min^{-1}

Number and Location of Electrodes	%Clay Removal	% Clear Effluent	Clear Effluent Prod. Rate	Cost Per 1,000 Gal.
All	96	37	2.6 GPH	0.25
Top $\frac{1}{2}$	94	50	3.5 GPH	0.10

These experiments were all conducted with waters of very low electrical conductivities resulting in very slow rates of electrolysis. At higher conductivities the products of electrolysis would likely cause sufficient turbulence in this system to render the electrophoretic transport ineffective and thus appreciably disrupt the entire operation of the system.

Porous Filter Model

At low electrical conductivities the 5-micron filter proved to be slightly superior to the 165-micron filter in all considerations except plugging. In addition a flow rate of $224 \text{ ml min}^{-1} \text{ ft}^{-2}$ appeared to be preferable to the other two tested although each exhibited acceptable clay removal. Results of the flow rate and plate porosity tests are illustrated in Figures 26 through 28. These tests were run at very low electrical conductivities, ranging from 3 to 5 micromhos cm^{-1} .

One hundred percent turbidity removal was obtained with both

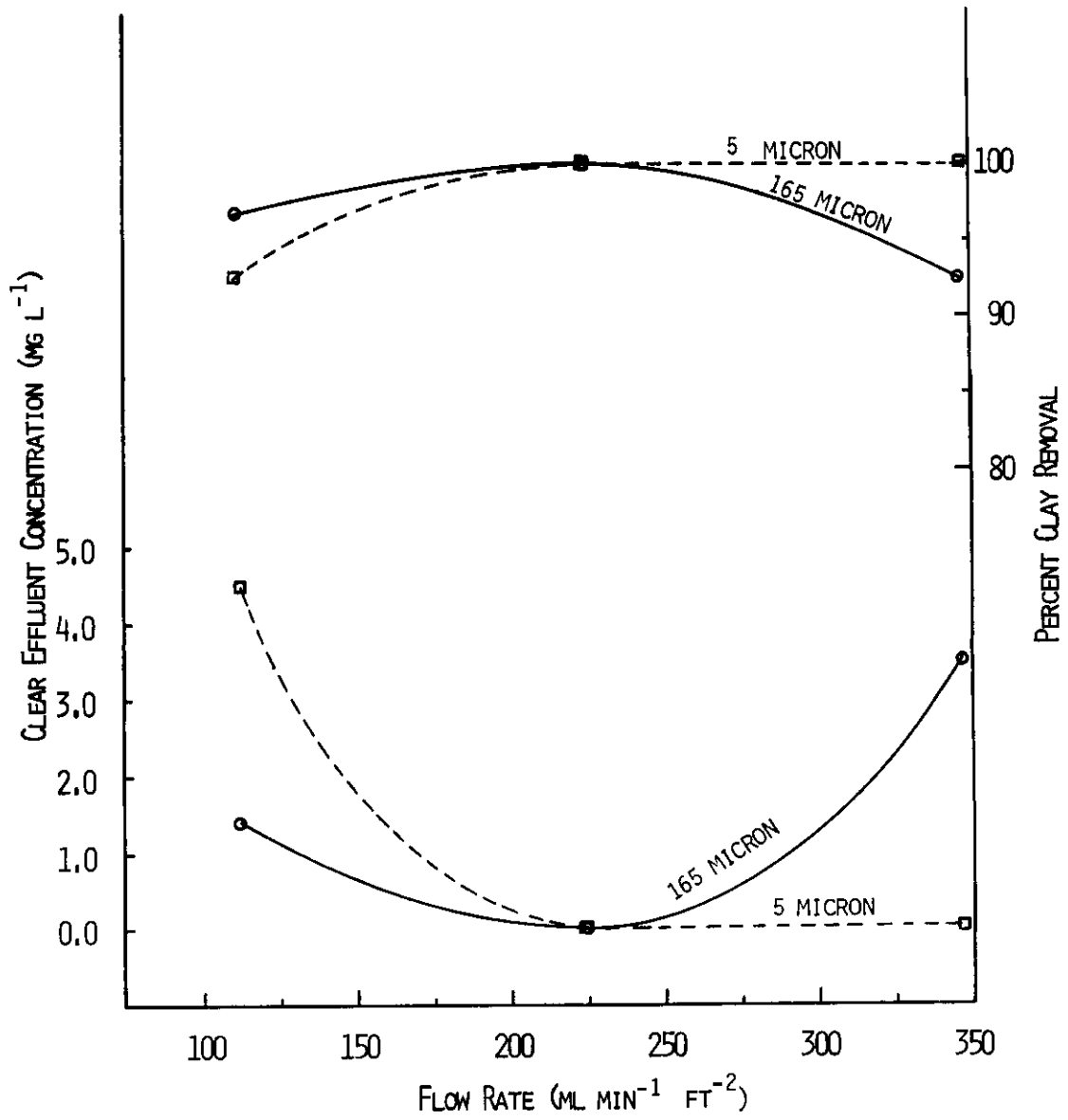


Figure 26. Evaluation of filter porosity on performance at various flow rates - porous filter model.

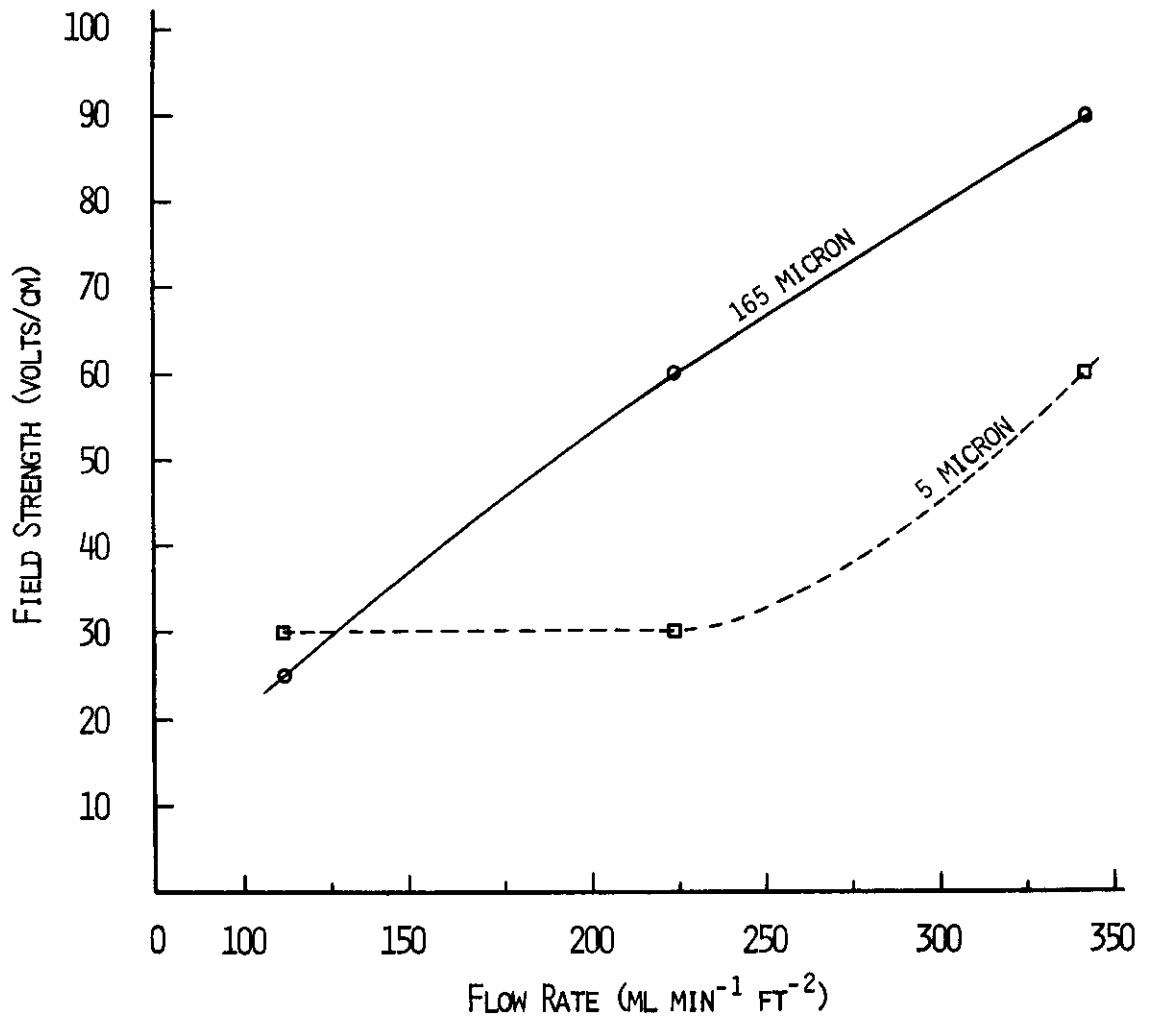


Figure 27. Field strength at maximum clay removal versus flow rate - porous filter model.

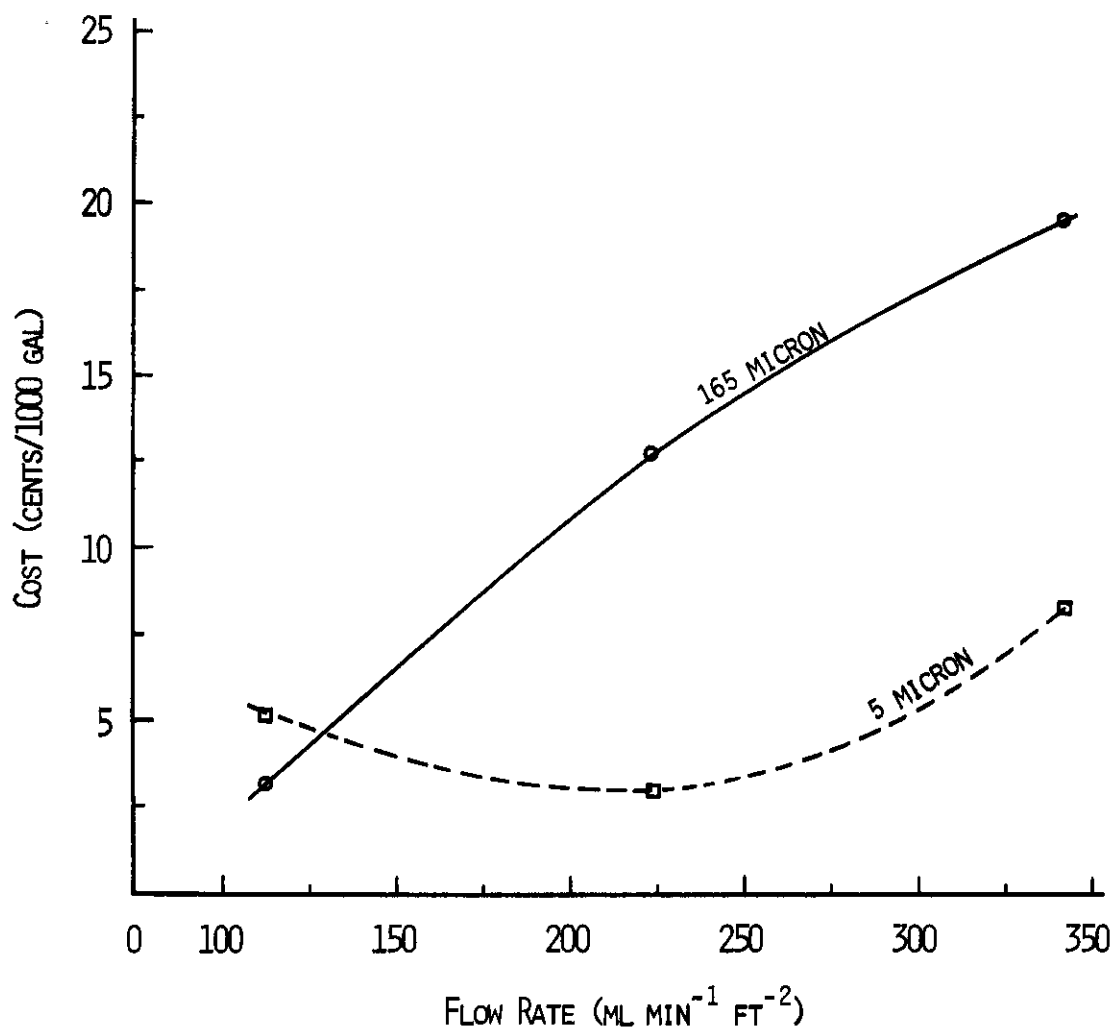


Figure 28. Power cost versus flow rate - porous filter model.

filter porosities at the $224 \text{ ml min}^{-1} \text{ ft}^{-2}$ flow rate. The 5-micron plate also yielded 100 percent removal at the $342 \text{ ml min}^{-1} \text{ ft}^{-2}$ flow rate. Clear effluent concentration and percent clay removal are both plotted versus flow rate in Figure 26.

The electric field strengths necessary to obtain the clarification indicated in Figure 26 are shown in Figure 27. Note that on the 165-micron filter the relationship of field strength to flow rate is almost a straight line and the field strength is approximately 3 times that which is considered optimum at each flow rate. Clarification was probably entirely due to electrophoresis with this porous plate. The minimum field strength at optimum clarification was generally lower for the 5-micron filter, indicating some filtration may have taken place. This is also suggested due to increased plugging of the less porous plate as determined by flow rate decrease. The average flow rate decrease due to deposition was approximately $6 \text{ ml min}^{-1} \text{ ft}^{-2} \text{ hr}^{-1}$ for the 5-micron plate as compared to about 1.5 for the 165-micron filter.

