

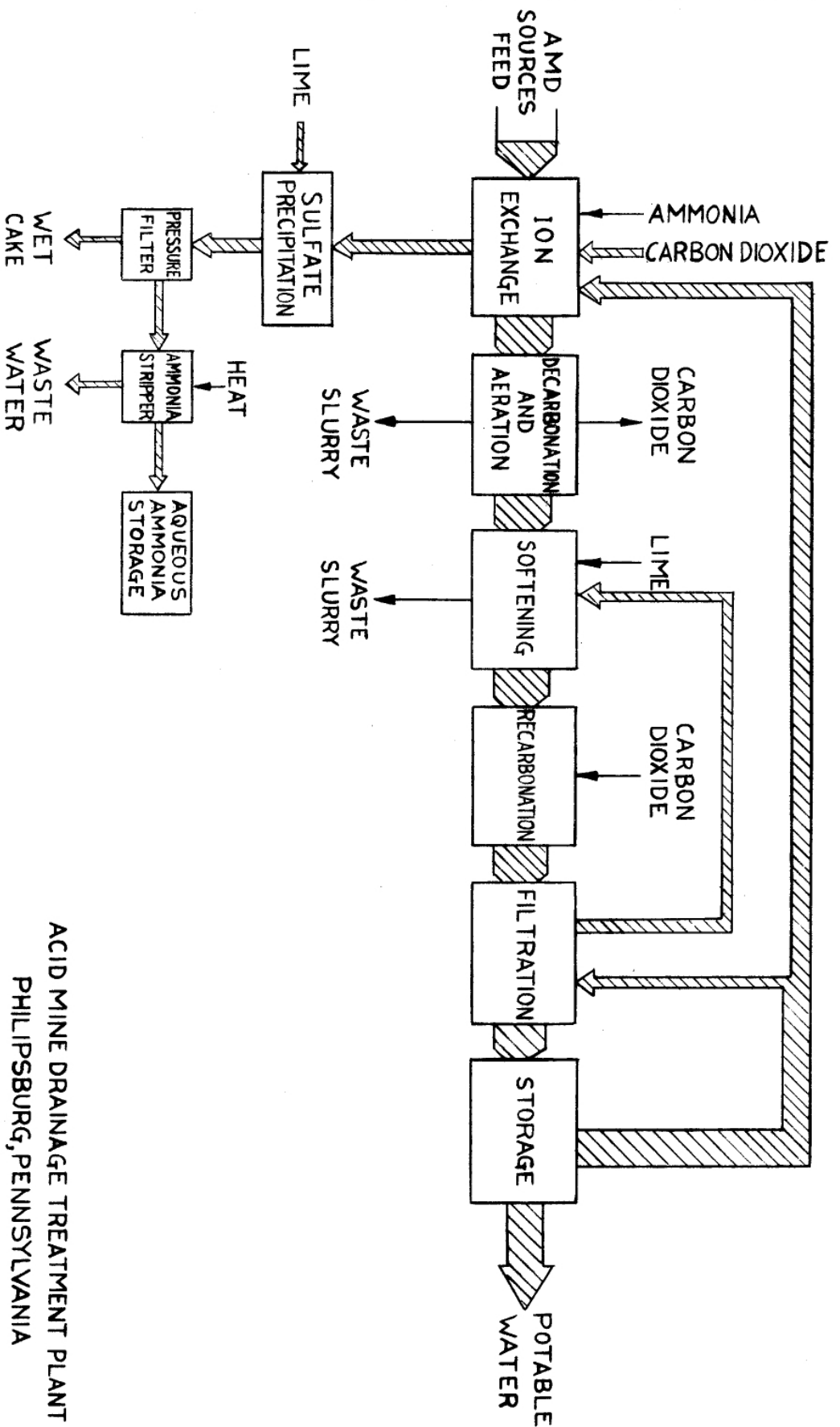
DESCRIPTION OF PLANT AND PROCESS

The proposed plant is designed to produce potable water of a quality that will meet U.S. Public Health Service Drinking Water Standards. The following tabulation is a partial listing of recommended physical and chemical characteristics which will govern the quality of the product. This list has been abstracted from U.S.P.H.S. Standards 1962:

U.S. Public Health Service Drinking Water Standards

<u>Substance</u>	<u>Limit</u>
Physical Characteristics	
Turbidity	5 units
Color	15 units
Threshold Odor Number	3
Chemical Characteristics	
Alkyl Benzene Sulfonate (ABS)	0.5 mg/l
Chloride (Cl)	250.
Copper (Cu)	1.
Cynide (CN)	0.01
Iron (Fe)	0.3
Manganese (Mn)	0.05
Nitrate (NO ₃)	45.
Sulfate (SO ₄)	250.
Total Dissolves Solids (TDS)	500.
Zinc (Zn)	5.

A schematic flow diagram, Figure 7, shows the principal steps of the proposed treatment plant. Raw AMD water is pumped from collection sumps at the sources and is passed through ion exchange towers. In the towers the acid-forming anions are replaced by bicarbonates. The water is then sprayed into a decarbonator, in which the carbon dioxide is removed. The CO₂ - free water is collected in a catch basin under the decarbonator and is transported by gravity through the remaining four



ACID MINE DRAINAGE TREATMENT PLANT
 PHILIPSBURG, PENNSYLVANIA

SCHEMATIC OF PROPOSED TREATMENT PLANT

DATE 6-30-69
 FIGURE 7

W. O. 2674 - 01

process steps. It is fed to an aerator system in which the heavy metals are separated, and then to a softener in which the hardness is removed by liming. The softened water is stabilized in the recarbonation step, is further clarified by filtration and finally stored for use. The schematic qualitatively indicates the amount of water recycled from storage to the process, mainly to the ion exchange step. The recycle to the filtration step is reclaimed in the softener.

The ammonia recovery is shown directly under ion exchange. The ammonia is released from the sulfate-rich, spent regenerant solution by chemical precipitation with lime. The solids are separated from the solution in a pressure filtration step and the resulting ammonia-bearing filtrate is fed to a distillation step. Ammonia is recovered from the condensate as a pure concentrate and stored.

The factors which determine sizing of equipment are (1) the concentration of dissolved solids, (2) the concentration of particulates, (3) the extent of treatment necessary to produce the desired quality and (4) the desired hydraulic load.

The ion exchange resin type and the size of exchange equipment are determined by factors (1) and (4), the dissolved solids, specifically the anion content, and the hydraulic load. The magnitude of aeration which removes the heavy metals from solution is controlled by the ferrous iron concentration which, in turn, determines the oxygen demand and the length of the reaction period (detention time). The size of the softening-coagulation unit depends only on the hydraulic load; the lime consumption will vary with the water hardness and completeness of CO₂ removal in the previous step. The turbidity of the feed is expected to have a minor affect on the equipment because visual inspection of the waters indicated a very small particulate content.

The treatment plant will be designed to process 684,000 gallons per day of acid mine drainage water based on the design water analysis shown in Table 2. The excess flow is required to provide for internal use in the treatment process as follows:

Ion exchange regeneration and first rinse	24,000 gpd
Ion exchange - second rinse	114,000 gpd
Lime slurry	7,500 gpd
Filter backwash and rinse	28,500 gpd
Boiler feed	<u>10,000</u> gpd
Total	184,000 gpd

It is expected that with some plant operating experience most of the second rinse water can be returned to the process for recovery by decarbonation and aeration. The filter backwash and rinse can also be recycled to the softener. With full recirculation, the plant output would be 620,000 gal per day.

The following sections describe in detail the complete plant and process:

Raw Water Supply

Ion Exchange System - Service

Regeneration

Ammonia Regenerant Recovery

Decarbonation, Aeration and Settling

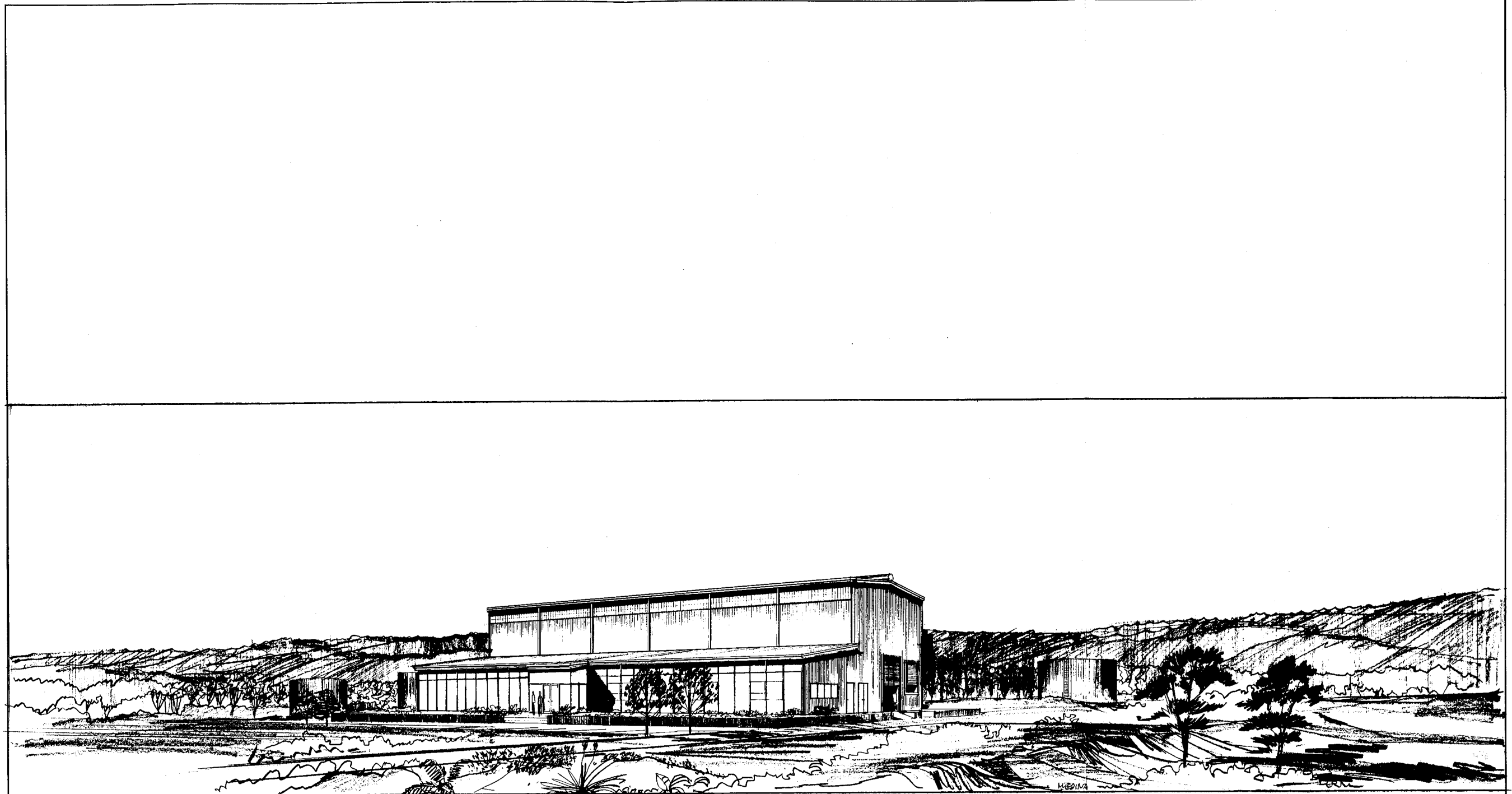
Softening

Recarbonation

Gravity Filtration

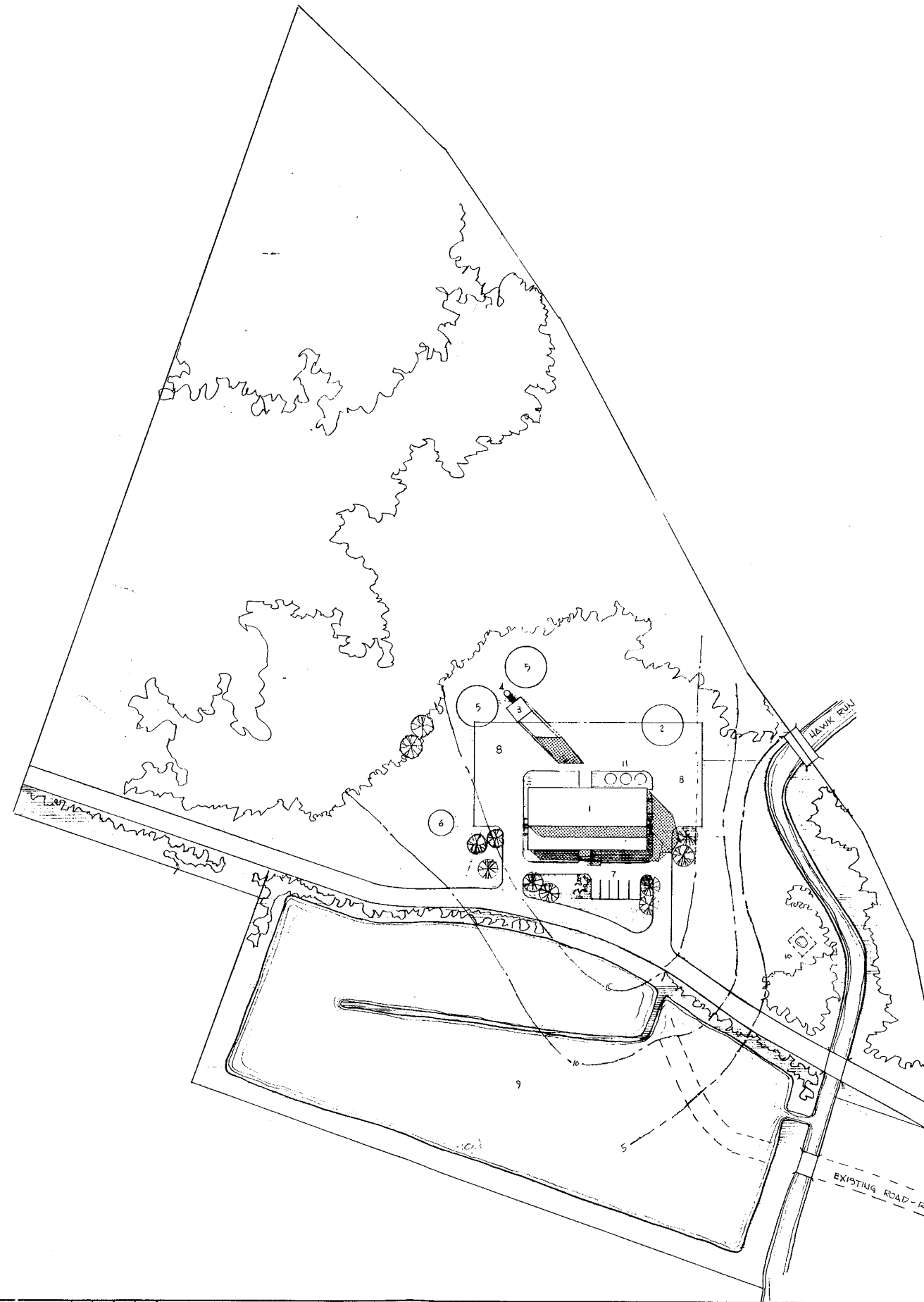
Product Storage

Waste Collection

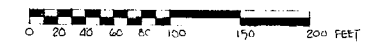
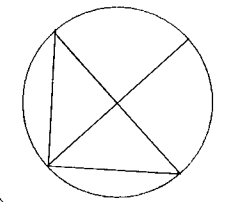


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 ENGINEERS AND CONSTRUCTORS
 ORADELL, N. J. HEMPSTEAD, N. Y. LOS ANGELES, CALIF.
 ACID MINE DRAINAGE TREATMENT PLANT
 PHILIPSBURG, PENNSYLVANIA
 PERSPECTIVE
 COMMONWEALTH OF PENNSYLVANIA
 DEPARTMENT OF MINES & MINERAL INDUSTRIES
 HARRISBURG, PENNSYLVANIA
 DRAWN BY *Hedulla* DATE 6/25/69 SCALE
 CHECKED BY *W. O. 2674* ARCHITECT
 DWG. 1401



- DATA
- 1 SERVICE BUILDING
 - 2 WATER STORAGE TANK
 - 3 DECARBONATOR
 - 4 AERATOR
 - 5 SETTLING BASIN
 - 6 SOFTENER
 - 7 PARKING
 - 8 SERVICE PARKING AND LOADING
 - 9 LAGOON
 - 10 MINE HOLE No 1
 - 11 PRECIPITATORS AND LIME SILO



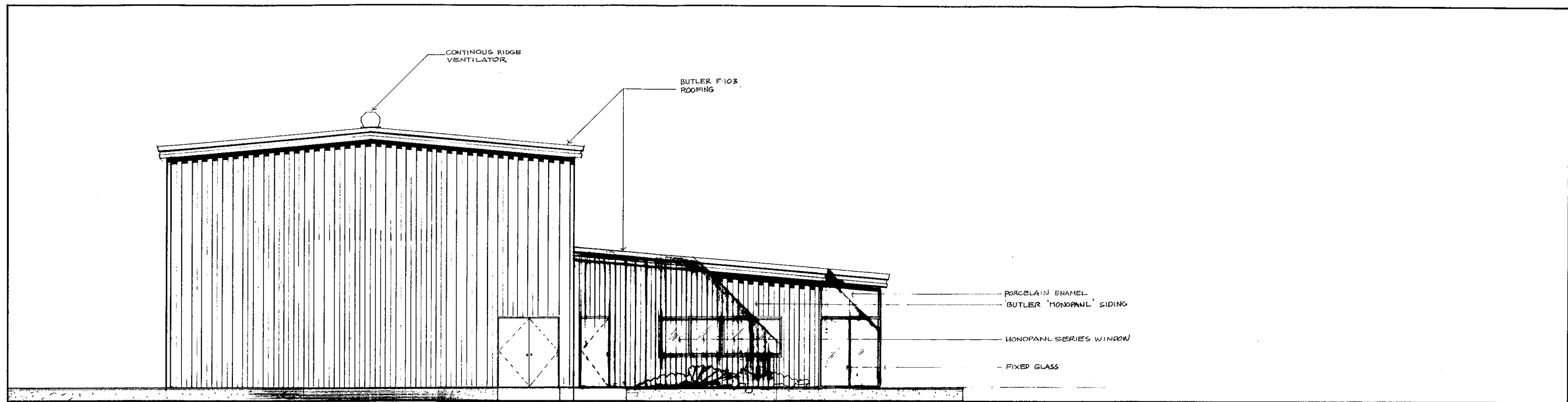
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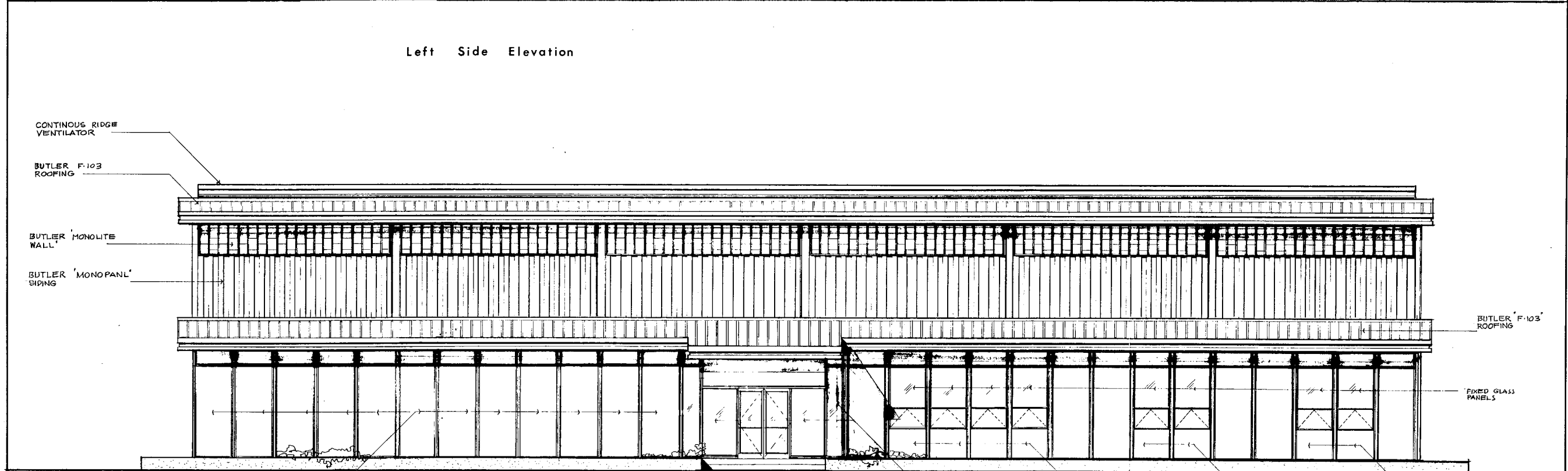
ACID MINE DRAINAGE TREATMENT PLANT
 PHILIPSBURG, PENNSYLVANIA
 SITE PLAN

COMMONWEALTH OF PENNSYLVANIA
 DEPARTMENT OF MINES & MINERAL INDUSTRIES
 HARRISBURG, PENNSYLVANIA

DRAWN BY: *W. O.* DATE: 6/25/69 SCALE:
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 DWG. 1402



Left Side Elevation



Front Elevation

NOTE:
REFER TO SECTION A-A FOR
DIMENSIONAL HEIGHTS.

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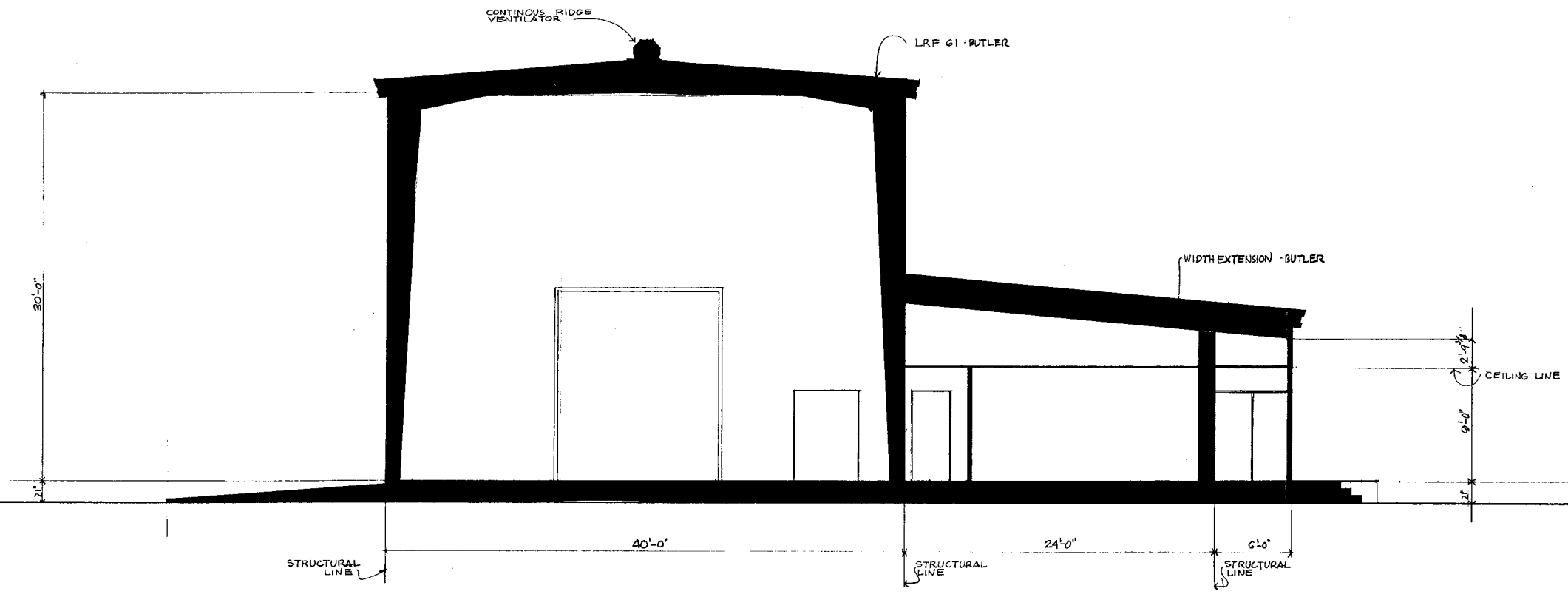
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AOD MINE DRAINAGE TREATMENT PLANT
PHILIPSBURG, PENNSYLVANIA
ELEVATIONS