Operating costs are shown in Figure 28. This figure indicates that the 5-micron filter operating at $224 \text{ ml min}^{-1} \text{ ft}^{-2}$ was the most efficient and economical for these sets of tests.

Conductivity tests. Due to the initial tests a clear effluent flow rate of $224 \text{ ml min}^{-1} \text{ ft}^{-2}$ was used for evaluating influent electrical conductivity and clay concentration. It was found that at higher conductivities the plate porosity had little or no effect on

the results or operation of the system. Several porosities were used in these tests.

A high clay removal percentage was obtained at all influent electrical conductivities, amounting to 97 percent or greater. The highest clay concentration that remained in the clear effluent was only 1.4 mg l^{-1} . The minimum field strength which would maintain maximum clay removal decreased with increasing conductivity as is indicated in Figure 29. This decrease was due primarily to the electrochemical reactions that occurred in the model which are more dependent on current density than on electrical field strength. The primary clarifying action appeared to be from electrochemically induced flocculation which was then filtered by the porous stainless steel plates. This is supported by the fact that field strengths lower than optimum were adequate for clarification at the higher conductivities. In addition the flow rate decrease caused by a filter cake buildup of flocculated particles whose repelling charges had been neutralized was much greater at the higher conductivities. The flow decrease at the higher conductivities averaged more than $20 \text{ ml min}^{-1} \text{ ft}^{-2} \text{ hr}^{-1}$ compared to about 1.5 to 6 at the low conductivities.

Due to the above facts, the operating costs did not rise as fast with electrical conductivity as would be expected when using electrophoresis solely. If clear effluent only is considered usable, the cost per 1,000 gallons ranged from an average of \$0.08 at 4 micromhos

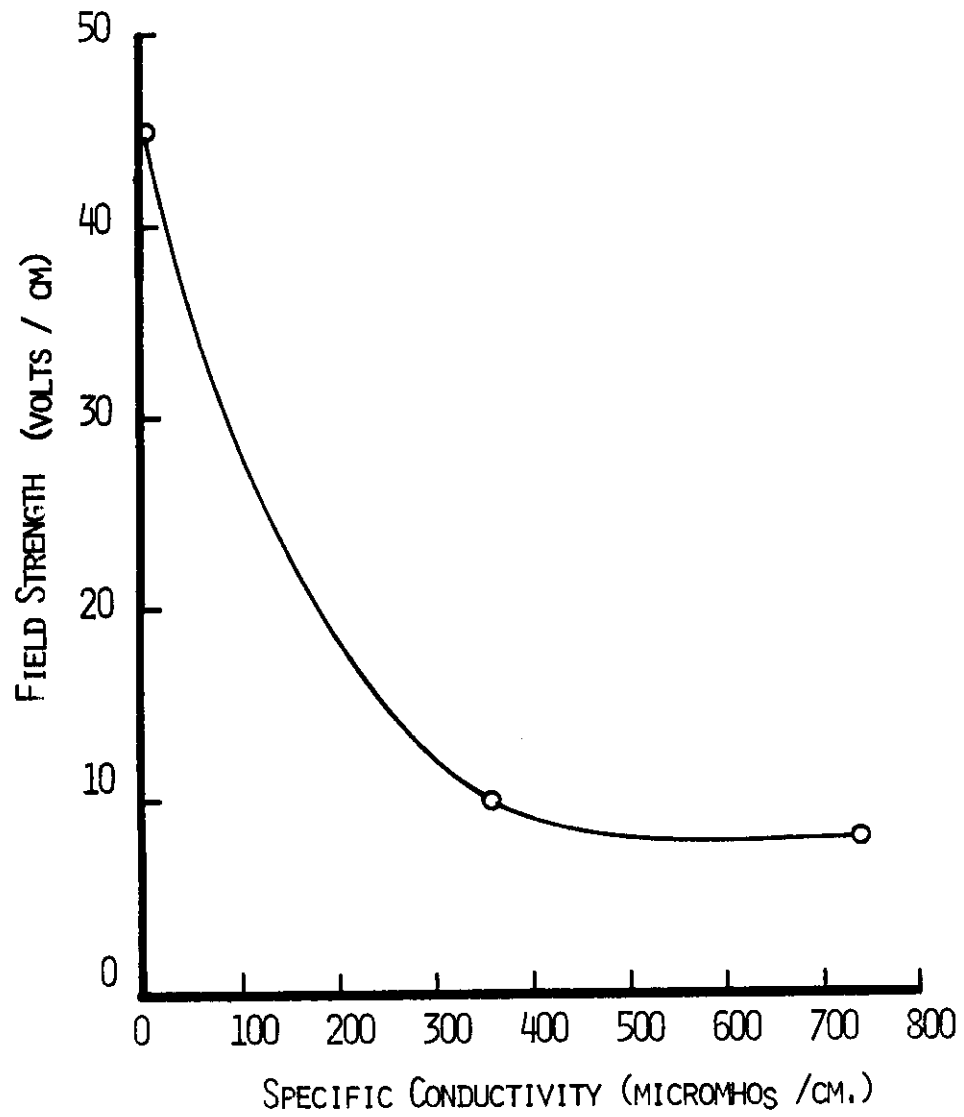


Figure 29. Field strength at maximum clay removal versus specific conductivity - porous filter model.

cm^{-1} to \$0.16 at 740 micromhos cm^{-1} . This relationship is shown in Figure 30.

The particulate matter in what was normally considered waste flow or turbid effluent was likewise highly flocculated. It was found that after a period of settling, an estimated 75 percent or more of this water was relatively free of turbidity. If it is assumed that 75 percent (a very conservative estimate) of the turbid effluent is recoverable for use, the cost then drops to below \$0.10 per 1,000 gallons for the entire conductivity range tested. This is shown in Figure 31.

Clay concentration tests. A high percentage of clay was removed with each influent concentration tested. One hundred percent removal was achieved at the low influent concentration of 3 $\text{mg} \cdot \text{l}^{-1}$. At 93 $\text{mg} \cdot \text{l}^{-1}$ influent concentration, 1.3 $\text{mg} \cdot \text{l}^{-1}$ remained in the purified effluent giving 98.6 percent removal. The electrochemical effect was also the predominant means of clarification in these tests, since the specific electrical conductivity ranged from 300 to 400 micromhos cm^{-1} . The flow rate decrease again averaged approximately 20 $\text{ml} \cdot \text{min}^{-1} \cdot \text{ft}^{-2} \cdot \text{hr}^{-1}$ for each concentration tested with exception of the highest concentration (93 $\text{mg} \cdot \text{l}^{-1}$), where the flow decrease climbed to 27 $\text{ml} \cdot \text{min}^{-1} \cdot \text{ft}^{-2} \cdot \text{hr}^{-1}$.

The minimum field strength required for maximum clay removal increased as influent clay concentration increased. Only 2 volts cm^{-1} was required for the low concentration. This increased to 16

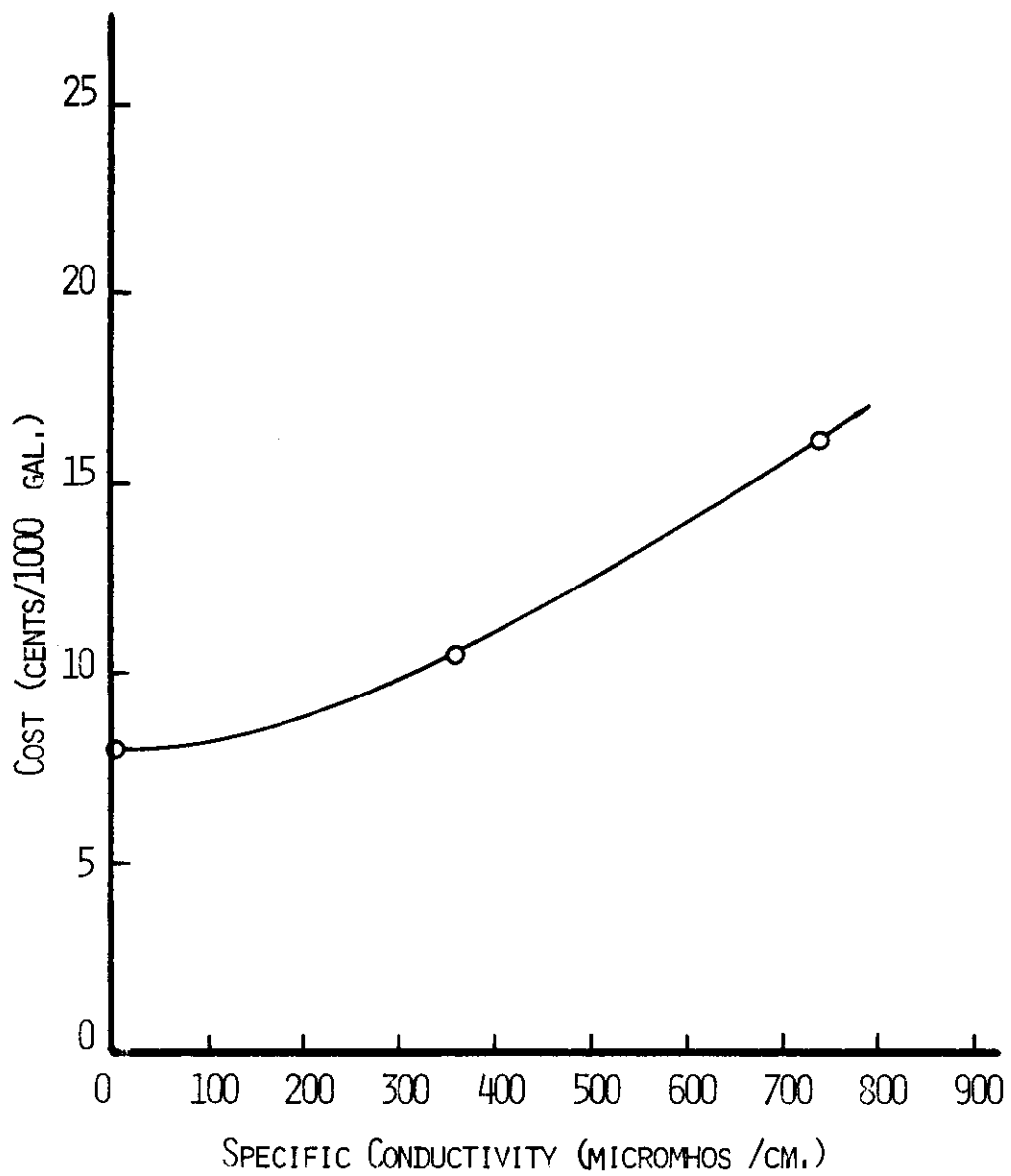


Figure 30. Operating costs versus specific conductivity considering clear effluent only - porous filter model.

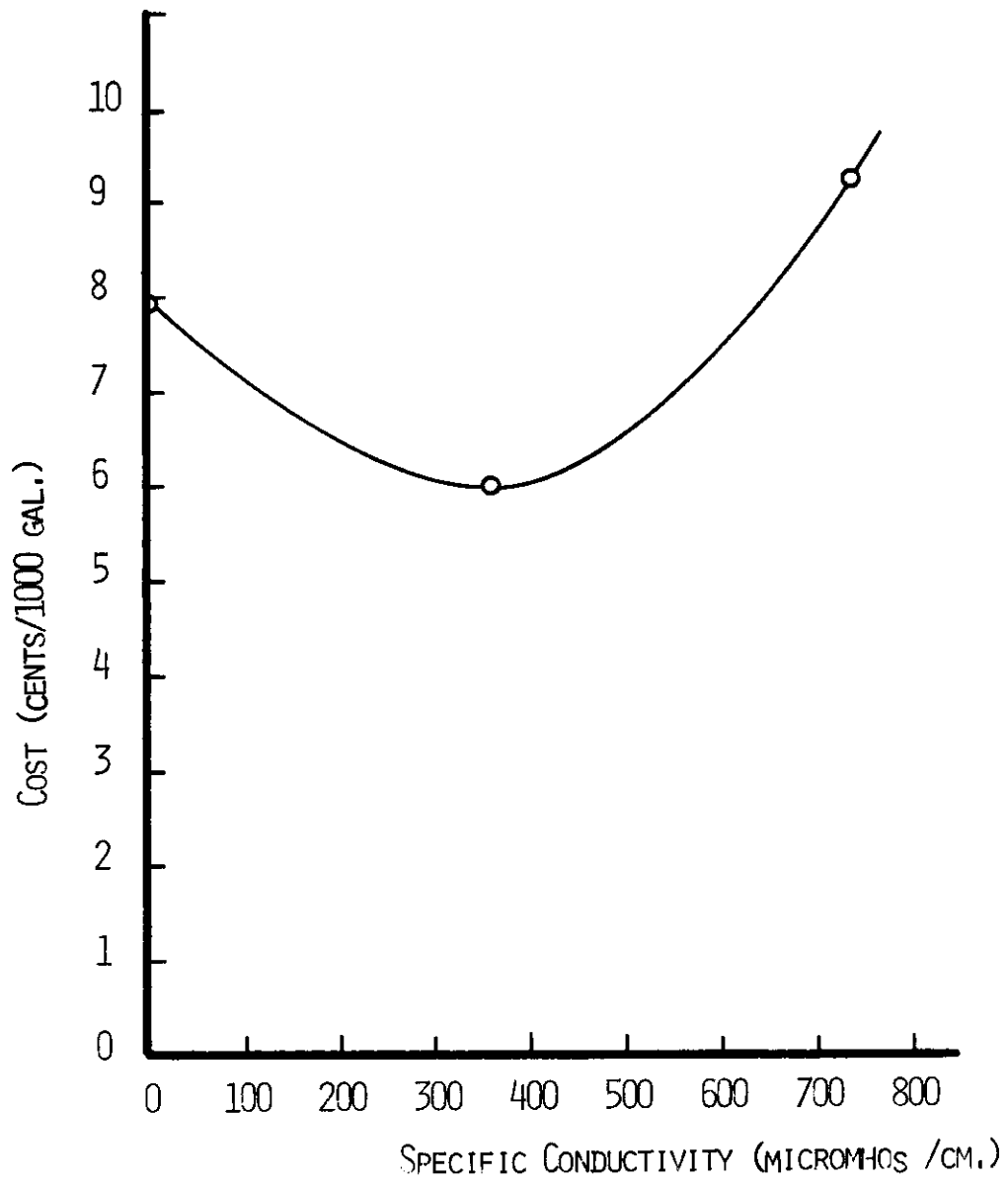


Figure 31. Operating costs versus specific conductivity, assuming 75% recovery of turbid effluent due to flocculation and settling - porous filter model.

volts cm^{-1} at 93 mg l^{-1} but still was only 0.8 times the optimum field strength. This is illustrated in Figure 32. This of course results in an increase in operating costs with concentration as shown in Figure 33 considering only clear effluent production.

For very low influent concentrations the operating cost is on the order of a few mils per 1,000 gallons and increases to \$0.38 per 1,000 gallons at the higher concentration. Again these costs are reduced substantially when the flow of turbid effluent after settling is included.

Electrode Grid Models

Stainless steel electrode tests. Tests with the stainless steel electrodes included three electrical conductivities. The average value for the low conductivity was $91.2 \text{ micromhos cm}^{-1}$; the medium conductivity averaged $341 \text{ micromhos cm}^{-1}$; and the high value averaged $716 \text{ micromhos cm}^{-1}$. Three influent clay concentrations were tested at each electrical conductivity and averaged 18.5 mg l^{-1} at the low concentration tested; 48.9 mg l^{-1} at the medium concentration; and 96.3 mg l^{-1} at the highest concentration. Total flow rate for the electrode grid tests was 330 ml min^{-1} or approximately 5.2 gal hr^{-1} . This flow was split at the electrode grid and left the model through two outlets.

The effluent concentration was found to increase in these tests with increasing influent concentration. In addition it

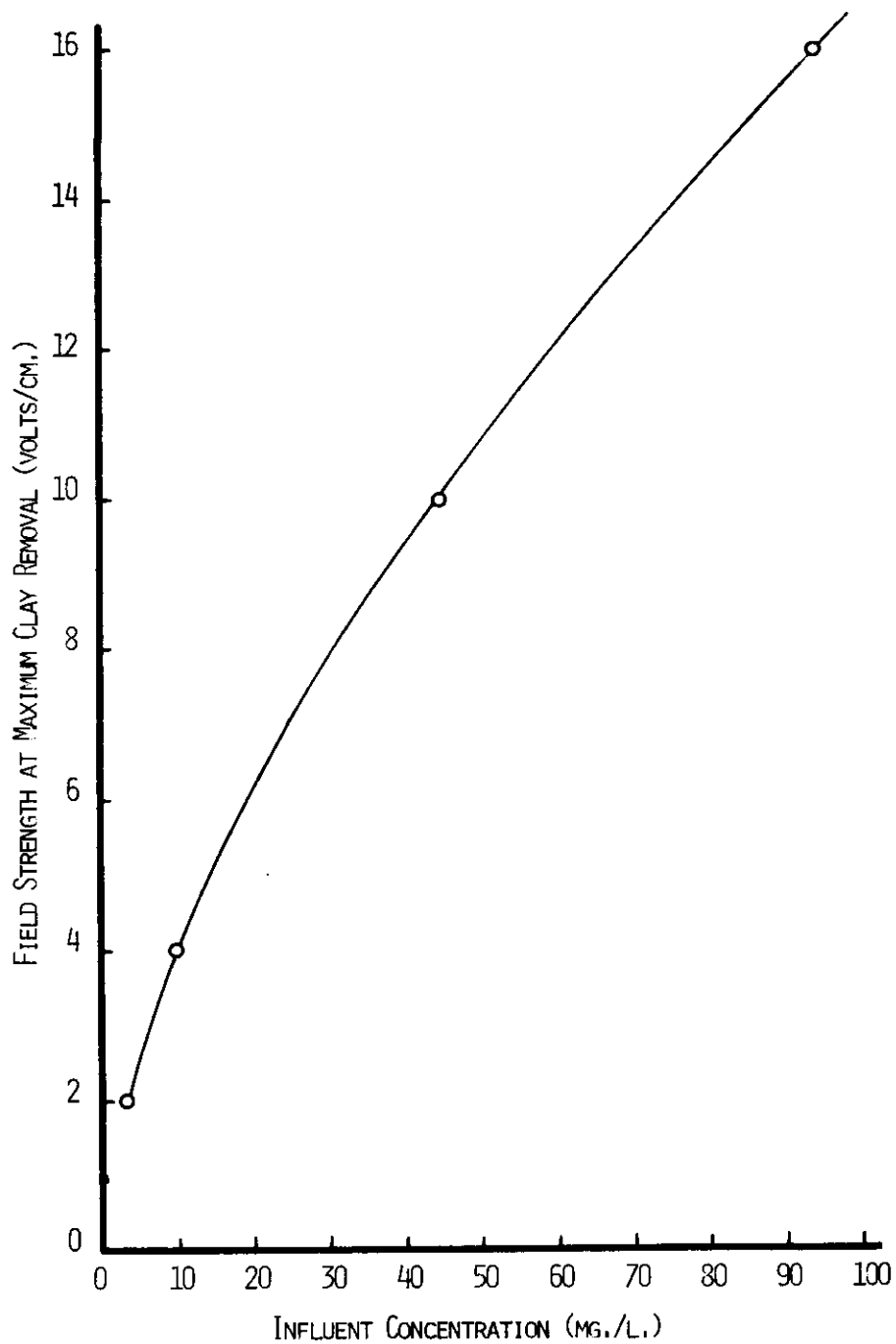


Figure 32. Field strength versus influent concentration - porous filter model.

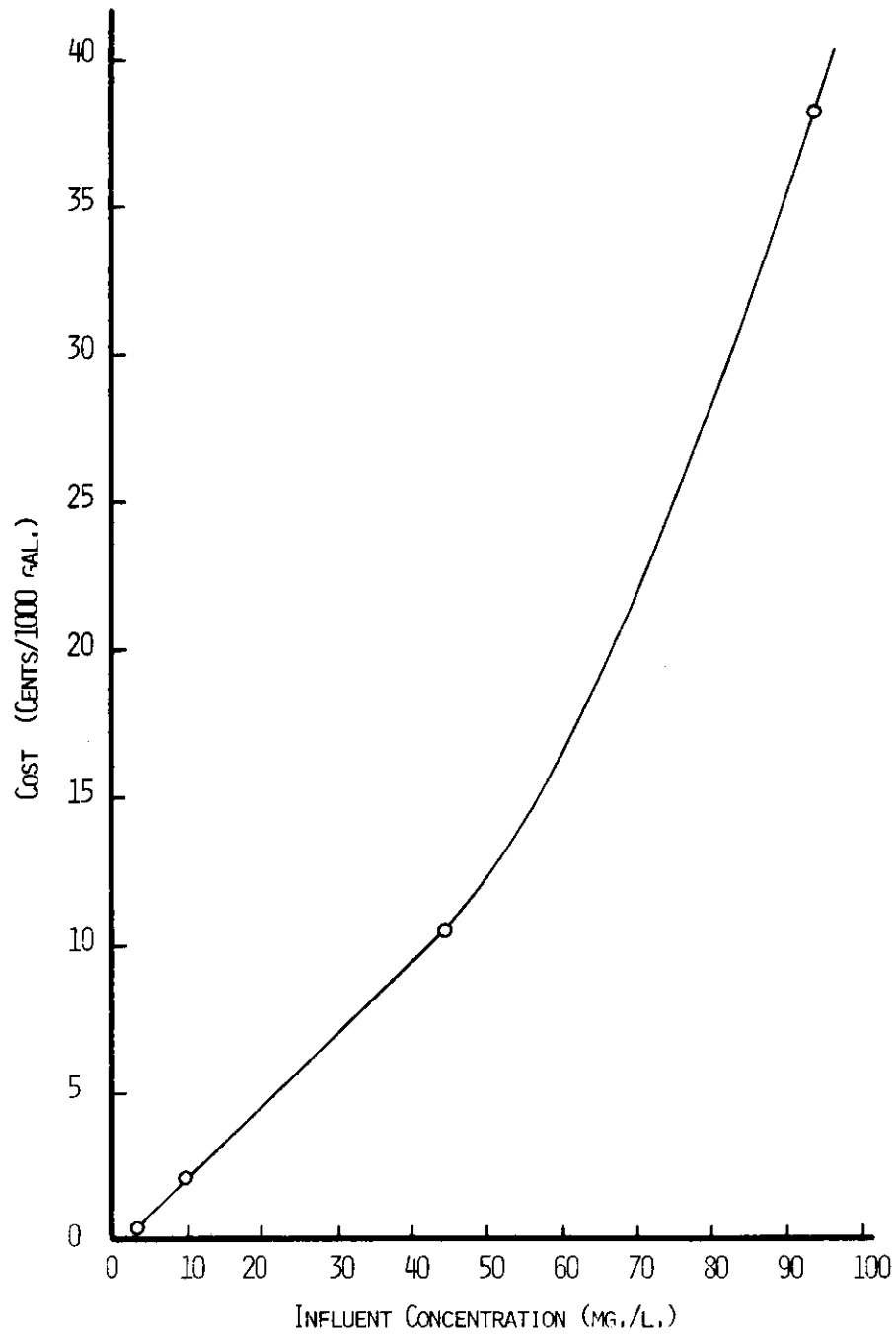


Figure 33. Operating cost versus influent concentration considering clear effluent only - porous filter model.

increased slightly with electrical conductivity at the low and high concentration but was relatively unaffected by conductivity at the medium concentration. These results are shown in Table IV. The effluent turbidity for the low and medium influent concentrations are very close to or below the 10 mg l^{-1} that is required by the U.S. Public Health Service. The effluent produced from the high concentration was slightly above this.

TABLE IV. Effluent Concentration, mg l^{-1} - Electrode Grid Model, Stainless Steel Electrodes

Average Influent Concentration	Specific Electrical Conductivity		
	$91.2 \mu\text{mhos cm}^{-1}$	$341 \mu\text{mhos cm}^{-1}$	$716 \mu\text{mhos cm}^{-1}$
18.5 mg l^{-1}	3.75	4.90	6.25
48.9 mg l^{-1}	10.90	8.90	9.90
96.3 mg. l^{-1}	11.90	12.40	13.90

Even though the effluent turbidity increased with influent concentration, the clay removal efficiency also increased as concentration increased (Table V).

The power costs for these tests may be adjusted so as to represent those which would be obtained when using optimum electrode grid dimensions. Referring to Chapter III, at the optimum design the average potential between the grids is 94 percent of that applied to the grid. Operating at the actual test configuration the average

potential available was 81.5 percent of that applied to the grids. Therefore, in maintaining the same current the cost at the optimum design may be lowered by a factor of $0.815/0.94 = 0.867$ or should amount to 86.7 percent of the actual test costs.

TABLE V. Clay Removal Efficiency, Percent - Electrode Grid Model, Stainless Steel Electrodes

Average Influent Concentration	Specific Electrical Conductivity		
	91.2 $\mu\text{mhos cm}^{-1}$	341 $\mu\text{mhos cm}^{-1}$	716 $\mu\text{mhos cm}^{-1}$
18.5 mg. l^{-1}	72.5	79.4	64.7
48.9 mg. l^{-1}	78.0	81.8	79.4
96.3 mg. l^{-1}	87.9	87.4	84.9

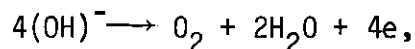
Table VI represents the adjusted power costs per 1000 gallons of effluent produced for the stainless steel tests. These are the costs of obtaining the effluent concentrations shown in Table IV. It is obvious as well as logical that the cost decreased with increasing electrical conductivity. At the higher conductivity the potential required is less to maintain an adequate current for the necessary electrochemical reaction rate. The power costs for waters with an average specific electrical conductivity of 341 micromhos cm^{-1} and above were less than \$0.20 per 1000 gallons in all cases and the majority of natural waters treated should fall within this range.

TABLE VI. Power Costs Per 1000 Gallons - Electrode Grid Model, Stainless Steel Electrodes

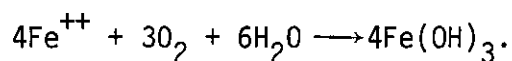
Average Influent Concentration	Specific Electrical Conductivity		
	91.2 $\mu\text{mhos cm}^{-1}$	341 $\mu\text{mhos cm}^{-1}$	716 $\mu\text{mhos cm}^{-1}$
18.5 mg l^{-1}	30.3¢	11.4¢	6.5¢
48.9 mg. l^{-1}	21.8¢	17.8¢	9.8¢
96.3 mg. l^{-1}	58.5¢	16.8¢	9.6¢

The minimum current which would maintain adequate flocculation consistently fell within the range of three to four amperes which corresponds to current densities of 8.33×10^{-4} to 1.11×10^{-3} amp cm^{-2} .