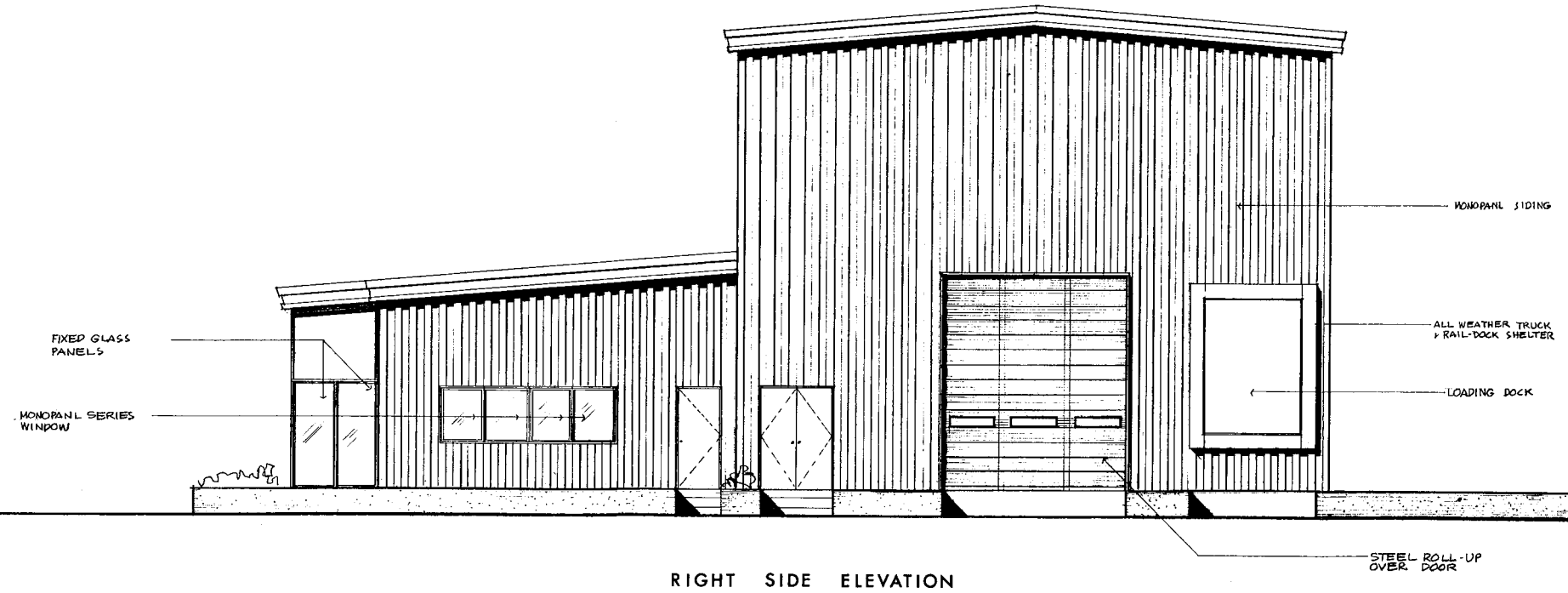
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DEPARTMENT OF MINES & MINERAL INDUSTRIES
HARRISBURG, PENNSYLVANIA

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DATE: 6/27/69
ARCHITECT

W. O. 2674
DWG. 1403



SECTION A - A



RIGHT SIDE ELEVATION

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BURNS AND ROE, INC.
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ACID MINE DRAINAGE TREATMENT PLANT
PHILIPSBURG, PENNSYLVANIA
SECTION & ELEVATION

COMMONWEALTH OF PENNSYLVANIA
DEPARTMENT OF MINES & MINERAL INDUSTRIES
HARRISBURG, PENNSYLVANIA

DESIGNED BY: *W. O. 2674*
CHECKED BY: *[Signature]* DATE: 6-16-69 SCALE: 1/4" = 1'-0"

[Signature]
CHIEF ARCHITECT

W. O. 2674
DWG. 1404

Raw AMD Water Supply System

As previously stated, the three sources of acid water available to the plant, are Acid Mine Hole No. 1, Acid Mine Hole No. 3 and Red Moshannon River at Hawk Run. Rather than construct 3 stations each with a large motor and have 2 stations idle while the third is on line, it was decided to have the main pumping station at Mine Hole No. 1 capable of pumping 475 gpm at a water head of 180 ft through the ion exchangers to the decarbonator and to have the other stations pump to this main one. This arrangement makes pump station No. 1 operate continuously but substantially reduces the power and pipe requirement for the others. The piping to the main station from the other two is a 6" plastic pipe. Plastic was selected because of its internal and external corrosion resistance, its ease of installation and its economy.

The pump sumps at each site are similar. The sumps are set deep enough below the level of the water source to fill the sump by gravity flow. A sluice gate is provided on the inlet to the sump which allows the water level to be pumped down when the gate is closed. A hatch in the top slab of the sump will allow access to the interior of the sump, when the level is down, for inspection and maintenance. To resist corrosion, the pumps are constructed of stainless steel. The pumps are duplicated at each location and equipped with a low level shutoff to stop them if flow into the sump ceases.

The inlet used at the river differs somewhat from those used at the mine holes. At the mine holes, a stainless steel well point is placed in the bottom of the pools. This will strain most foreign matter out of the water before reaching the pumps. This same objective is met at the river by enclosing the intake pipe with a stainless steel mesh. The mesh is welded to a stainless steel pipe frame which is embedded in a concrete slab. The concrete slab also acts as an anchor for the inlet piping.

Ion Exchange

Ion exchange treatment is generally performed in fixed resin beds with the liquid flowing downward through the bed. The pressure drop through such beds is a function of the vertical velocity and the size, gradation and porosity. If the liquid being treated contains any suspended particles, the beds will also act as filters and the pressure drops will increase as suspended particles are trapped in the filter.

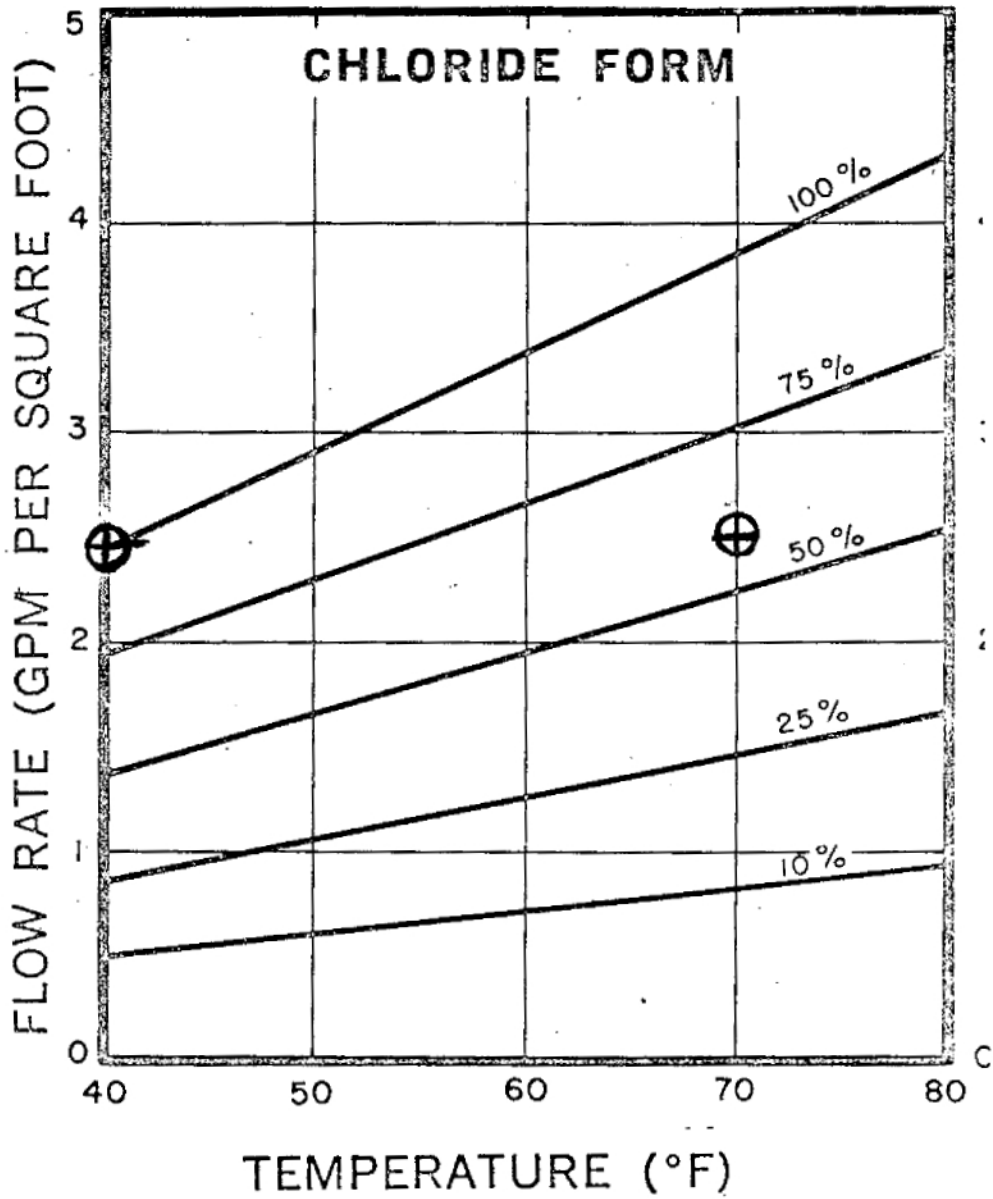
While the mine drainage water to be treated in the proposed plant does not contain any substantial amounts of suspended matter, it does contain several ions which upon contact with the resin may be converted into relatively insoluble compounds. These precipitates would be trapped in the resin bed and would cause a rapid increase in the loss of head through the bed, ultimately plugging the bed completely. They would also tend to deposit on the resin beads and would not be removed during regeneration of the resin. This would cause a loss of exchange capacity as well.

To prevent this condition, it is proposed to operate the ion exchange system in upflow mode. The velocity of flow will be sufficient to expand the bed by about 100% thus continuously flushing out any precipitates that might be forming. The operation of ion exchange beds in the upflow mode requires close control of the flow velocity, since excessive velocity will cause the bed to compact at the top of the vessel and again result in a high pressure drop.