Aluminum electrode tests. The reddish brown precipitate and flocculating agent appearing in the stainless steel electrode grid tests, as well as porous filter tests, was originally thought to be ferric hydroxide which was being formed by the oxidization of the ferrous ion Fe^{++} in the raw water. The oxidizing agent was thought to be oxygen produced at the anode by electrolysis, the reaction being



and the ferric hydroxide forming as follows:



The stainless steel electrodes were considered to be largely passive

due to their corrosion resistance as well as from the formation of a protective oxide film at the current densities used in the test.

In order to test this hypothesis, additional tests were run using aluminum as the electrode grid material. If the above were correct, the reddish brown ferric hydroxide would likewise appear in the aluminum electrode tests. This failed to occur, however, indicating the ferric ion was derived primarily from the dissolution of the stainless steel electrodes where it then hydrolyzed to form ferric hydroxide.

The aluminum electrodes consistently gave much better results than did the stainless steel electrodes. All of the aluminum electrodes were tested at a medium electrical conductivity that averaged 330 micromhos cm^{-1} . The three influent concentrations tested were 18.2 mg l^{-1} , 56.8 mg l^{-1} , and 105 mg l^{-1} .

The results of these tests are shown in Figures 34 through 36 and are compared with the stainless steel electrodes at the same conductivity. Much lower effluent clay concentrations were obtained with aluminum as is shown in Figure 34. At the high influent concentration of 105 mg l^{-1} the effluent concentration was 8.6 mg l^{-1} which resulted in a 91.8 percent removal efficiency. The clay removal efficiencies are given in Figure 35 and are likewise considerably higher than those obtained with stainless steel.

The adjusted power costs using aluminum electrodes were less than \$0.10 per 1000 gallons for each concentration tested. For the medium

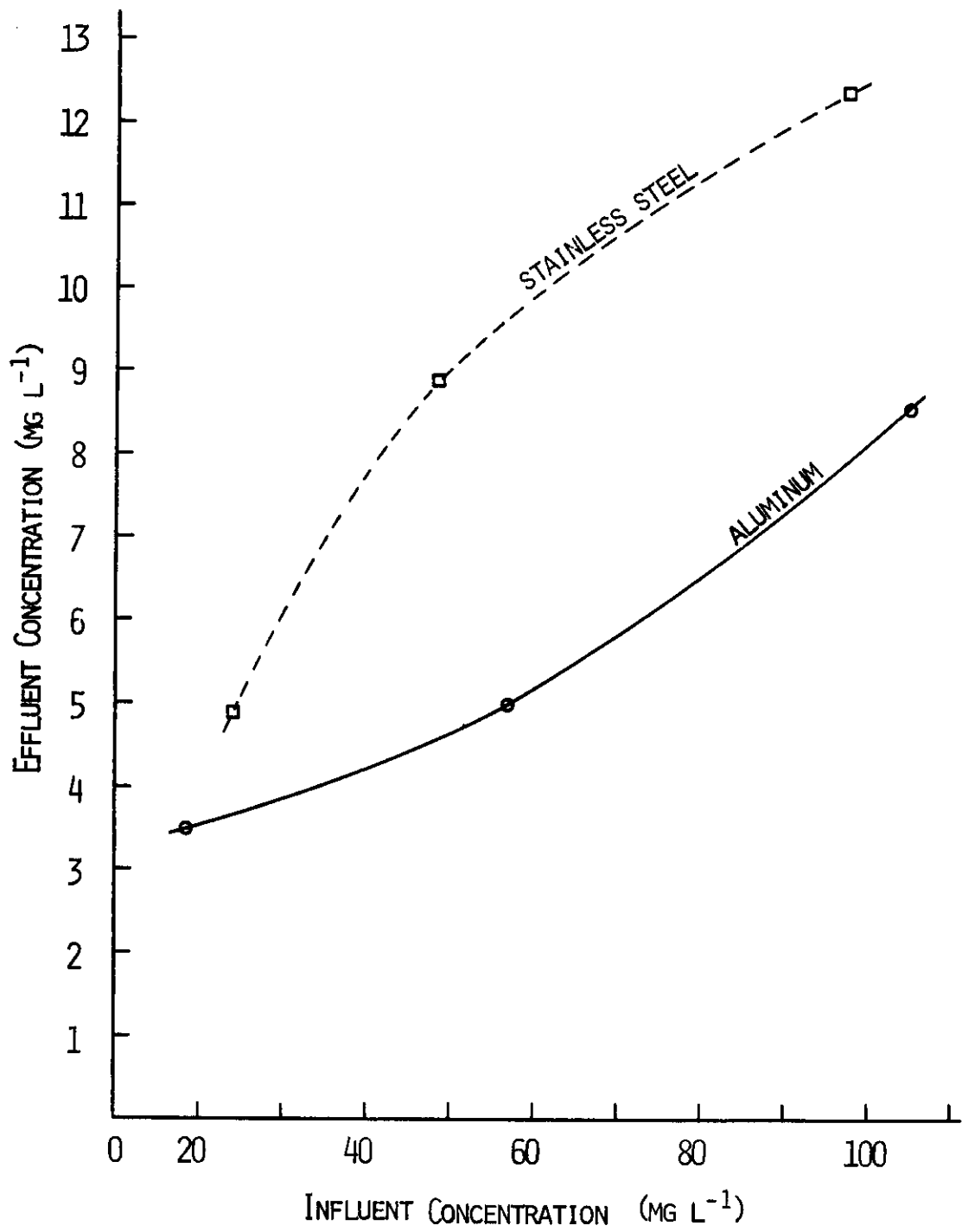


Figure 34. Effluent concentration versus influent concentration-electrode grid model.

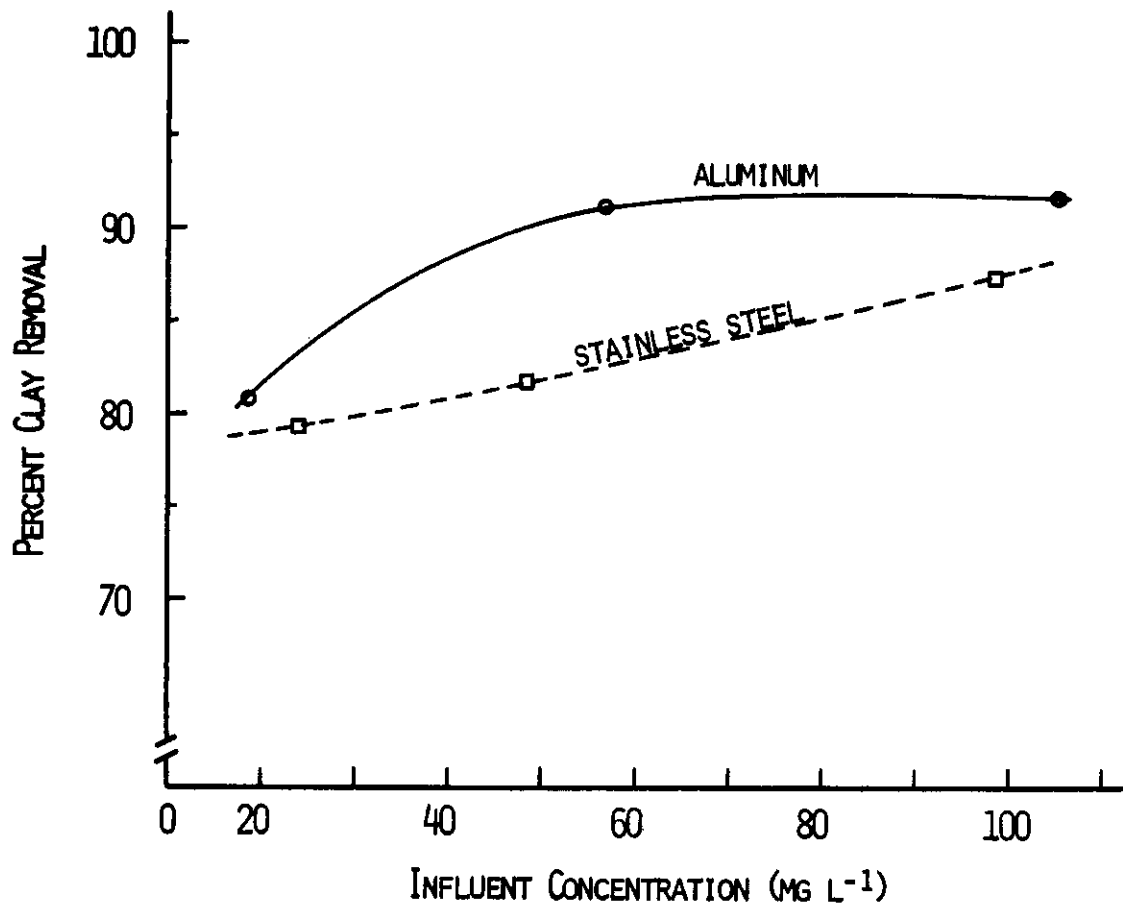


Figure 35. Percent clay removal versus influent concentration-electrode grid model.

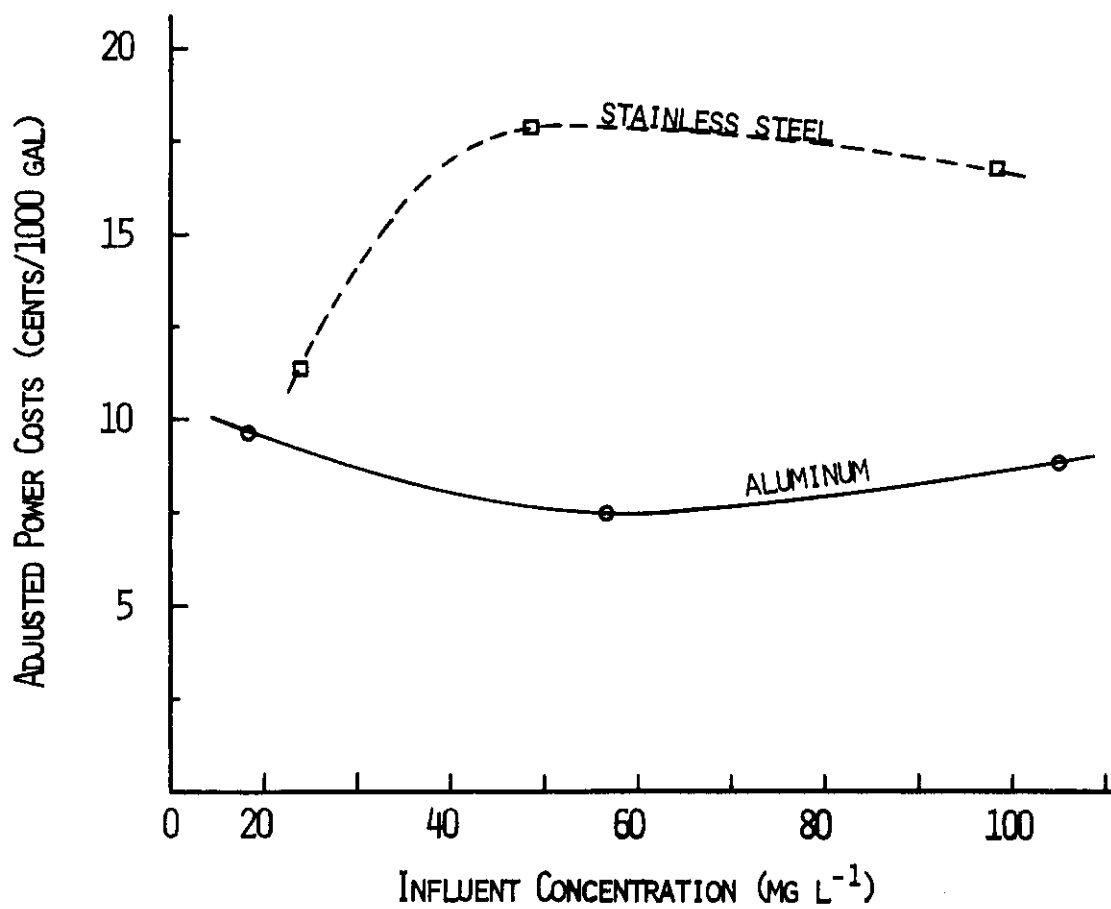


Figure 36. Adjusted power costs versus influent concentration - electrode grid model.

and high concentration these power costs were approximately one-half those of the stainless steel as is illustrated in Figure 36.

In each aluminum test the minimum current which would maintain flocculation at the flow rate of about $5.2 \text{ gallons hr}^{-1}$ was approximately 3 amperes. This gave a current density of $8.3 \times 10^{-4} \text{ amp cm}^{-2}$.

Bacteria Determinations

Tests were made for coliform bacteria on water sampled as the effluent left the flocculator and also after settling. The addition of bacteria to the influent suspension is described in the first section of this chapter. These tests were possible only with suspensions of low conductivities. The tap water when mixed in higher proportions with deionized water contained sufficient residual chlorine to kill the bacteria before they reached the model. These tests were conducted only with the electrode grid model and primarily with the stainless steel electrodes. One determination only was made with the aluminum plates.

Complete bacteria removal was achieved in each test. There were no indications of coliform bacteria present after settling or after the flocculation operation in the stainless steel tests. An MPN of 3.6 coliform per 100 ml remained after flocculation in the aluminum test but none remained after settling.

There were three possible ways in which bacteria could be removed in this system. The first involved the migration of the

negatively charged bacteria to the anode where it would adhere to the anode along with yet unflocculated clay particles. When the polarity reversal occurred this deposited material was usually discharged as large flocs or as consolidated plate like material which rapidly sank to the bottom of the model. The second possibility was through combining with flocs formed by gentle turbulent mixing before and after reaching the electrodes. The bacteria which combined with flocs in either of these two ways should have been removed either in the model or later in the settling basins. The third possibility was by the oxidation and killing of bacteria by chlorine which was possibly being discharged at the anode.

As a consequence of the above observations, residual chlorine measurements were made on a number of remaining tests. Total cell potentials of from 5 to 40 volts were tested with the aluminum electrodes with current densities ranging from 5.6×10^{-4} to 1.39×10^{-3} amps cm^{-2} . Total chlorine residuals greater than 0.05 mg l^{-1} were not detected in any of these tests.

Similar residual chlorine tests were made with stainless steel electrodes. In these tests appreciable residual chlorine was detected when the total potential between electrodes reached approximately 15 volts. Below this very little residual chlorine could be measured. The results of the chlorine tests are presented in Table VII.

The minimum potential applied for adequate clarification in the stainless steel tests in which bacteria determinations were

TABLE VII. Residual Chlorine Determinations

Potential Between Electrodes (volts)	Current Density (amp cm ⁻²)	Total Chlorine Residual (mg l ⁻¹)	Free Chlorine Residual (mg l ⁻¹)	Combined Chlorine Residual (mg l ⁻¹)
<u>Aluminum Electrodes</u>				
5.5	5.5 x 10 ⁻⁴	<0.05	—	—
9.0	8.3 x 10 ⁻⁴	0.00	—	—
9.0	1.1 x 10 ⁻³	<0.05	—	—
10.0	8.3 x 10 ⁻⁴	0.00	—	—
12.5	1.4 x 10 ⁻³	0.00	—	—
40.0	8.3 x 10 ⁻⁴	<0.05	—	—
<u>Stainless Steel Electrodes</u>				
8.0	5.5 x 10 ⁻⁴	<0.05	—	—
10.0	8.3 x 10 ⁻⁴	<0.05	—	—
10.5	8.3 x 10 ⁻⁴	<0.05	—	—
13.5	2.8 x 10 ⁻⁴	<0.05	—	—
14.2	1.1 x 10 ⁻⁴	<0.05	—	—
15.5	1.4 x 10 ⁻³	1.80	1.20	0.60
15.75	3.4 x 10 ⁻⁴	0.80	0.50	0.30
16.0	1.4 x 10 ⁻³	2.30	2.00	0.30
25.0	5.5 x 10 ⁻⁴	1.85	1.30	0.55
41.0	1.1 x 10 ⁻³	2.30	1.45	0.85
54.0	1.7 x 10 ⁻³	1.80	1.30	0.50

made was in each case well above 15 volts. Therefore, the bacteria removal can be attributed to the chlorine evolution in these tests. However, the first two possibilities must be given credit for the bacteria removal in the aluminum test.

Electrophoretic Characterization of Soils

When electrophoresis is the sole method of clarification the electrophoretic transport velocity is usually balanced against the hydraulic flow velocity in one way or another. This was necessary in the parallel plate model as well as the porous filter model to obtain optimum results at low conductivities. It is therefore necessary to know the electrophoretic mobility of the pollutants which are to be removed from the water.

Seven Texas soils, representing several different areas of the State as well as a wide range of chemical and physical characteristics, were analyzed to determine their electrophoretic mobility. Only the clay fraction of the soil was tested and included all of the soil particles of two micron size and smaller. The clay mineral types as determined by X-ray diffraction were available for the soils tested and are included with the mobility values. In addition, the percent clay as was determined by the pipette method is given. These data are presented in Table VIII.

At the onset of this reported research project it was hoped that electrophoretic or electrochemical methods would be well adapted to

TABLE VIII. Electrophoretic and Mineralogical
Characterization of Texas Soils

Soil Type	% Clay	Clay Mineral Types *	EM @ 25°C ($\mu \text{ sec}^{-1} \text{ volt}^{-1} \text{ cm}$)
Lufkin fs1	13.1	KQ	2.04
Lufkin clay	40.3	MKQ	2.20
Houston Black clay	55.5	MKQ	1.23
Victoria clay	40.4	MKQI	2.07
Hoban scl	29.8	IKQF	1.47
Amarillo fs1	11.3	IKQF	2.04
Amarillo scl	21.6	IKQF	2.13

* Legend: M - Montmorillonite, K - Kaolinite, I - Illite,
Q - Quartz, F - Feldspars, listed in order of prominence.

clarification of playa lake water in the Texas High Plains prior to ground water recharge. As a result, sediment samples were taken from playa lakes throughout this area from locations which included the three major soil types of the area. The soil within the lakes are, however, all classified as Randall clay despite the differences in soil types of the contributing adjacent soils. The electrophoretic mobilities of these soils are given in Table IX along with X-ray diffraction data and percent clay.

The high electrophoretic mobilities obtained indicate that waters containing these soil sediments could very readily be clarified

TABLE IX. Electrophoretic and Mineralogical Characterization of Playa Lake Sediments From the Texas High Plains

<u>Location</u>	<u>% Clay</u>	<u>Clay Mineral Type, <2μ</u>	<u>EM @ 25$^{\circ}$C</u>
Amarillo	43.5	I ₁ M ₂ K ₂ Q ₂ *	2.88
Bushland	42.3	I ₁ M ₂ K ₂ Q ₂	3.47
Abernathy	46.5	I ₁ M ₂ K ₂ Q ₂	3.45
Lubbock	61.8	I ₁ M ₂ K ₂ Q ₂	2.61
Lamesa	61.8	I ₁ M ₂ K ₂ Q ₃	2.67

*Legend: M - Montmorillonite, K - Kaolinite, I - Illite,

Q - Quartz; 1 - 40%, 2-10 to 40%, 3 - 10%.

by electrophoretic means. However, experimental results subsequent to this have virtually ruled out this method of clarification for recharge on a practical basis. Treated effluent flow rates are much too low and the cost per unit of water treated would be prohibitive for recharge purposes. However, utilizing the electrophoretic technique for domestic water supplies from these sources is definitely a practical possibility.

CHAPTER V

DISCUSSION OF RESULTS

The purpose of this reported research project was to investigate the development of a water purification system which utilized electrophoretic or electrochemical processes and was both operationally and economically practical. A small domestic water system was upmost in mind although the possibility of application to municipal, industrial, and other specialized water treatment systems was not discounted. The following evaluations and discussion of the tested models are offered with this in mind.

Discussion of Models

Parallel Plate Model

The parallel plate model was designed to utilize electrophoresis entirely for the removal of suspended solids. Waters with very low electrical conductivities met this requirement and when treated with this model were successfully clarified. Operationally, at these conductivities, the system worked well and could have been automated easily to accomplish the flushing cycle which periodically would be needed. In addition, power costs, for these specific tests could be brought down to acceptable values by accepting fairly low rates of purified effluent production. The power cost was approximately \$0.10 per 1000 gallons when using the top electrodes only and a clear

effluent production rate of 3.5 GPH, which was about the optimum operating conditions found for this model. The approximate cost of treating surface waters for Texas cities is \$0.15 per 1000 gallons which includes both capital costs and operating costs.* This cost of course does not include distribution facilities.

At this point reality must be faced in that waters of very low electrical conductivities rarely if ever exist in nature. In a system in which the electrodes are exposed to the water, appreciable electrolysis takes place as the electrical conductivity of the water increases. The electrolysis products cause sufficient bubbles and turbulence in the system to appreciably disrupt the electrophoretic transport process. In addition, many of the products of electrolysis join in primary and secondary chemical reactions which alter the characteristics of the suspension. Products formed due to these chemical reactions included coagulants which resulted in corresponding floc formation. The horizontal parallel plate model did not facilitate their removal.

Porous Filter Model

The porous filter model was designed to facilitate clarification either by electrophoresis alone or by electrochemically induced flocculation or both simultaneously. The test results in which the electrical conductivity was very low indicate performance due to electrophoresis alone. Operationally the model worked very well at

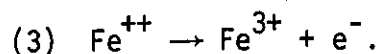
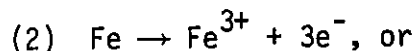
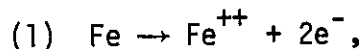
*Personal communication with Dr. T. D. Reynolds, Environmental Engineering Department, Texas A&M University.

this conductivity, removing 100 percent of the suspended clay (at a flow rate of 224 ml min^{-1}) and without an appreciable tendency to plug the porous stainless steel filter. In addition the average power costs at 224 ml min^{-1} was below \$0.10 per 1000 gallons. It is possible that this cost could be lowered with a better design concerning the flow regime or flow entering and passing through the filter. Part of the excess electric field (excess over theoretical optimum) was no doubt necessary to overcome some convective as well as local acceleration forces which could probably be decreased through better design. However, it must again be recognized that this situation, i.e., low electrical conductivity, would rarely if ever occur naturally.

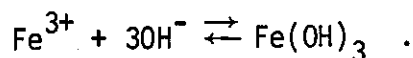
When operating at electrical conductivities high enough to result in considerable electrolysis rates, four and perhaps five unit operations in a conventional water treatment process are incorporated into one unit with this model. These include the addition of chemical coagulants, mixing, floc formation, filtration, and depending on the potential between the plates and the current density, possibly disinfection.

The coagulant present in this model was ferric hydroxide derived primarily from the dissolution of iron from the stainless steel electrodes (as determined in the electrode grid model). The initial dissolution was possible at two degrees of oxidation or from ions at both valency states. The following three possibilities exist

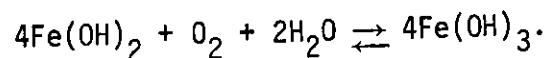
for dissolution:



The ferric ion produced in reactions (2) and (3) hydrolyzes directly to form ferric hydroxide as follows:



In a like manner ferrous hydroxide is secondarily formed from the primary reaction (1) where in the presence of oxygen it is oxidized to form ferric hydroxide:



The presence of oxygen and/or chlorine is very likely at the anode since the potential was greater than the discharge potential of either ion. It is possible for the metal to pass into solution in this case with a current efficiency of less than one, the remaining current discharging OH^{-} or other anions. If chlorine is being produced rather than or along with oxygen it will likewise oxidize the ferrous ion to the ferric state to form ferric hydroxide.

The mixing and flocculating operation was achieved by the turbulence caused by rising gaseous electrolysis products, O_2 , H_2 , and possibly Cl_2 . The filtration was of course by the stainless steel porous filter. Chlorine residual determinations were not made during the porous filter tests; however, its potential production was proven in the electrode grid tests.

The porous filter model worked well from several aspects. First, very good clarification was achieved at all conductivities and influent concentrations tested. This included clay removal almost entirely by electrophoresis on one hand to complete flocculation and filtration on the other. Economically, the direct power cost did not vary as much with electrical conductivity as it did with influent clay concentration. Cost was not prohibitively high in any test and had the potential of being much lower due to the settling of the flocculated turbid effluent.

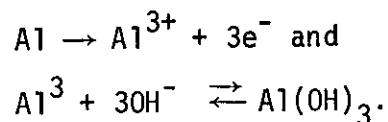
Operationally, however, the model was not completely successful. It was designed such that when operating electrophoretically the negatively charged porous cathode would repel most particulate matter in the system. The filter porosity was large enough in each case for clay size particles to pass through if they were not repelled. When operating electrochemically, however, many of the flocs which formed were drawn into the filter instead of settling to be removed by the turbid flow. Adequate clarification could not be achieved when the turbid flow rate was increased enough to prevent this deposition on the filter. As a result these flocs caused appreciable clogging of the porous stainless steel filters. In some cases backwashing was not successful in restoring the porosity of the filter. It is concluded that the filtration operation should be made a separate unit operation.

Electrode Grid Model

The electrode grid model was designed to perform similarly to the porous filter model but eliminate the filtration operation. In this way the clogging problem was also alleviated. In addition the original model was designed to decrease turbulence and acceleration forces.

As previously explained, however, the model as originally designed failed when operating electrochemically due to this lack of turbulence and mixing action. Even though adequate power was supplied to the electrodes to release sufficient flocculating electrolysis products, the flocs failed to form properly in this configuration. This points out the necessity for gentle turbulent mixing for adequate floc formation.

After modification the model worked much more satisfactorily and both stainless steel and aluminum electrodes were tested in the modified configuration. It is obvious from the results that aluminum electrodes are much better suited for electrochemical flocculation than are stainless steel. The aluminum electrodes dissolve readily and hydrolyze directly to give aluminum hydroxide as follows:



Satisfactory clarification, in the aluminum tests, was

accomplished using only electrochemical flocculation and settling. A high percentage of suspended materials was successfully removed and the effluent was in each case within U.S. Public Health Service standards. Even though this is acceptable for health standards, for esthetic reasons the addition of filtration will no doubt be desired by some persons and be necessary in some situations.