The permissible range of velocity in the upflow mode is a function of temperature density and velocity of flow. Figure 8 shows the expansion characteristics of the proposed resin under a variety of temperature conditions. Using these data for the design of ion exchange vessels an upflow velocity of 2.5 gpm per sq ft is dictated since this velocity will result in a bed expansion of 60% at 40° F and of 100% at 70° F.

The ion exchange system will consist of 4 parallel towers, 1.1 ft diameter by 13 ft straight side. Each vessel will contain, 570 cu ft of ion exchange resin. The resin bed is supported by a false bottom plate which contains fluid distributor strainers. The settled bed depth will

AMBERLITE IRA-93
BACKWASH CHARACTERISTICS



be 6 ft. The regenerant inlet distributor is located just above the settled bed in order that the freeboard space above the settled bed does not fill with excess regenerant solution. A treated water collector is located at the top of the vessel.

The ion exchange vessels will be operated in pairs, with one pair in service while the second pair is in regeneration or standby service. The water to be treated will be pumped into a manifold connected to each of the exchange vessels. The total water throughout will be measured in the pump discharge line. Instantaneous flow to each unit will be determined at the unit inlet. The raw water pH, specific conductivity and total hardness will be monitored at the inlet line.

Ion Exchange Resin

The proposed plant will use Rohm and Haas Company Amberlite IRA-93 ion exchange resin. This resin, a weakly basic anion exchange resin, is approved for use in the processing of potable water supplies by the U.S. Food and Drug Administration. Amberlite IRA-93 was substituted for the IRA-68 resin since the latter has not yet been approved for use in the processing of potable water supplies. The characteristics of Amberlite IRA-93 will show in the following table.

The resin will be utilized in the bicarbonate form. During the exhaustion part of the cycle the resin will exchange sulfates in the mine drainage water for bicarbonates. At the present time it is proposed to operate the plant on a fixed time cycle with only partial exhaustion. The theoretical capacity of the ion exchange system, when supplied with water of the analysis used as the basis of design, will be 820,000 gallons per regeneration cycle, based on a new resin capacity of 24 kg per cu ft. Treating less than this full capacity, will improve the resin efficiency, lower operating costs and provide longer resin life. For the initial operations of the proposed plant the ion exchange units will operate on a 12-hour day cycle.

PHYSICAL AND CHEMICAL PROPERTIES

Ionic Form	
(as supplied)	Free Base
Shape	Spherical Particles
Moisture*	50 to 58 percent after one exhaustion-regeneration cycle
Density*	38 to 42 lbs./cu. ft. after one exhaustion- regeneration cycle
Shipping Weight	38 lbs./cu. ft.
Effective Size	0.40 to 0.50 mm.
Screen Grading	16 to 50 mesh, wet (U.S. Standard Screen)
Uniformity Coefficient	2.0 maximum
Fines Content	1.0 maximum through a 50 mesh U.S. Standard Screen
Swelling	Approximately 23 percent upon complete conversion from the free base to the chloride or sulfate form
Voids	37 to 40 percent
Total Exchange Capacity	
a) Volumetric	1.4 meq./ml, minimum 30.5 kgr./cu. ft. as CaCO ₃ 4.4 lbs.-eq./cu. ft.
b) Weight	3.8 meq./g. of dry resin, min.
Temperature Limitations	212°F. maximum

Regeneration

Regeneration of the resin is accomplished by passing a solution of ammonium hydroxide through the bed. At the start of the regeneration cycle, the pressure in the vessel is released and the water level is dropped to just above the top of the resin bed.

The solution of ammonium hydroxide is then added and allowed to seep through the bed. The exact volume required for each regeneration is determined from the quantity of water processed between regenerations and the anion content of the water. At the design flow and raw water characteristics, 2,960 lbs. of ammonia are required to regenerate all four units on a once per day basis. It is proposed to apply the ammonia in the form of 8% solution of ammonium hydroxide.

The spent regenerant solution consisting of ammonium sulfate and excess ammonium hydroxide will flow by gravity to the ammonia recovery system. Immediately following the application of the regenerant, the resin bed is rinsed with 5,000 gallons per unit of product water. This rinse is required to displace the excess ammonia from the resin bed and is also collected for ammonia recovery.

After completion of the first stage rinse the unit is pressurized to the operating pressure of 50 psi and the second stage rinse introduced through the bottom distributors. This rinse consists of 28,500 gallons per unit of product water to which carbon dioxide has been added in an amount sufficient to convert the resin from the hydroxide form to the bicarbonate form. The second stage rinse water leaves the vessel through the top collector and re-enters the main product water stream. Once a unit has been converted to the bicarbonate form, it is ready to be returned to service. However, in order to avoid loss of carbon dioxide the unit must be kept under pressure until it is put into service.

Ammonia Recovery

The spent regenerant solution and the first stage rinse water are collected in one of two 15,000 gallon tanks which are equipped with stirring paddles. Hydrated lime slurry is then added to the spent regenerant until the ammonium sulfate is completely converted to calcium

sulfate and free ammonia. When the reaction has been completed the slurry is pumped to one of two pressure filters to separate the solids from the liquid. Vertical mounted baffles are required to keep the slurry in suspension and prevent caking on the bottom and sides of the tank.

The pressure filters consist of a series of plates containing filter cloth covered frames. The slurry solution is pumped into the frames where the solids are retained by the cloth and the liquid passes through the cloth into a plate container.

When the frames are full of solids the flow is stopped, the plates are separated and the moist solids are dumped on a conveyor belt. The conveyor carries the solids away from the plant for disposal in the waste lagoon.

The filtrate is pumped to an elevated storage tank which will feed a constant flow to an ammonia stripper still. The filtrate will contain 1.6% ammonia solution. The stripper still will produce an overhead distillate product containing 28% ammonia solution. The still consists of a column containing a stack of perforated sieve trays equally spaced. The perforations are equally distributed across the tray and at one end, there is a liquid downflow area for the lean liquid to flow down to the next tray. The vapor bubbles up through the perforations and become enriched by contacting the liquid.

In order to obtain the desired fractionation of ammonia and water vapor the feed solution is introduced approximately in the center of the distillation column. Ammonia vaporizes and rises, bubbling through the trays to the top of the column. The difference in concentration and vapor pressure of ammonia and water cause the ammonia to boil and vaporize from the liquid solution on the trays. The ammonia-rich vapor at the top of the column enters a condenser. The condenser consists of a nest of parallel tubes fixed inside a tank shell. Cooling water flows through the shell portion, cooling and condensing the ammonia water vapor flowing through the tubes. A portion of the cooled condensed liquid is

returned to the first tray for further enrichment by means of a reflux apparatus. The remainder of the liquid flows to a storage tank.

The ammonia stripped water flows downward through the lower trays where further ammonia stripping occurs. At the bottom of the column, the liquid water enters a reboiler, which is similar to the condenser, except that steam flows through the tubes heating and boiling the residual water product in the shell area. The ammonia-water vapor mixture is introduced back into the bottom of still column and the remaining liquor is drawn off as a bottom waste product. As an alternate heat source, raw steam may be introduced into the bottom of the still column.

The 28 percent ammonia distillate product is collected in a 1500 gal storage tank. It is then transferred to a day tank, where the ammonia solution is made up to the required 8% strength. It is estimated that the addition of 160 lb ammonia and 3,210 gal of fresh water will be required to produce sufficient ammonia for the daily regeneration requirements.

Decarbonation, Aeration and Settling

The purpose of the Decarbonation and Aeration section is to remove from the process water the alkalinity and the heavy metals, iron, aluminum, and manganese. The heavy metals are present in the raw water and the alkalinity was introduced in the ion exchange step by exchanging sulfates and chlorides for bicarbonates on the resin; since there is a direct ion substitution, the bicarbonate concentration is the sum of the chlorides and the sulfates in the raw water. As discussed under Water Analyses, for design purposes a high value of 1000 mg/l sulfate as CaCO_3 was selected. The iron and aluminum along with the other cations, are unaffected by the processing step and their concentrations are those of the original raw water. The representative heavy metal concentrations are 250 mg/l Fe as CaCO_3 and 100 mg/l Al as CaCO_3 . The pH of the pressurized ion exchange effluent is unknown but laboratory data indicate that immediately after pressure is released, the pH will be between 5.-6.

The removal of alkalinity and heavy metals from the influent water occurs as follows: The water from the ion exchange passes through a pressure reduction and control valve leading to the elevated spray decarbonator. The 50 psi pressure that is initially required to keep the desired CO₂ concentration in solution is utilized for two other purposes: (1) to lift the water to an elevation suitable for gravity flow through the remaining plant and (2) to degas the water by spraying it through venturi nozzles and thereby making available for decarbonation a substantial water surface.

This spraying step involves two gas transfer operations, the removal of CO₂ and the saturation with O₂, each of which depend on the gas concentration difference between two phases, surface area and temperature. The principal function of the spraying however is to decarbonate i.e. remove CO₂ from water, and thereby raise the pH to 8.0 - 8.3. The CO₂ removal is necessary to decrease the quantities of lime used in the softener step and to increase the pH to an acidity range in which the oxidation of ferrous iron occurs at a substantially more rapid rate. The ferric hydroxide formed in this reaction precipitates out of solution along with the aluminum floc.

The work of Stumm & Lee, "Oxidation of Ferrous Iron"^(A) shows that the iron oxidation time is a function of the original ferrous, hydroxyl, and oxygen concentrations and the water temperature. Kinetic calculations based on this reference indicate that the oxidation at normal ambient temperatures can be reduced by a factor of 10, from 50 to 5 minutes by increasing the pH from 7.5 to 8.0. The second function of the spray is to introduce oxygen into the water in order to obtain the highest possible oxidation rate for ferrous iron. Complete oxidation of the iron requires 20 mg/l of oxygen and saturated water contains only 10 mg/l. Since the oxygen requirement is twice the amount available in saturated water, the initial oxygen in the water must be first consumed in the oxidation reaction and then the water must be resaturated at least once.

^(A) Stumm, W. and Lee, G.F., Ind. Eng. Chem. 53:143 (1961)

Spray decarbonation can only partially saturate the water because the exposure time of droplets in air is too short, and can only cause partial oxidation of the iron because the detention time in the spray collection sump is also too small. Additional exposure and detention time is provided in a second aeration step which consists of a slat filled, forced-draft tower and two settling (detention) basins. This second step is gravity fed by the overflow from the collection sump to the distributor on top of aeration tower. As the water cascades down through the tower over the surfaces provided by the 45°-mounted, wooden slats, it is continually exposed to a fresh supply of upward flowing air fed in at the base by 500 CFM blower. The aerator is 22 ft high with 30 sq ft cross sectional area and is located between the two open-top detention basins. The oxygen-saturated service water flows from the aeration tower by gravity into these basins where the reaction goes to completion; the newly-formed ferric hydroxide and aluminum hydroxide precipitate out of solution at this point and settle to the bottom of the tank. The settled sludge is scraped to a collection point with a rotating rake. The sludge is discharged from the basin to the waste lagoon by "blowing down" in the usual manner i.e. the same as the softener sludge. The basins are capable of alternate or simultaneous operation, and can be "blown down" continuously or intermittently as operating conditions permit. Each detention basin is 40 ft in diameter, eight ft high and holds 75,000 gallons, equivalent to two and a half hours detention time. This reaction period is a conservative estimate of the time required for oxidation of the 140 mg/l iron to 1.4 mg/l and for the settling of the iron and aluminum sludge during average summer temperatures. For cold weather service when both the reaction rate and settling is slow, the detention period can be doubled to five hours by operating both basins simultaneously in parallel.

The overflow from the detention basins is fed to the softener where the remaining hardness will be reduced to 70-100 mg/l. The softener is a protection against leakage passing through the aeration system.