Power costs were lower with the aluminum electrodes on this model than with any other model or electrodes tested at comparable conductivities and influent concentrations. These were below \$0.10 per 1000 gallons for each test. At the high influent concentration near 100 mg l^{-1} the adjusted cost with aluminum electrodes was 8.9¢ per 1000 gallons compared to 16.8¢ for the stainless steel electrodes and 21.8¢ (assuming 75% of the turbid effluent usable) for the porous filter tests.

Significant chlorine residuals could not be detected when using aluminum electrodes. In this case the aluminum electrodes probably had an equilibrium potential less positive than the discharge potential of the chloride ion; thus no chlorine was evolved. This is not considered a disadvantage though. For maximum efficiency the disinfection should no doubt be a separate electrochemical operation.

It is certain that for electrochemical operation the electrode grid model can be improved significantly, for the design as tested

is definitely not considered an optimum or even a satisfactory design for this purpose.

Application of Results in Future Models or Systems

Results from the final model tested confirmed the potential for successful water clarification primarily through electrochemical means. In addition, disinfection may also be achieved in a like manner but not necessarily combined with the flocculation process. By utilizing these methods, the physical application of chemicals for water treatment is entirely eliminated. This is one of the primary problems with present small domestic water systems. The total water purification process is thus reduced to only three or possibly four unit operations, all of which may readily be automated with the exception of replacing electrodes. If filtration is desired a replaceable cartridge type filter might be particularly applicable for this system. Due to the low turbidity already achieved, the time between filter cartridge replacement should be fairly long. A proposed flow diagram for a small water treatment system is given in Figure 37.

As stated previously, the design of the electrode grid model was definitely not considered the optimum possible. The following

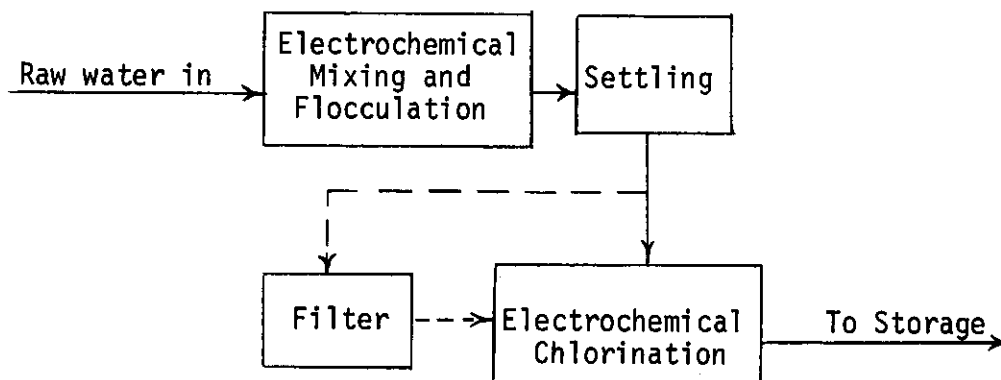


Figure 37. Schematic diagram of proposed small domestic water treatment system.

is an example of a system design, such as that above, optimized for electrochemical operation entirely and considers the chlorination as a separate unit operation. This design will include the power consuming electrochemical operations only.

Flocculator Design

This design will be for a small domestic plant with a 500 gallon per day capacity. This should be adequate for the household needs of most farm families. It will be assumed also that the electrical conductivity of the raw water is 1,000 micromhos cm^{-1} or that the specific resistance is likewise 1,000 ohms-cm. It will also be assumed that the water is very turbid or is a very difficult water to treat.

The amount of Al^{3+} ions required from the dissolution of the aluminum electrodes for the secondary formation of $\text{Al}(\text{OH})_3$ will

be estimated by stoichiometric means. The necessary amount of filter alum, $\text{Al}_2(\text{SO}_4)_3 \cdot 14\text{H}_2\text{O}$, for water treatment is determined by the quality of water being treated. The required treatment dosage ranges from 3.5 ppm up to a high of 85 ppm for very difficult waters.* It will be assumed that 85 ppm of filter alum would be necessary in this case if treatment were by conventional means.

The equivalent amount of Al^{3+} required is

$$(85 \text{ mg l}^{-1}) (54/600) = 7.65 \text{ mg l}^{-1},$$

the approximate molecular weight of filter alum being 600 and the atomic weight of Al about 27. This is equal to approximately 29 mg Al^{3+} per gallon required for flocculation. If the capacity is 500 gal day⁻¹ and assuming 24 hour operation, the hourly flow rate is 20.83 gal hr⁻¹. Thus the hourly rate of Al^{3+} supplied by electrolysis must be

$$(0.029 \text{ g gal}^{-1}) (20.83 \text{ gal hr}^{-1}) = 0.604 \text{ g hr}^{-1}.$$

According to Faraday's Law, 96,500 coulombs (one faraday) will produce one gram equivalent weight of a substance by electrolysis. In the case of Al^{3+} this is $27/3 = 9$ grams. This assumes a current efficiency of 100 percent, which is the ratio of the quantity of substance obtained by electrolysis to the quantity obtained theoretically by Faraday's Law. Gas was visually observed

*Personal communication with Dr. T. D. Reynolds, Environmental Engineering Department, Texas A&M University.

being evolved from the anode in the experimental model. This was probably O_2 since no chlorine residuals were found. Therefore, the current efficiency for Al^{3+} dissolution is going to be conservatively estimated as 50 percent at this time for lack of a better figure.

The number of faradays, F, required is thus

$$0.604 \text{ g hr}^{-1} / (9 \text{ g F}^{-1}) (0.5) = 0.1342 \text{ F hr}^{-1}.$$

This is equal to

$$(0.1342 \text{ F hr}^{-1}) (96,500 \text{ amp-sec F}^{-1}) = 12,950.3 \text{ amp-sec hr}^{-1}$$

which gives the required current of

$$12,950.3 \text{ amp-sec hr}^{-1} / 3,600 \text{ sec hr}^{-1} = 3.6 \text{ amperes.}$$

The electrical potential must be great enough to cause the dissolution of the aluminum electrode. The equilibrium potential or oxidation potential for the reaction $Al \rightarrow Al^{3+} + 3e^-$ is about 1.66 volts. This is the half-cell reaction so that the total potential between electrodes must be at least 3.32 volts. Four volts will be used to provide a slight overpotential although this might need to be higher in actual operation.

It will initially be assumed that the electrode spacing, d , should be as close as possible. One-tenth of an inch or 0.254 cm will be tried. From Ohm's Law the resistance, R , which will satisfy the voltage and amperage requirement is $V/I = 4.0/3.6 = 1.11$ ohms. Recalling that the specific resistance, r , is 1,000 ohms-cm the required area, A , of the electrodes is

$$A = rd/R = (1,000 \text{ ohms-cm}) (0.254 \text{ cm}) / 1.11 \text{ ohms} = 231 \text{ cm}^2.$$

This small area probably would be supplied by only two aluminum sheets of approximately 230 to 240 cm² each. First 16-gauge aluminum sheets will be considered which are 0.05 inches thick and weigh about 0.706 lbs ft⁻². An alternating current of low frequency will be used so that both electrodes will be consumed. The total weight of metal in this case is

$$(462 \text{ cm}^2 / 929 \text{ cm}^2 \text{ ft}^{-2}) (0.706 \text{ lbs ft}^{-2}) = 0.35 \text{ lbs,}$$

or is equal to 158.8 grams. The electrodes are consumed at a rate of

$$(0.604 \text{ g hr}^{-1}) (24 \text{ hr day}^{-1}) = 14.5 \text{ g day}^{-1}.$$

At this rate the electrodes would last

$$158.8 \text{ gram} / 14.5 \text{ g day}^{-1} \approx 11 \text{ days.}$$

In this case it would be helpful to space the electrodes farther apart in order to increase the time between electrode replacement. By increasing d to 1.0 cm. the necessary electrode area is increased to

$$(1000 \text{ ohm-cm}) (1.0 \text{ cm}) / 1.11 \text{ ohms} = 900.9 \text{ cm}^2.$$

In addition, 10-gauge metal which is 0.1 inch thick and weighs 1.41 lbs ft⁻², will be used in place of 16 gauge. The total weight of metal is now

$$(1801.8 \text{ cm}^2 / 929 \text{ cm}^2 \text{ ft}^{-2}) (1.41 \text{ lbs ft}^{-2}) = 2.37 \text{ lbs}$$

or 1,238.3 grams. The time between electrode replacement may now be computed as

$$1,238.3 \text{ grams} / 14.5 \text{ grams day}^{-1} = 85 \text{ days.}$$

For practical purposes this will probably range between 70 to 80

days.

Operating costs. Operating costs will include both power costs and the cost of the consumable electrodes. It will be assumed that the cost of electricity is \$0.02 per KWH and that the sheet aluminum costs \$0.94 per pound.

Since the flow rate or system capacity is 500 gal day^{-1} it will therefore take 48 hours to produce 1000 gallons. The power requirement is

$$(4 \text{ volts}) (3.6 \text{ amps}) = 14.4 \text{ watts} = 0.0144 \text{ KW.}$$

The power cost per 1,000 gallons is therefore

$$(48 \text{ hrs}) (0.0144 \text{ KW}) (\$0.02 \text{ KWH}^{-1}) = \$0.013 \text{ per 1000 gallons.}$$

Assuming the electrodes are replaced every 75 days the electrode costs may be estimated as follows:

$$(2.73 \text{ lbs/75 days}) (2 \text{ days/1000 gallons}) (\$0.94 \text{ lb}^{-1}) = \\ \$0.068 \text{ per 1000 gallons.}$$

The total flocculation cost is therefore

$$\$0.068 + \$0.013 = \$0.081 \text{ per 1000 gallons.}$$

It must be remembered that this was considering a very difficult water to treat. It is possible that this could be cut by more than tenfold depending on the quality of raw water being treated. However this figure is competitive with conventional coagulation and sedimentation costs plus it has the advantage of automation.

Chlorinator Design

For chlorine production it is desirable to have an inert electrode so that its dissolution does not compete with chlorine evolution. The most common inert electrodes are platinum, gold, arc carbon or graphite. However, graphite is the only anode material of industrial use in chlorine cells.

Normally anions other than OH^- are not discharged at the anode except under peculiar conditions. This is because the discharge potential of the OH^- ion from aqueous solutions is one of the lowest and hence will be one of the first to be discharged. The exception of this is with the halide ions in which the mechanism of the discharge process is not well known. The reversible electrode potential of the oxygen electrode in a neutral solution is +0.81 volts. On the other hand the standard electric potential of the chlorine electrode is +1.36 volts. Theoretically, therefore, reversible conditions for the simultaneous discharge of both ionic species at pH 7 could be obtained only if it were possible to raise the Cl^- ion concentration to an extremely high value. In practice, however, the effective discharge potential of the OH^- ion is much higher than the reversible value considered above. This is mainly due to the fact that the overpotential required to discharge the OH^- ion is in general much higher than that for the Cl^- ion. In addition, the anodic discharge is always accompanied by a decrease in pH, which further shifts the reaction condition in favor of

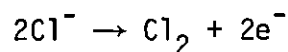
chlorine gas evolution.

As a result of the above facts it is possible for chlorine evolution at the anode although there will always be a certain amount of oxygen produced. The proportion of Cl_2 produced to O_2 produced depends on such things as Cl^- ion concentration, the anode current density, temperature, and hindrance of OH^- ion migration to the anode.

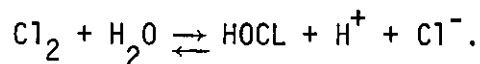
The chloride ion is one of the most common ions present in practically all waters but occurs with tremendous variations in concentration. However, it was demonstrated in these research results that adequate chlorine residuals could be obtained with chloride concentrations as low as 10 to 20 ppm. Therefore, even though there is no control on the concentration of Cl^- ions, the concentration in most waters should be adequate for sufficient chlorine gas production.

The only parameter which will probably be practical to control as far as optimizing chlorine evolution in a small system is the anode current density. It should be kept as high as possible in order to increase the overpotentials of other competing anionic species in favor of Cl^- . The anode current density usually ranges between 0.02 and 0.12 amps cm^{-2} in industrial applications.

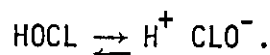
The primary reaction at the anode for the chlorine evolution is



followed by secondary reactions



The hypochlorous acid formed is a weak acid that ionizes to form the hypochlorite ion as follows:



The actual amount of chlorine gas necessary to produce sufficient residual to insure disinfection depends on several things; some of which are the chlorine demand, time of contact, and pH of the water. It will be assumed here that the time of contact will be sufficiently long so that predominately combined chlorine residuals will insure adequate disinfection. It will be assumed that a total chlorine residual of up to 2.0 mg l^{-1} can be obtained from the evolution of about 6.0 mg l^{-1} chlorine gas.

The requirements of 6.0 mg l^{-1} will equal $22.71 \text{ mg gal}^{-1}$ and require a production rate of

$$(0.02271 \text{ g gal}^{-1}) (20.83 \text{ gal hr}^{-1}) = 0.473 \text{ g hr}^{-1}.$$

One gram equivalent weight of chlorine is $35.4/2 = 17.7$ grams; thus it follows that

$$0.473 \text{ g hr}^{-1}/17.7 \text{ g F}^{-1} = 0.0267 \text{ faradays hr}^{-1}$$

is required for sufficient evolution of chlorine. However, this assumes a current efficiency of 100 percent. Current efficiencies in industrial plants range from 80 to 97 percent in the production of chlorine gas. It will be conservatively estimated in this design to be 60 percent. Therefore, the needed current is estimated

to be

$$(0.0267 \text{ F hr}^{-1}) (96,500 \text{ amp-sec F}^{-1}) / (3600 \text{ sec hr}^{-1})(0.6) = 1.20 \text{ amps.}$$

Recalling the current density requirement, the lower value of 0.02 amp cm^{-2} will be used to obtain the electrode area needed as

$$1.2 \text{ amps} / 0.02 \text{ amp cm}^{-2} = 60 \text{ cm}^2.$$

The applied potential between electrodes in industry usually ranges between 3.7 and 4.5 volts although the requirement could possibly be higher in this case. However, at this time 4.5 volts will be used. In this example the total resistance, R, must be 4.5 volts/1.2 amps = 3.75 ohms. Solving for the distance between the electrodes, d, we obtain

$$d = RA/r = (3.75 \text{ ohms}) (60 \text{ cm}^2) / 1,000 \text{ ohm-cm} = 0.225 \text{ cm}$$

or approximately 0.1 inch. If this spacing is not practical it may be increased by increasing the applied potential proportionally.

Costs. The cost of chlorination will primarily be power costs. The inert graphite electrodes are very inexpensive and should have to be replaced infrequently if ever.

The power consumed in this example is

$$(4.5 \text{ volts}) (1.2 \text{ amps}) = 5.4 \text{ watts or } 0.0054 \text{ KW.}$$

Thus the cost per 1000 gallons is estimated as

$$(48 \text{ hrs}) (0.0054 \text{ KW}) (\$0.02/\text{KWH}) = \$0.0052/1000 \text{ gal}$$

or about 5 mils per 1000 gallons.

The total operating costs for the power consuming operations

in this example are thus estimated in cents per 1000 gallons as follows:

Al electrodes for flocculation ---6.8¢

Electricity for flocculation -----1.3¢

Electricity for chlorination-----0.52¢

Total 8.6¢/1000 gallons.

The primary advantage of this proposed electrochemical system over conventional water treatment methods is the ease with which it may be automated. This system should require only a small amount of attention, either monthly or bi-monthly. This is a greater asset at the present time to small rural systems than to municipal systems. However, with the increasing shortage of labor, this may be an important consideration for municipal plants in the future.

The costs presented here are operating costs only. However, they have been estimated considering the most difficult conditions probable. With the addition of capital costs, the total cost of water treatment should compare very favorably with the cost of conventional treatment. This aspect needs to be evaluated more fully through the construction of a pilot system.

CHAPTER VI

SUMMARY AND CONCLUSIONS

An investigation was made into the use of electrophoretic and electrochemical means to clarify and purify water. Numerous design concepts were advanced and tested through the use of laboratory models. The design concepts were evaluated both from an operational and economical standpoint.

This research was pertinent for there is a pressing need for new and superior water treatment methods and systems which can be used to alleviate the alarming growth and wide variety of pollutants present in waters today. Of particular need at the present time is a small water treatment system for rural families which will require a minimum amount of attention for satisfactory operation. Systems utilizing electrophoretic and electrochemical means appeared well adapted for this purpose, for they have the ability to be readily automated. This investigation was focused primarily on the small domestic type system, although the possibility of application to other municipal and industrial uses was not dismissed.

The models constructed for evaluation incorporated designs for pollutant removal entirely by electrophoresis on one hand to designs which could accommodate both electrophoretic and electrochemical processes on the other. One model was tested which was

designed for electrophoretic pollutant removal only. This was the horizontal parallel plate model which was tested with very low conductivity water in order to limit the process occurring in the model primarily to electrophoresis.

A porous stainless steel filter model was designed which could function either electrophoretically or electrochemically. To limit the process to electrophoresis, again low electrical conductivity water was used. When operating electrochemically at higher conductivities, this model essentially combined the entire spectrum of conventional water treatment operations into one unit. A problem with clogging, however, indicated that filtration should be made a separate operation. As a consequence, an electrode grid model was also designed to operate either electrophoretically or electrochemically and combined the same operations as did the porous filter model with the exception of the filtration process. It was tested with two electrode materials, stainless steel and aluminum.

Clay suspensions using commercial kaolinite clay in various concentrations were mixed in order to evaluate suspended particulate matter removal by each of the models. These measurements were made by turbidimetric means. Sewage effluent was mixed with the suspension in order to evaluate bacteria removal or disinfection as measured by the most probable number (MPN) of coliform bacteria. In addition, electrophoretic mobility determinations

were made on several Texas soils for the purpose of characterizing them for ease of electrophoretic removal. These mobilities were determined by the microscope method of electrophoresis.

As a result of exhaustive testing of the various models and the design concepts which they represented, the following conclusions are drawn:

1. Turbid waters can be successfully clarified by electrophoresis, but the process is limited to waters of very low electrical conductivity if the electrodes are exposed to the bulk suspension being clarified. At higher electrical conductivities, the electrolysis products are sufficient to disrupt entirely the electrophoretic transport. In addition, primary and secondary chemical reactions of the electrolysis products cause changes in the properties of the suspension. In some cases, they neutralize the negative charges of the colloidal particles rendering them unaffected by the applied electric field. Under this situation, the use of electrophoresis as a sole means of treatment for natural waters is virtually impossible.

2. In order to obtain clarification of natural waters entirely by electrophoresis, the electrodes must be separated from the bulk colloidal transport area. The method described by Bier (3,4,11,19) appears to be a logical way to accomplish this.

3. The economic justification of treatment purely by electrophoresis is debatable if suspended colloidal particulate matter removal is the primary purpose of treatment. The operating costs

are much too dependent upon the electrical conductivity of the water. This method would probably be limited to waters with electrical conductivities of 200 to 300 micromhos cm^{-1} and below. Even at these conductivities, the price paid for convenience and automation over conventional methods would be considerable. However, if ionic pollutant removal is also necessary, the utilization of electrophoresis along with dialysis might be a practical dual purpose use of the applied electric field.

4. Water clarification by electrochemical means was successful both operationally and economically. For optimum economic operation, a design should consider only electrochemical flocculation without regard to any additional electrophoretic clarification capabilities.

5. Aluminum appears to be better than stainless steel as an electrode material for electrochemical flocculation.

6. When electrochemical flocculation is used for clarification, the costs are roughly proportional to the pollutant concentration and inversely proportional to the electrical conductivity.

7. Disinfection as well as flocculation is possible by electrochemical means. It is accomplished by chlorine gas evolution at the anode due to electrolysis and should be a separate unit operation for maximum efficiency and effectiveness.

8. A small practical water system appears entirely feasible which consists of electrochemical flocculation, settling, and electrochemical disinfection operations. The use of filtration would be optional but if desired, a replaceable cartridge type filter might be well adapted here. In this case the entire system could be automated with the exception of periodic replacement of flocculator electrodes and filter if used. The settling basin may even be designed so that it flushes itself automatically at given intervals. The physical addition of chemicals is entirely eliminated with this system.

CHAPTER VII

SUGGESTIONS FOR FUTURE RESEARCH

It is suggested that future research into the use of electrophoretic and electrochemical phenomena in the field of water purification take two primary and separate approaches as follows:

1. The application of electrophoresis in combination with electro dialysis in which both processes share the same electric field. This could be accomplished through the appropriate use of ion selective membranes, simple cellulosic dialyzing membranes, and controlled porosity filters. In this way the power cost for both processes combined should be little more than that for the electro dialysis alone and would eliminate the conventional treatment which normally would accompany the desalting operation.

2. The construction of a small pilot water treatment plant which incorporates the electrochemical system outlined in Figure 37, Chapter V. Evaluation should include such things as life expectancy, maintenance difficulties, and actual fixed and operating costs which would be incurred by the system in the field.

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APPENDIX A

GLOSSARY OF TERMS AND SYMBOLS

Terms

- Electrophoresis - the movement of electrically charged particles suspended in a fluid under the influence of an electric field
- Electrokinetic - normally considered to encompass electrophoresis, electro-osmosis, sedimentation potential and streaming potential
- Electrochemical - pertaining to the relationship between chemical transformations and energy in reactions which involve electrical energy external to the system
- Electrolysis - process by which electrical energy is converted into chemical energy
- Dialysis - separation of solutes, usually by membrane processes
- Electrodialysis - dialysis accelerated by the action of an electric field which causes ions to flow against a concentration gradient

- Zeta Potential - effective electrical potential on a colloidal particle in uniform motion in a fluid
- Nephelometric - determination of concentration or particle size in suspension by measurement of transmitted or reflected light
- Hydrolyze - a chemical reaction in which a compound or element reacts with the ions of water (H^+ or OH^-)
- Stoichiometric - quantitative relationship with which elements combine
- Reversible Electrode Potential - the electrode potential adapted when the dissolution of an electrode and the deposition of its atoms are in equilibrium;
 $m \rightleftharpoons m^+ + e$, where $m = \text{metal}$
- Overpotential - the difference between the actual potential of an electrode operating irreversibly and the reversible potential
- Discharge or Dissolution Potential - the electric potential at which appreciable electrolysis or faradic current will flow; equal to the reversible potential plus some overpotential

Standard Electrode Potential - potential of electrodes which are in their standard state, i.e., standard pressure and temperature and at unit activity.

Symbols

- EM - Electrophoretic Mobility
- MPN - Most Probable Number
- V_F - Liquid Flow Velocity
- V - Electrophoretic Transport Velocity
- D - Dielectric Constant
- F - Faraday (96,500 coulombs)
- V - Volt
- I - Ampere
- R - Resistance
- A - Area
- fsl - Fine Sandy Loam
- sc1 - Sandy Clay Loam
- g - Gram
- l - Liter
- r - Specific Resistance
- d - Distance Between Electrodes
- x - Electric Field Strength
- η - Absolute Viscosity
- ζ - Zeta Potential

ϕ - Potential

μ - Micro or Micron

APPENDIX B

PRESENTATION OF DATA

Parallel Plate Tests

Test No.	Flow Rate (ml min ⁻¹)	Field Strength (volts cm ⁻¹)	Average Conductivity (μmhos cm ⁻¹)	Electric Current (amperes)	Effluent Concentration (mg l ⁻¹)	Percent Clay Removal	Percent Clear Effluent
A-1	1760	50.0	8.8	4.86	37.0	20	0
A-2	880	100.00	5.3	7.55	2.0	97	60
A-3	880	50.0	4.2	3.43	6.5	87	30
A-4	880	42.5	7.7	5.46	4.0	93	42
A-5	880	25.0	6.2	2.98	6.3	85	35
A-6	880	25.0	4.4	3.02	7.6	82	27
A-7	880	12.5	5.1	1.32	23.0	46	0
A-8	660	75.0	4.8	5.0	1.0	98	75
A-9	660	50.0	5.0	3.93	4.0	91	35
A-10	660	37.5	4.6	3.72	5.5	88	28
A-11	660	18.8	17.5	5.22	4.0	92	55
A-12	660	12.5	17.8	4.76	3.0	94	35

All Electrodes Charged

Test No.	Flow Rate (ml min ⁻¹)	Field Strength ₁ (volts cm ⁻¹)	Average Conductivity (μmhos cm ⁻¹)	Electric Current (amperes)	Clear Effluent Concentration (mg l ⁻¹)	Percent Clay Removal	Percent Clear Effluent
A-13	660	9.4	19.7	2.29	8.0	86	35
A-14	660	4.7	5.0	0.415	17.0	65	0
A-15	440	50.0	4.9	2.61	3.0	95	90
A-16	440	50.0	4.6	4.21	1.8	96	85
A-17	440	25.0	10.4	4.18	0.0	100	80
A-18	440	25.0	21.4	5.64	5.0	91	75
A-19	440	12.5	10.9	1.80	0.0	100	50
A-20	440	12.5	4.2	0.92	3.5	94	35
A-21	440	12.5	5.2	1.28	3.0	94	25
A-22	440	8.35	4.5	0.90	8.6	84	10
A-23	440	3.12	18.7	0.78	1.6	97	30
A-24	220	50.0	3.6	3.53	0.6	99	90
A-25	220	25.0	17.8	4.42	3.5	93	90
A-26	220	25.0	3.3	3.20	0.0	100	85
A-27	220	12.5	4.5	1.12	3.5	93	60
A-28	220	6.25	3.8	0.41	1.4	97	35

Test No.	Flow Rate (ml min ⁻¹)	Field Strength (volts cm ⁻¹)	Average Conductivity (μmhos cm ⁻¹)	Electric Current (amperes)	Clear Effluent Concentration (mg l ⁻¹)	Percent Clay Removal	Percent Clear Effluent
A-29	220	4.15	18.6	0.74	1.8	96	50
A-30	220	3.12	5.2	0.19	12.8	74	0
<u>Bottom 1/2 Electrodes Charged</u>							
B-1	440	75	4.1	3.15	4.2	91.8	85
B-2	440	50	4.2	2.24	4.3	91.6	50
B-3	440	25	3.5	1.24	9.2	82.8	0
<u>Top 1/2 Electrodes Charged</u>							
T-1	440	75	3.9	3.38	2.2	95.1	88
T-2	440	50	4.0	2.54	2.0	95.6	50
T-3	440	25	4.1	1.40	3.0	94.3	55
T-4	440	12.5	3.6	0.74	3.0	94.4	50