The importance of removing the aluminum in the aeration step should be pointed out. This MD water has an unusually high manganese content (15-30 mg/l) which must be reduced to 0.05 mg/l to make the water potable. The simplest means is to oxidize the manganous ion to the manganic and to allow this hydroxide to precipitate at a pH = 10. The high pH and therefore the precipitation of manganese are assured because, fortunately, the softener normally operates at a pH of 10.0-10.5 in order to remove the specified calcium and magnesium hardness. However, at this high pH, aluminum is quite soluble (21 ppm) and any aluminum present in the sludge blanket of the softener would redissolve and pass into the recarbonator. In the recarbonator the pH is lowered to 8.0 - 8.5 by introduction of carbon dioxide. The solubility of aluminum is limited at this pH and any excess over 3 mg/l will precipitate out and plug the filters or collect in the ground storage tank. It is therefore advantageous to remove the aluminum in the aerator step along with the iron and avoid the reprecipitation problems.

Softening

In order to reduce the hardness of the final product water, the effluent from the settling tanks will flow to a solid contact type lime softening unit. The following operations are accomplished in this unit, mixing of the water and the lime, flocculation, and separation of floc and supernatant. The unit is of the upflow, variable cross section, sludge recirculation type.

Complete contact between the feed chemicals and the aerator effluent occurs in a central mixing zone in which the flow is upward. After mixing, floc is formed in a downflow annular coagulation zone before entering the upflow clarification zone. As the water rises, in the clarification zone, the cross sectional area increases and the rise rate decreases. As the rise rate is reduced to a value where the force of gravity exceeds the upward-acting forces on a floc particle, the particle no longer rises and is separated from the flow. These particles increase in size and weight until they settle down at the bottom

of the clarification zone. Part of the sludge is recirculated into the mixing zone to assist the formation of new floc. The remaining sludge is removed or blown off periodically.

The design of the unit is such that the rise rate, as calculated in the clarified water zone immediately below the effluent collector, does not exceed 1.0 gpm/sq ft at the design net flow of 475 gpm. The gross flow is somewhat greater to account for sludge blow-off and filter backwash. Although this rise rate is lower than that which is normally encountered in a softening unit, it was selected to assure that there would be no floc carryover at the relatively low operating temperature (5° C) which may be encountered.

The solids contact unit will have a diameter of 25 ft and a side height of 15 ft. It is estimated that a lime dosage of 250 mg/l will be required to reduce the hardness to 70 mg/l.

The underflow from this unit will be primarily calcium carbonate. Present plans are to discharge this sludge to the waste lagoon. However, it would be feasible to recalcine this sludge and produce both lime for reuse and carbon dioxide for recarbonation. The technology for recalcining of calcium carbonate sludges is well established. However, the economics for a plant of the size contemplated at Hawk Run are not favorable.

The use of carbon dioxide from recalcining for carbonation of the resin would require further processing to separate carbon dioxide from the gas stream and compress it to the pressures necessary for resin regeneration.

Re carbonation

Recarbonation, or the addition of carbon dioxide (CO₂), is done to prevent the calcium carbonate present from precipitating as scale in the filter bed or in the piping. The CO₂ added converts the calcium carbonate to the bicarbonate form and lowers the pH. The end point is set at a pH range of 8.0-8.5 since at a lower pH, the water becomes aggressively corrosive.

An added benefit in this case is that any aluminates present in the softener effluent will precipitate as alumina. This can be removed in the filters.

Filtration

Three rapid sand filters will be provided, each designed to operate at a service rate of 2 gpm/sq ft when all units are in service. The 2 gpm/sq ft rate is within the normal operating range for this type of filter. When a unit is removed from service for backwashing, the remaining units can process the entire flow from the softener.

The filter media will consist of two layers. The top layer consists of 24 inches of 1 mm diameter anthracite. The bottom layer will consist of 12 inches of 0.4 mm diameter sand. The anthracite, being coarser, has larger pores, allowing the turbidity to penetrate further into the bed to achieve longer filter runs. The sand is used to filter out the finer material which may pass through the anthracite. Separation of the beds is maintained after backwash by the difference in specific gravities. The sand, being heavier, settles before the anthracite.

Eventually, the pores will become clogged, and the pressure loss across the filter will rise. When the pressure drop reaches 3-4 psi, the unit is backwashed. Backwashing is done by reversing the direction of flow through the unit. Wash water enters the underdrain system where it is uniformly distributed throughout the filter cross section. A wash rate of 10 gpm/sq ft of filter area is employed. This rate provides sufficient bed expansion and scouring to remove the entrapped particles and is low enough to prevent the media from being washed out. The wash

water then flows into the wash troughs and out of the filter to a basin. Filtered water for backwashing comes from the ground storage tank. The duration of the backwash is ten minutes. After backwashing, the filter is rinsed with softener effluent flowing in the service direction for ten minutes at 2 gpm/sq ft. However, the filter effluent is recycled to the softener via a backwash collection sump to conserve the treated water. After rinsing, the unit is returned to service.

Waste backwash will be returned to the inlet of the solids contact unit.

The backwash cycle will be semiautomatic. Upon reaching a desired head loss, the operator will start the backwash cycle. The filter then will automatically initiate first, a surface scouring, then a low rate of backwash, followed by high backwash rate and another low rate period to settle the filter bed. Restoration to service will be manual.

Product Storage

After the water has been filtered, it flows, by gravity, to a storage tank. This water is used within the plant for regeneration dilution water, slaking and feeding of lime, and filter washing. When the plant is developed into a source of potable water, the tank can provide equalizing storage. Such storage allows the service demand to fluctuate while the plant operates at a constant rate.

The tank has a capacity of 180,000 gallons, is 40 ft in diameter and is provided with an inlet connection, two outlet connections, an overflow and a drain. One outlet is connected to the pump for inplant use and the other is used as the plant effluent connection.

The ultimate goal of this plant is a potable water to be used as a municipal supply. Therefore, the tank is covered to prevent contamination by dust, rain, insects and birds. The cover is also a safeguard against pranksters and vandals. Both the overflow and the vent are designed to prevent contamination.

Waste Collection

The sludge lagoon collects all liquid and solid waste materials discharged from the treatment process, allows the solids to settle out of solution and allows a small volume of neutral, supernatant liquid to overflow into Hawk Run. The wastes from individual steps are conducted from the several discharge points to the lagoon via an underground sewer network.

The lagoon covers an area of 2.6 acres, is five ft deep and is located in the low land north of the access road and east of Hawk Run Creek. This land was selected because it requires the minimum amount of earthwork to attain the proper elevation for gravity flow from the plant, to construct the earth dam between the lagoon and Hawk Run and to form the diversion dam. This land is not part of the site donated by Lewis Stein and must be purchased.

Two operations in the plant form and discharge heavy precipitates. The softener blows down 300,000 lb/day of waste water containing primarily calcium carbonate and magnesium hydroxide. The settling basins following the aerator blow down 45,000 lb/day of waste water containing hydroxides of iron, aluminum and manganese. Sludge from these units is less than 5% solids by weight.

In addition to these units, additional sludge and waste water are produced at the ion exchange regeneration station. The ammonia stripper still bottom waste water contains calcium carbonate, calcium sulfate and less than 0.1% ammonia. The treated spent regenerant is dewatered by filtration. The filter cake produced, chiefly calcium sulfate, is 47% solids by weight.

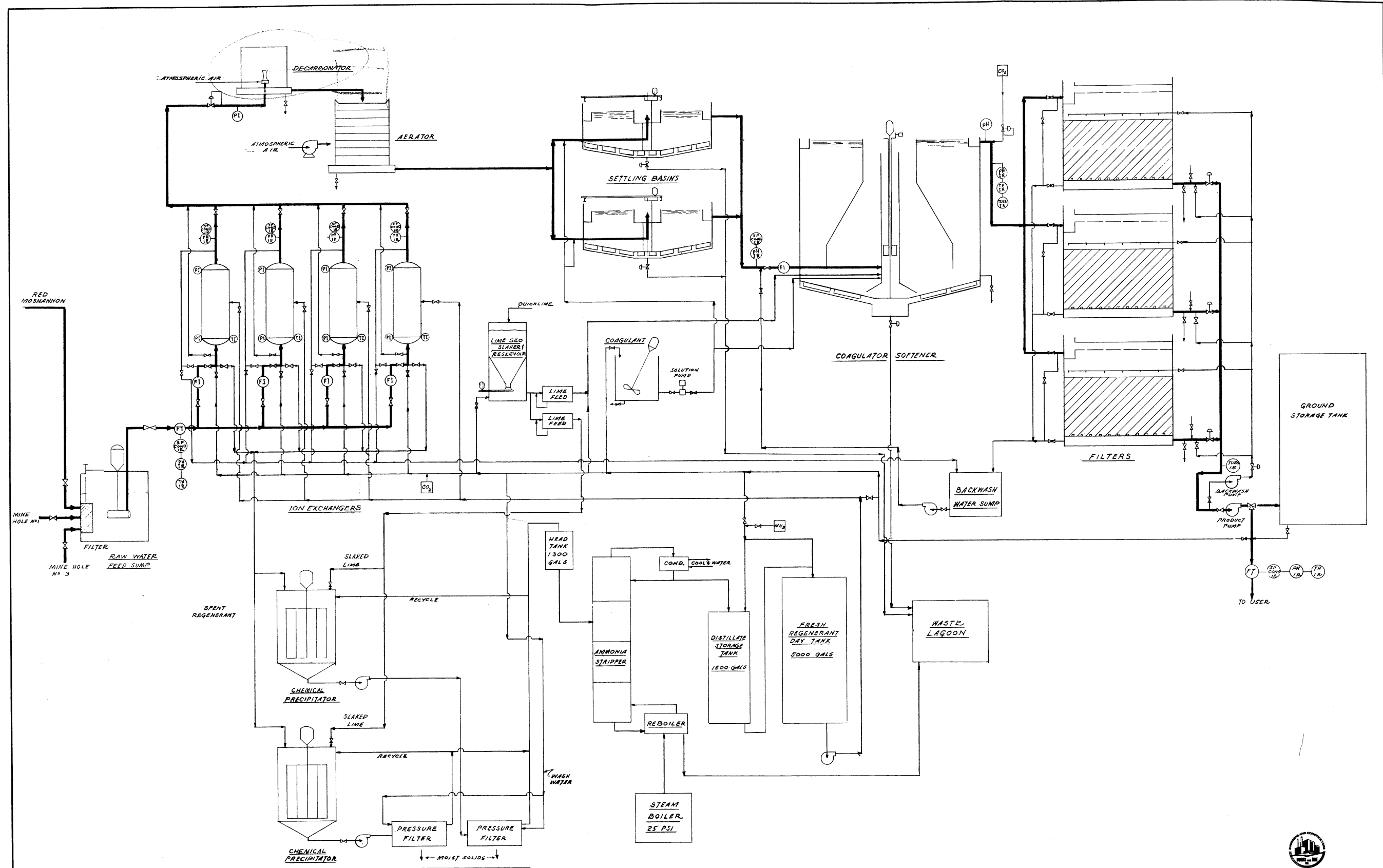
-With the exception of the filter cake, these wastes can flow by gravity through sewers to the lagoon. The sewers are designed so that liquids mix with the filter cake and resulting liquid velocity is great enough to carry the filter cake to the lagoon. At the point where the cake is dropped into the mixing sump a hose connection is provided to clean the sump and flush the sewer as required. The hose connection will be tapped into the raw water line feeding the ion exchangers. Tank

drains and overflows and building floor drains will also discharge into the sewer flowing to the lagoon.

The solids entering the lagoon are relatively insoluble. After they settle out of solution, the water, although somewhat hard, can be discharged to Hawk Run without contributing acidic pollution to the stream. The pond provides a detention time of greater than two months. This is ample time for the solids to settle. It will be necessary to drain periodically the lagoon and remove the settled solids with earth moving equipment. It is estimated that the lagoon will fill with solids in 6 years.

Process Flow Diagram

Drawing 1201 is the detailed engineering flow diagram showing the treatment plant, the regeneration system and waste disposal system.



BURNS AND ROE, INC. ENGINEERS AND CONSTRUCTORS ORADELL, N. J. HEMPSTEAD, N. Y. LOS ANGELES, CALIF.									
ACID MINE DRAINAGE TREATMENT PLANT PHILPSBURG, PENNSYLVANIA FLOW DIAGRAM									
COMMONWEALTH OF PENNSYLVANIA DEPARTMENT OF MINES & MINERAL INDUSTRIES HARRISBURG, PENNSYLVANIA									
DESIGNED BY W. J. G.		DRAWN BY J. L. P.		CHECKED BY J. L. P.		DATE 6/25/69		SCALE AS SHOWN	
REV. BY C. S.		DATE 7/1/69		REV. NO. 1		REVISION CORRECTED		BY C. S.	
ENG. DES. W. J. G.		ELEC. ENG. W. J. G.		MECH. ENG. W. J. G.		CIVIL ENG. W. J. G.		DATE 6/25/69	
W. O. 2674 DWG. 1201									

Composition of Process Streams

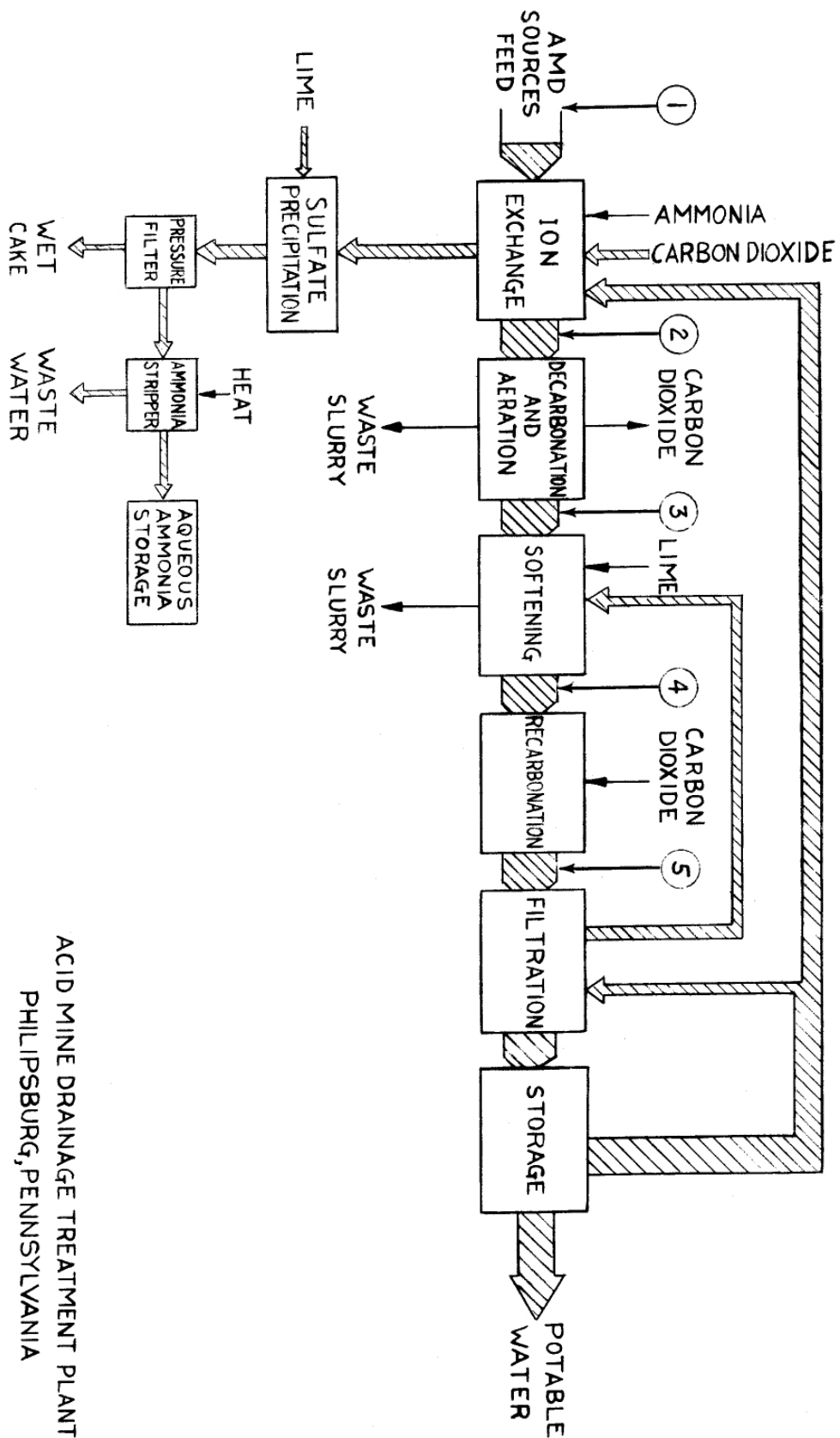
Table 3, Estimated Water Quality Throughout Treatment Process, shows the composition of the effluent from the five critical steps of the process. The cation and anion concentration are listed as mg/l (ppm) as CaCO₃.

The encircled numbers on Table 3 coincide with those on the associated schematic flow diagram, Figure 9.

Table 3
Estimated Water Quality
Throughout Treatment Process

Substance (mg/l as CaCO ₃)	① Raw Water	② Ion Exch. Effluent	③ Aerator Effluent	④ Softener Effluent	⑤ Product After Recarbonation
<u>Cations</u>					
Calcium	300	300	300	14	14
Magnesium	250	250	250	55	55
Iron	250	250	0	0	0
Manganese	30	30	22	0	0
Aluminum	<u>100</u>	<u>100</u>	<u>0</u>	<u>0</u>	<u>0</u>
Total Cations	930(A)	930	572	69	69
<u>Anions</u>					
Bicarbonate	0	930	572	0	61
Carbonate	0	0	0	56	8
Hydroxide	0	0	0	13	0
Sulfate	1000	0	0	0	0
Chloride	<u>10</u>	<u>0</u>	<u>0</u>	<u>0</u>	<u>0</u>
Total Anions	1010	930	572	69	69
Total Hardness	550	550	550	69	69
<u>Alkalinity</u>					
M.O. Alkalinity	0	930	572	69	69
Phen. Alkalinity	0	0	0	41	4
CO ₂ PPM as CO ₂	0	35	15	0	0
Turbidity JTU	<50	<50	<50	<5	<5
pH	3-4	5.5-6.0	7.9	10.5-10.6	8.5

Note: (A) Hydrogen ions not included in Total Cations.



ACID MINE DRAINAGE TREATMENT PLANT
 PHILLIPSBURG, PENNSYLVANIA

SCHEMATIC OF PROPOSED TREATMENT PLANT

DATE 5-30-57
 FIGURE 9

M. O. 2674 - 01

Material Balance:

The material balance (Table 4) shows the quantities of material consumed in treating the designated volume of AMD water. The nominal daily output of the plant is 500,000 gallons while operating by standard water treatment criteria and under the most adverse conditions. The total volume of AMD water that must be treated by the plant under these conditions is 684,000 gallons, and this amount of feed is the design volume of the plant. As noted previously, it is expected that the water-use efficiency will be increased by recovering most of the internally used product water which normally would be discharged to waste. The water that will be recovered and recycled to the process will come from ion exchange rinse, filter backwash, and ammonia recovery. Also, the volume of water used will be reduced by increasing the ammonia concentration used for resin regeneration from 4% to 8% and by cutting the volume of the second rinse to about one-half the water required in the standard ion exchange regeneration (50-75 gal/cu ft resin). All of the above water conservations have been included in the design of the plant. The result is an expected output of 620,000 gallons potable water which is 24% above the nominal or minimum volume. The maximum output will depend on influent water composition and will be determined only through actual plant operation.

Under the design conditions the plant consumes 6-1/2 tons of chemicals and discharges 265 tons of waste slurry. The amount of 1,200 pounds of carbon dioxide gas is vented to the atmosphere. The waste slurry contains 9-1/2 tons of solids, mainly calcium carbonate and sulfate, which are separated from the liquid in the waste lagoon. The supernatant, clear and neutral, will be discharged to Hawk Run.

The Desal Process is distinguished by the use of volatile carbon dioxide which replaces the sulfate and chloride anions on the AMD waters. Upon release of pressure the carbon dioxide leaves the solution. Three tons of carbon dioxide are daily consumed in the bicarbonation of the

ion exchange resin. About two hundred pounds are added to the effluent of the softener to drop the pH from 10.5 to 8.5 and stabilize the softened water.

Carbon dioxide will not be reclaimed in this phase of the project. Although potentially economically rewarding, a carbon dioxide recovery system would add to the complexity of operating the plant and would not assist in the process evaluation which is the primary purpose of the installation. Therefore, carbon dioxide will be purchased in the most convenient form, as liquid in cylinders, and will be discharged to the atmosphere. Alternate possible cheaper sources of CO₂, such as exhaust gases from coal or oil combustion or from lime sludge recalcining will be investigated for the plant expansion to 1 million gpd.

Similarly, lime will not be recovered. Over three and a half tons quicklime 95% pure will be consumed daily and discharged to the waste lagoon. In fact the total amount of calcium discharge from the plant will be larger than that which entered as lime. About 500 lb of calcium from the AMD water feed will be precipitated in the softener as hardness and will be sluiced to the waste lagoon.

The ammonia inventory circulating in the regeneration section of the plant is about 3000 lb. The daily makeup rate is expected to be about 160-200 lb due mainly to the losses in the still-bottoms discharged from the ammonia stripper. Small amounts will also leak from the mechanical seals of pumps and precipitator agitators.

A small boiler is provided to produce 25 psi steam for stripping ammonia from waste water, for vaporization of the liquid carbon dioxide and ammonia and for heating the facility. The steam demand is about 4000 lb/hr and the oil consumption of 35 gallons per hour necessitates a minimum storage tank of 12,000 gallons. Oil is the fuel recommended for this installation because of small operator attention needed and its ease of handling. The average daily consumption is approximately 350 gallons.

Table 4
Treatment Plant Through-put

Daily Input

Feed: Raw AMD water	684,000 gal
Carbon Dioxide	6,238 lb
Lime, 95% CaO	6,430 lb
Ammonia, Anhydrous	160 lb

Daily Output

Product water	620,000 gal
Carbon Dioxide	1,190 lb
Sludge from softener and aerator	345,000 lb
Still bottoms	184,000 lb
Ammonia	160 lb
Solid in sludge	19,160 lb

Hydraulic Profile

The plant was designed to utilize gravity flow wherever possible. The hydraulic profile, Figure 10, shows the energy gradient of the process flow at normal conditions. The datum (zero elevation) is taken at the surface of Acid Mine Hole #1. From this diagram, it is seen that gravity flow of the process stream is feasible from the decarbonator to the ground storage tank. The sludge sewer to the lagoon is shown as a dotted line.

At Mine Hole #1, the pumps develop a head of 180 ft of water.

The pipe and fittings leading to the inlet valve of the ion exchange column have caused a head loss of 12 ft. The head loss through the ion exchanger is 12 ft and 4 ft more are lost in the piping to the decarbonator. The head at this point is 152 ft. This would require a unit with high walls to prevent the wind from carrying the spray away. By installing a pressure relief valve upstream from the spray nozzle, sufficient head is dissipated to contain the spray within the decarbonator. Upon exposure to atmospheric pressure, the water is now dependent on gravity for the pressure differential.