Porous Filter Tests

Test No.	Flow Rate (ml min ⁻¹)	Field Strength (volts cm ⁻¹)	Average Conductivity (μmhos cm ⁻¹)	Electric Current (amperes)	Clear Effluent Concentration (mg l ⁻¹)	Percent Clay Removal	Average Flow Rate Change (ml min ⁻¹ hr ⁻¹)
<u>165μ Filter Porosity, Low Conductivity</u>							
C-1	342	210	3.1	0.675	4.5	90.3	+2.7
C-2	342	180	2.9	0.567	8.6	81.5	-3.17
C-3	342	150	2.8	0.445	8.2	82.4	-9.1
C-4	342	120	3.5	0.40	3.0	93.5	+2.0
C-5	342	105	3.6	0.31	3.0	93.5	-0.9
C-6	342	90	3.7	0.27	3.5	92.5	+1.2
C-7	342	75	3.8	0.234	8.7	81.2	-1.4
C-8	224	80	5.3	0.31	0.6	98.6	-5.7
C-9	224	70	5.5	0.29	0.0	100.0	0.0
C-10	224	60	3.7	0.23	0.0	100.0	-2.0
C-11	224	50	4.0	0.19	1.4	97.1	+2.4
C-12	224	40	4.0	0.12	1.4	97.1	-3.4
C-13	224	30	4.5	0.10	3.3	93.3	+1.1
C-14	224	20	4.3	0.07	9.2	81.2	+3.7
C-15	224	10	4.0	0.025	24.5	50.0	-1.4

Test No.	Flow Rate (ml min ⁻¹)	Field Strength (volts cm ⁻¹)	Average Conductivity (μmhos cm ⁻¹)	Electric Current (amperes)	Clear Effluent Concentration (mg l ⁻¹)	Percent Clay Removal	Average Flow Rate Change (ml min ⁻¹ hr ⁻¹)
C-16	112	60	3.2	0.2	1.4	96.8	-1.9
C-17	112	50	3.1	0.13	1.4	96.8	-3.4
C-18	112	40	3.0	0.11	1.4	96.8	-3.2
C-19	112	35	3.15	0.11	1.4	96.8	+1.8
C-20	112	30	3.2	0.085	1.8	95.9	+2.3
C-21	112	25	3.2	0.07	1.4	96.8	-6.6
C-22	112	20	3.2	0.05	2.5	94.4	-3.0
C-23	112	15	3.3	0.03	3.5	92.1	-1.5
C-24	112	10	3.33	0.02	8.2	81.5	0.0
C-25	112	5	3.4	0.01	24.5	44.8	0.0
<u>5μ Filter, Low Conductivity</u>							
H-1	342	120	3.6	0.395	3.0	93.6	-5.6
H-2	342	90	4.2	0.31	1.8	96.1	0.0
H-3	342	75	4.2	0.285	0.0	100.0	0.0
H-4	342	60	4.3	0.23	0.0	100.0	0.0

Test No.	Flow Rate (ml min ⁻¹)	Field Strength (volts cm ⁻¹)	Average Conductivity (μmhos cm ⁻¹)	Electric Current (amperes)	Clear Effluent Concentration (mg l ⁻¹)	Percent Clay Removal	Average Flow Rate Change (ml min ⁻¹ hr ⁻¹)
H-5	342	45	4.2	0.17	1.4	97.0	-4.9
H-6	342	30	4.1	0.11	4.0	91.4	-10.7
H-7	342	15	4.4	0.06	28.2	39.5	-11.5
H-8	224	80	3.5	0.256	0.6	98.7	-12.0
H-9	224	70	3.6	0.24	0.0	100.0	-2.2
H-10	224	60	3.6	0.22	0.0	100.0	-1.6
H-11	224	50	3.7	0.19	0.0	100.0	-4.1
H-12	224	40	3.8	0.145	0.0	100.0	+11.5
H-13	224	30	3.8	0.105	0.0	100.0	-2.6
H-14	224	20	3.9	0.07	11.4	74.7	0.0
H-15	224	10	4.0	0.03	21.8	51.7	-0.9
H-16	112	50	5.2	0.18	4.0	93.4	-0.8
H-17	112	40	4.8	0.125	5.6	90.7	-5.4
H-18	112	30	5.0	0.095	4.5	92.6	+2.2
H-19	112	20	5.0	0.07	5.2	91.4	-0.5

Test No.	Flow Rate (ml min ⁻¹)	Field Strength (volts cm ⁻¹)	Average Conductivity (μmhos cm ⁻¹)	Electric Current (amperes)	Clear Effluent Concentration (mg l ⁻¹)	Percent Clay Removal	Average Flow Rate Change (ml min ⁻¹ hr ⁻¹)
H-20	112	15	5.1	0.04	7.2	88.1	-0.5
H-21	112	10	5.1	0.02	11.8	80.5	+0.6
H-22	112	5	5.2	0.01	14.0	76.8	-0.3
<u>Electrical Conductivity Tests</u>							
E-1	224	10	690	2.86	2.5	95.1	-10.8
E-2	224	8	740	2.21	0.6	98.8	-19.5
E-3	224	6	760	1.59	3.2	93.8	-35.6
E-4	224	4	765	0.97	6.0	88.3	-17.0
E-5	224	2	770	0.31	16.0	68.9	-
E-6	224	10	359	1.15	1.4	96.8	-25.5
E-7	224	6	364	0.60	4.5	89.8	-24.0
E-8	224	4	369	0.383	16.0	63.9	-11.2
E-9	224	2	370	0.14	18.2	58.9	-

Test No.	Flow Rate (ml min ⁻¹)	Field Strength (volts cm ⁻¹)	Average Conductivity (μmhos cm ⁻¹)	Electric Current (amperes)	Clear Effluent Concentration (mg l ⁻¹)	Percent Clay Removal	Average Flow Rate Change (ml min ⁻¹ hr ⁻¹)
<u>Influent Concentration = 3.25 mg l⁻¹</u>							
D-1	224	10	383	1.47	0.0	100.0	-33.2
D-2	224	8	387	1.15	0.0	100.0	-52.9
D-3	224	6	400	0.92	0.0	100.0	-8.4
D-4	224	4	413	0.59	0.10	97.1	-7.6
D-5	224	2	421	0.23	0.0	100.0	-16.3
D-6	224	1	431	0.10	0.75	75.0	-17.0
<u>Influent Concentration = 9.6 mg l⁻¹</u>							
D-7	224	10	398	1.55	1.4	85.2	-21.2
D-8	224	8	413	1.29	1.6	83.2	-19.0
D-9	224	6	427	0.94	1.4	85.26	-5.4
D-10	224	4	434	0.60	0.6	93.85	-1.6
D-11	224	2	442	0.26	0.6	93.85	-

Test No.	Flow Rate (ml min ⁻¹)	Field Strength (volts cm ⁻¹)	Average Conductivity (μmhos cm ⁻¹)	Electric Current (amperes)	Clear Effluent Concentration (mg l ⁻¹)	Percent Clay Removal	Average Flow Rate Change (ml min ⁻¹ hr ⁻¹)
D-12	224	20	337	2.74	2.5	97.33	-
D-13	224	16	367	2.61	1.3	98.6	-26.1
D-14	224	14	328	1.93	23.5	74.7	-22.8
D-15	224	10	328	1.31	14.5	84.4	-32.6

Influent Concentration = 93.5 mg l⁻¹

Electrode Grid Tests

Test No.	Grid Potential (volts)	Electrical Current (amperes)	Average Conductivity ($\mu\text{mhos cm}^{-1}$)	Influent Concentration (mg l ⁻¹)	Average Effluent Concentration (mg l ⁻¹)	Percent Clay Removal
<u>Stainless Steel Electrodes</u>						
SS-1	45.8	6.0	86.9	13.1	3.1	76.3
SS-2	42.6	5.0	82.0	13.9	3.6	74.1
SS-3	36.8	4.0	84.0	13.9	3.6	74.1
SS-4	30.5	3.0	84.0	13.2	3.75	70.0
SS-5	25.5	2.5	86.2	13.2	5.4	59.1
SS-6	21.3	2.0	87.7	13.2	5.75	56.4
SS-7	12.4	1.0	87.7	13.2	14.5	-
SS-8	24.0	2.7	100.0	48.4	10.9	77.5
SS-9	20.0	2.35	101.0	48.4	20.25	58.2
SS-10	26.0	2.2	90.9	48.4	22.8	52.9
SS-11	35.2	5.0	100.0	98.0	11.9	87.9
SS-12	33.1	4.0	93.5	98.0	17.1	82.6

Test No.	Grid Potential (volts)	Electric Current (amperes)	Average Conductivity ($\mu\text{mhos cm}^{-1}$)	Influent Concentration (mg l^{-1})	Average Effluent Concentration (mg l^{-1})	Percent Clay Removal
SS-13	26.1	3.0	93.5	98.0	38.6	60.6
SS-14	15.5	5.0	317.4	24.0	3.85	84.0
SS-15	14.2	4.0	323.0	24.0	3.85	84.0
SS-16	11.5	3.0	333.3	24.0	4.95	79.4
SS-17	8.35	2.0	333.3	24.0	15.0	37.5
SS-18	15.8	5.0	328.0	48.5	8.1	83.3
SS-19	13.4	4.0	345.0	48.5	8.9	81.6
SS-20	10.6	3.0	357.1	48.5	18.2	62.5
SS-21	15.5	5.0	342.4	98.5	15.2	84.6
SS-22	12.7	4.0	357.0	98.5	12.4	87.4
SS-23	9.9	3.0	370.3	98.5	23.7	75.9
SS-24	11.1	6.0	657.8	17.9	4.5	74.9
SS-25	9.3	5.0	689.6	17.9	6.05	67.7
SS-26	7.8	4.0	714.2	17.9	5.8	67.6
SS-27	6.5	3.0	740.7	17.9	6.25	66.6
SS-28	5.1	2.0	740.0	17.9	8.95	50.0
SS-29	10.6	6.0	685.0	48.3	10.7	77.8

Test No.	Grid Potential (volts)	Electric Current (amperes)	Average Conductivity ($\mu\text{mhos cm}^{-1}$)	Influent Concentration (mg l ⁻¹)	Average Effluent Concentration (mg l ⁻¹)	Percent Clay Removal
SS-13	26.1	3.0	93.5	98.0	38.6	60.6
SS-14	15.5	5.0	317.4	24.0	3.85	84.0
SS-15	14.2	4.0	323.0	24.0	3.85	84.0
SS-16	11.5	3.0	333.3	24.0	4.95	79.4
SS-17	8.35	2.0	333.3	24.0	15.0	37.5
SS-18	15.8	5.0	328.0	48.5	8.1	83.3
SS-19	13.4	4.0	345.0	48.5	8.9	81.6
SS-20	10.6	3.0	357.1	48.5	18.2	62.5
SS-21	15.5	5.0	342.4	98.5	15.2	84.6
SS-22	12.7	4.0	357.0	98.5	12.4	87.4
SS-23	9.9	3.0	370.3	98.5	23.7	75.9
SS-24	11.1	6.0	657.8	17.9	4.5	74.9
SS-25	9.3	5.0	689.6	17.9	6.05	67.7
SS-26	7.8	4.0	714.2	17.9	5.8	67.6
SS-27	6.5	3.0	740.7	17.9	6.25	66.6
SS-28	5.1	2.0	740.0	17.9	8.95	50.0
SS-29	10.6	6.0	685.0	48.3	10.7	77.8

Test No.	Grid Potential (volts)	Electrical Current (amperes)	Average Conductivity ($\mu\text{mhos cm}^{-1}$)	Influent Concentration (mg l^{-1})	Average Effluent Concentration (mg l^{-1})	Percent Clay Removal
SS-30	9.0	5.0	722.0	48.3	7.85	83.7
SS-31	7.4	4.0	751.8	48.3	9.95	79.4
SS-32	5.7	3.0	763.0	48.3	17.9	62.9
SS-33	4.3	2.0	769.0	48.3	22.9	52.6
SS-34	10.5	6.0	694.4	92.4	17.6	81.0
SS-35	8.8	5.0	699.3	92.4	15.5	83.2
SS-36	7.2	4.0	751.8	92.4	14.0	84.8
SS-37	6.0	3.0	813.0	92.4	25.0	72.9
<u>Aluminum Electrodes</u>						
A1-1	13.0	4.0	305	18.2	3.0	83.5
A1-2	10.2	3.0	305	18.2	3.5	80.8
A1-3	7.1	2.0	305	18.2	9.95	45.3
A1-4	10.7	5.0	331	56.8	3.85	93.2
A1-5	9.4	4.0	333	56.8	4.5	92.1
A1-6	7.6	3.0	350	56.8	5.0	91.2
A1-7	5.5	2.0	350	56.8	11.3	80.1

Test No.	Grid Potential (volts)	Electrical Current (amperes)	Average Conductivity ($\mu\text{mhos cm}^{-1}$)	Influent Concentration (mg l^{-1})	Average Effluent Concentration (mg l^{-1})	Percent Clay Removal
A1-8	3.3	1.0	350.0	56.8	17.9	68.5
A1-9	12.4	5.0	319.0	105.0	5.8	94.5
A1-10	11.1	4.0	330.0	105.0	6.3	94.0
A1-11	9.0	3.0	334.0	105.0	8.6	91.8
A1-12	6.4	2.0	357.0	105.0	19.0	81.9
A1-13	32.0	4.0	89.7	46.0	5.2	88.7
A1-14	42.3	3.0	89.7	46.0	5.9	87.2
A1-15	30.1	2.0	89.7	46.0	14.95	67.5