Therefore the equipment following the decarbonator is positioned such that the water surface of each unit is lower than that in the preceding equipment by an amount equal to the head loss in the interconnecting piping. The topography of the site is such that this arrangement can be made without a great amount of earthmoving.

W.O. NO. 2674-01

BURNS AND ROE, INC.

DATE: 6-24-69

DRAWING NO. _____

ORADELL, N. J.

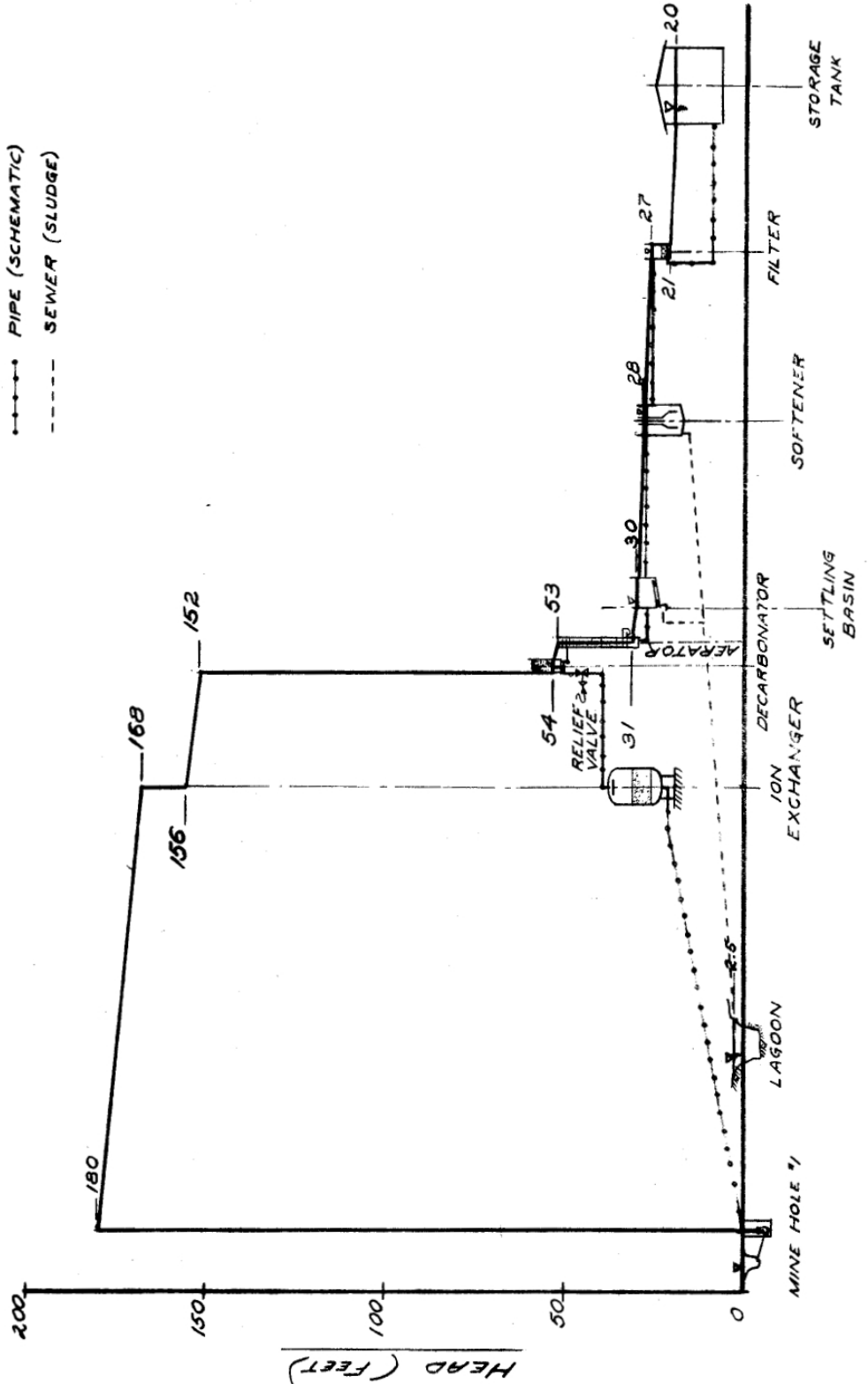
BY: CRM

TITLE: HYDRAULIC PROFILE

SHEET _____ OF _____

HYDRAULIC PROFILE
AMD DEMONSTRATION PLANT

- LEGEND
- HEAD
 - PIPE (SCHEMATIC)
 - - - SEWER (SLUDGE)



HOR. SCALE 1" = 120'
EQUIPMENT

Plant Operation and Manning:
Schedule Explanation

The Operating and Manning Schedules, Figures 11 and 12, dictate the plant operation over a 56 hour time period. The plant will operate 7 days per week, 24 hours per day and the schedules assume it is at a steady state condition. The Operating Schedule depicts the plant's major components, such as ion exchange aerator, softener, gravity filters, and Chemical Precipitation System; it shows their operating periods, and their sequence with respect to the overall plant cycle. The underlying factor, for the sequence of operations as presented here, is manpower consideration. The anticipated manning requirements for plant operation only are indicated in the Manning Schedule. The scale of both schedules is one hour per division.

The first two lines of the Operating Schedule show the ion exchange system which is separated into Unit No. 1 and Unit No. 2, each unit being composed of 2 ion exchange towers. The time required for one cycle in each ion exchange unit is 12 hours, the cycle lasting in Unit No. 1 from noon to midnight and in Unit No. 2 from midnight until noon. All regeneration is done during the day shift. In the regeneration cycle of Unit No. 1 (indicated by cross hatching) one tower is regenerated from 8 A.M. to 10 A.M. and the other tower is regenerated from 10 A.M. to 12 A.M. At noon the newly regenerated Unit No. 1 is put into service and Unit No. 2 is regenerated with one tower finishing at 2 P.M., and the other tower at 4 P.M. Unit No. 2 is now fully regenerated and is kept on standby for 8 hours until midnight at which time ion exchange service begins in Unit No. 2 as Unit No. 1 comes out of service. Thus ion exchange is carried out at all times and regeneration for the whole system takes place from 8 A.M. to 4 P.M. during the prime manpower period. Following ion exchange on the schedule we have gravity filter backwashing. The three gravity filters in the plant are backwashed and rinsed for approximately 20 minutes, once every 24 hours. They are depicted as being backwashed at 8:50, 9:50 and 10:50 but the exact time is dependent on filter backpressure. Filter servicing takes place during the day shift

when maximum manpower is available. It should also be noted that the gravity filters are located in the same general area of the main plant building as the ion exchange towers making them readily accessible to the plant operators; furthermore, the times indicated are guide lines and there is some flexibility with regard to scheduling the filter backwash.

The continuous bars on lines 4 and 5 indicate that both aeration and softening steps are occurring continuously. Lines 6, 7 and 8 depict the slaking operation. The slaker operation is represented by cross hatching and its duration is 8 hours per day from 4 A.M. to noon. This period allows for a four-hour buildup of lime slurry in the 5000 gallon reservoir prior to transfer into each of the two chemical precipitator tanks. The continuous line represents the activity of the slurry pump feeding lime to the softener. This pump operates continuously since the softener is in constant operation. Line 8 shows the timing of the transfer pumps. These pumps operate for 30 minutes twice a day at 7:30 A.M. and 11:30 A.M., and deliver the batches of lime slurry to the chemical precipitator tanks.

The Chemical Precipitation System represented by lines 9 and 10, is divided into Tank No. 1 and Tank No. 2 and corresponds to the regeneration of Units No. 1 and No. 2 shown for the two ion exchange units., At approximately 8:00 A.M. Tank No. 1 begins to fill with the spent regenerant from first tower of ion exchange Unit No. 1. Prior to this addition, a batch of lime slurry sufficient to neutralize all the regenerant from Unit No. 1, has been added to Precipitator Tank No. 1. Two hours later the remaining portion of spent regenerant, from the second tower of Unit No. 1, is added to Precipitator Tank No. 1 and fills it. The agitators in the precipitators operate whenever there is slurry in the tank in order to keep the solids in suspension and prevent scaling. The gypsum and lime slurry in Precipitator Tank No. 1 is then pumped through the plate and frame pressure filters where the solids are separated from the liquid. Meanwhile, the spent regenerate from ion exchange Unit No. 2 begins to fill Precipitator Tank No. 2, to which lime slurry from the slaker has already been added. The mixing continues in Precipitator Tank No. 2 from noon to 4:00 P.M. at which time its precipitated slurry is fed to the pressure filters.

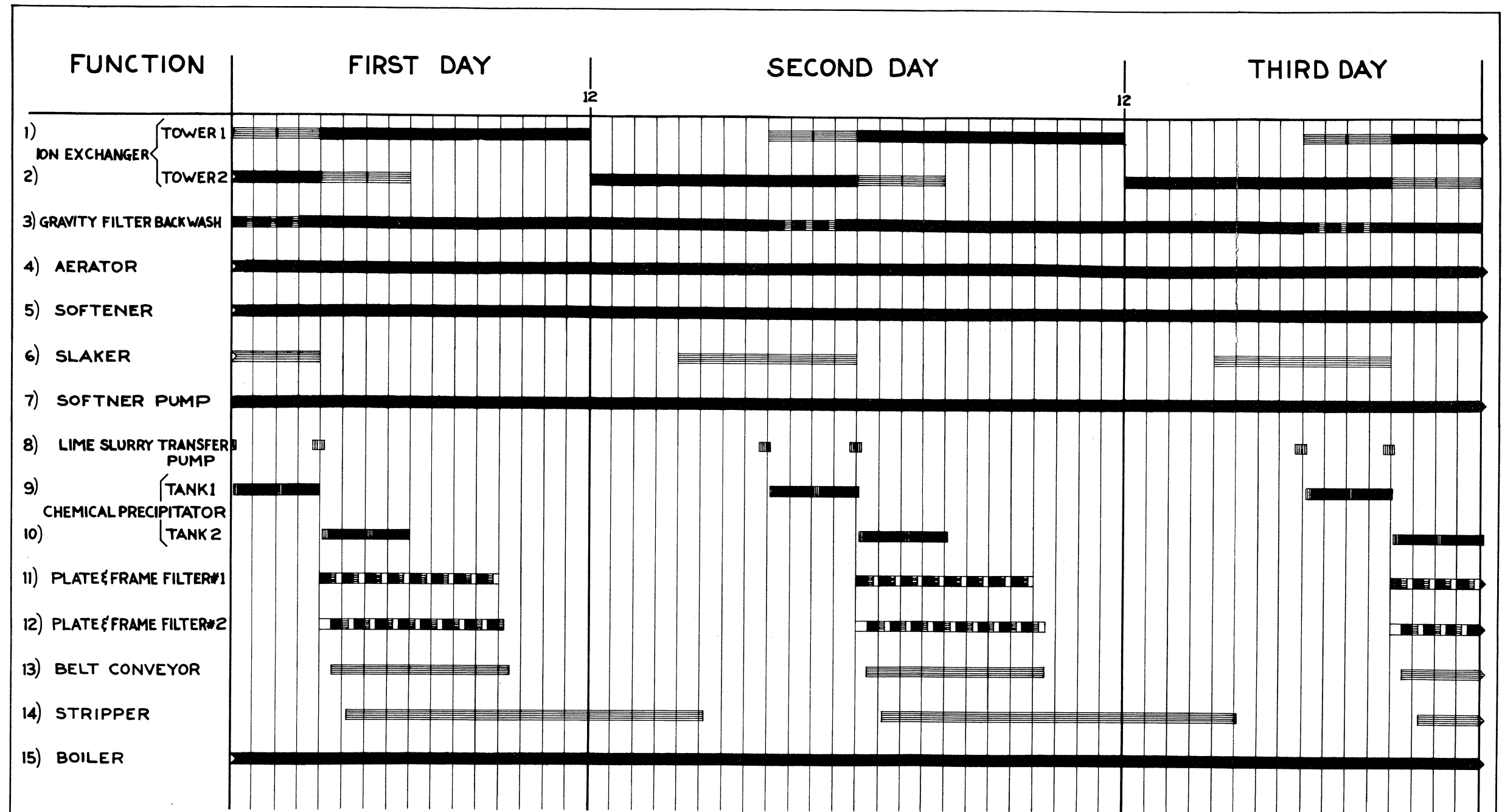
The two pressure filters which are mounted in tandem are depicted on lines 11 and 12; at noon Filter No. 1 begins to fill with the solids from the slurry pumped from Precipitator Tank No. 1 and after 30 minutes Filter No. 1 is filled. The slurry is then diverted to pressure Filter No. 2 for 30 minutes while Filter #1 is cleaned (represented by cross hatching) for 15 minutes, and then shutdown until 1:00 P.M. At this time it again receives slurry while Filter No. 2 is being cleaned. This procedure of alternate filter operation continues until 4:00 P.M., at which time Precipitator Tank No. 1 is empty, and Precipitator Tank No.2 is serviced in the same manner until empty at 8:00 P.M.

The 13th line shows the operation of the belt conveyor which is used to transport to the sewer sump the moist solids removed from the pressure filters. The solids are sluiced from the sump to the waste lagoon with waste water. The conveyor begins to operate at 12:30 P.M. as the moist solids from the cleaning of pressure Filter No. 1 begin to deposit on the belt, located below the filters. The conveyor will operate continually until 8:30 P.M. when all the moist solids have been deposited into the sewer sump.

Line 14 represents the operation of the ammonia stripper still. The still operation begins at 1:00 P.M. and continues for 16 hours, ending at 5:00 A.M. The filtrate and washings from the pressure filters begin to enter the 1300 gallon head tank at noon and by 1:00 P.M. there is sufficient head to open the valve leading to the still. At 5:00 P.M. the fresh regenerate tanks begin to fill, and the regenerate for the following day is ready for use.

The boiler which services the ammonia stripper and provides heat for the building is shown on line 15 and operates continuously.

The remaining four lines (16 through 19) deal with the manning schedule for the plant. For 24 hour day operation the plant employs five operators, a maintenance man, a supervisor, and a chemist. Three operators will work during the principle work shift (8:00 A.M. to 4:00 P.M.) with the 2 additional operators covering as main operators for the other two 8-hour shifts, the afternoon shift (4-12 M) and the midnight shift (12 M8:00 A.M.). It is advisable that the supervisor work 9:00 to 5:00 P.M.



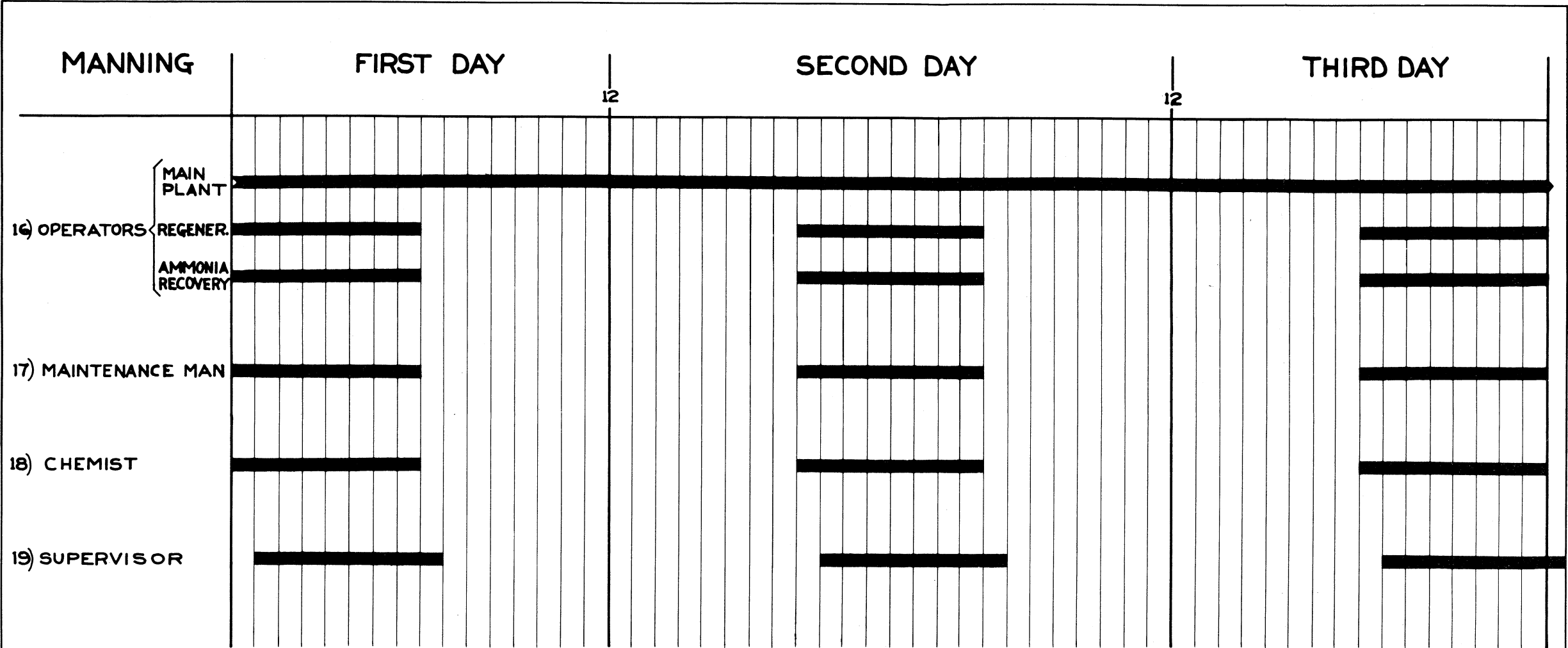
ACID MINE DRAINAGE TREATMENT PLANT
PHILIPSBURG, PENNSYLVANIA

OPERATING SCHEDULE

DATE 6-30-69

FIGURE 11

W.O. 2674-01



ACID MINE DRAINAGE TREATMENT PLANT
 PHILIPSBURG, PENNSYLVANIA

MANNING SCHEDULE

DATE 6-30-69
 FIGURE 12

W.O. 2674-01

in order to oversee the major operations during the day, and the transition to the 4 to 12 work shift. Line 16 depicts the work day for the operators, during the prime work shift (8:00 A.M. - 4:00 P.M.) there are three men present; one will oversee the regeneration in the ion exchange units, another operator will conduct the Chemical Precipitation, and ammonia recovery for the following days regeneration. The third, the main plant operator monitors and controls the whole water treatment plant and is based at the control panel located adjacent to the ion exchange towers.

Line 17 represents the maintenance man's work day and line 18 shows the chemist activity. Finally, line 19 depicts the supervisors schedule showing the aforementioned phase shift in his work day.

PLANT EXPANSION TO 1,000,000 GPD

The proposed Hawk Run AMD treatment plant has been designed using very conservative criteria to produce 0.5 million gallons of potable water per day. The initial design has also been made with consideration for the probable future expansion which will boost the output to 1.0 million gallons per day.

The following design for the expansion has been made assuming the laboratory data concerning certain water efficiencies provided by Rohm and Haas is substantiated in the initial plant. The water conservation, as used in the initial process design, allows for a production increase of about 25% in excess of the nominal half million gpm rating and therefore allows the expanded plant to double the nominal output without doubling the plant. Thus, instead of treating 1.37 million gpd (twice the input of the original plant) the expanded plant will only require a daily input of 1.11 million gallons to yield the desired 1 million gallons. Each of the three sources normally overflow the input amount, although, during summer drought conditions, flow from Hole No. 3 may be below it and may not be usable.

In the preliminary design of the expansion it is proposed to minimize cost by adding only three ion exchange towers and operating with an unbalanced service flow during a quarter of the service cycle. In order to limit the hydraulic problems all seven ion exchange towers will be of the same diameter. In a daily service cycle, four towers will treat 950 gpm for 6 hours and three towers will treat 712 gpm for 18 hours. Six towers will be regenerated once every 24 hours and the seventh tower will require regeneration once every 48 hours. Sufficient feed to the towers will be provided by adding a 250 gpm pump and a spare to the two existing 475 gpm pumps in the main feed sump. Combinations of these pumps will yield the required 950 gpm and 712 gpm flows. The decarbonation and aeration steps following ion exchange may require additional capacity as experience in the initial plant will dictate. At the present time the

extreme condition is considered. Additional nozzles and storage capacity will be installed in the existing decarbonator structure and an additional forced-draft aerator will be positioned parallel to the first. Additional settling basins will probably be required to separate the iron, aluminum and manganese from the new higher flows. At least one additional basin will probably be required just to smooth out the variations in flow due to the uneven service through the ion exchange section. Two basins are recommended. An additional softener will be needed in parallel to the original one in order to keep the rise rate constant. The recarbonation will be increased by increasing the carbon dioxide flow through the existing CO₂ feed nozzles. Two additional gravity filters will be required to comply with the filtration standard of 2 gpm/SF for a municipal water supply. A chlorination station will also be required. No increase in storage capacity is necessary.

The expanded plant will reclaim ammonia, carbon dioxide and lime. The ammonia recovery will be increased by one 15,000 gallon chemical precipitator, an additional slurry feed system, extension rods to hold additional plates and frames for the existing pressure filters, and an additional ammonia stripper system that has one half the capacity of the original system.

Lime will be reclaimed by feeding the blowdown containing calcium carbonate and magnesium hydroxide sludge from the softeners to a recalcining furnace which is located adjacent to the softeners. The resulting lime from the furnace will be conveyed to the silo for reuse and the CO₂ exhausted from the furnace stack will be collected.

It is estimated that 50% of the CO₂ fed to the plant can be reclaimed from the softener wastes. The CO₂ collected will be compressed to about 75 psi and reused to regenerate the ion exchange resin.

A 24 ft extension to the 30 ft high bay will be required to house the three new ion exchange towers and two new gravity filters. The following list is the equipment required for the plant expansion. A plot plan of the expanded 1 million gpd plant is shown on Figure 13.

Service Plant

- 1 Bay - Building extension 30 ft high x 24 ft long x 40 ft wide
- 2 - Feed pumps 250 gpm
- 3 - Ion exchange towers, 11 ft dia x 13 ft str side
- 2 - Decarbonator - spray nozzles
- 1 - Aerator, forced-draft type
- 2 - Settling basins, 40 ft diameter
- 1 - Direct contact softener 19 ft diameter x 15 ft high
- 2 - Gravity filters, 10 ft diameter
- 1 - Chlorination station
- Miscellaneous piping

Regeneration Plant

- 1 - Chemical precipitator - 15,000 gallons
- 1 - Feeder, lime slurry
- 2 sets - Plates and frames for pressure filter
- 1 - Stripper system, ammonia, capacity = 100,000 lb/day solution
- 1 - Feed tank, 700 gallons
- 1 - Product tank, 1000 gallons
- 1 - Recalcining furnace with auxiliaries
- 1 - CO₂ collection system
- 1 - CO₂ compressor
- 1 - CO₂ storage and feed station

BY CRM/BRN/DATE 6-25-69 SUBJECT AMP DEMONSTRATION PLANT SHEET NO. _____ OF _____
 CHKD. BY _____ DATE _____ TENTATIVE ARRANGEMENT OF JOB NO. 2674-01
 _____ FUTURE EXPANSION _____

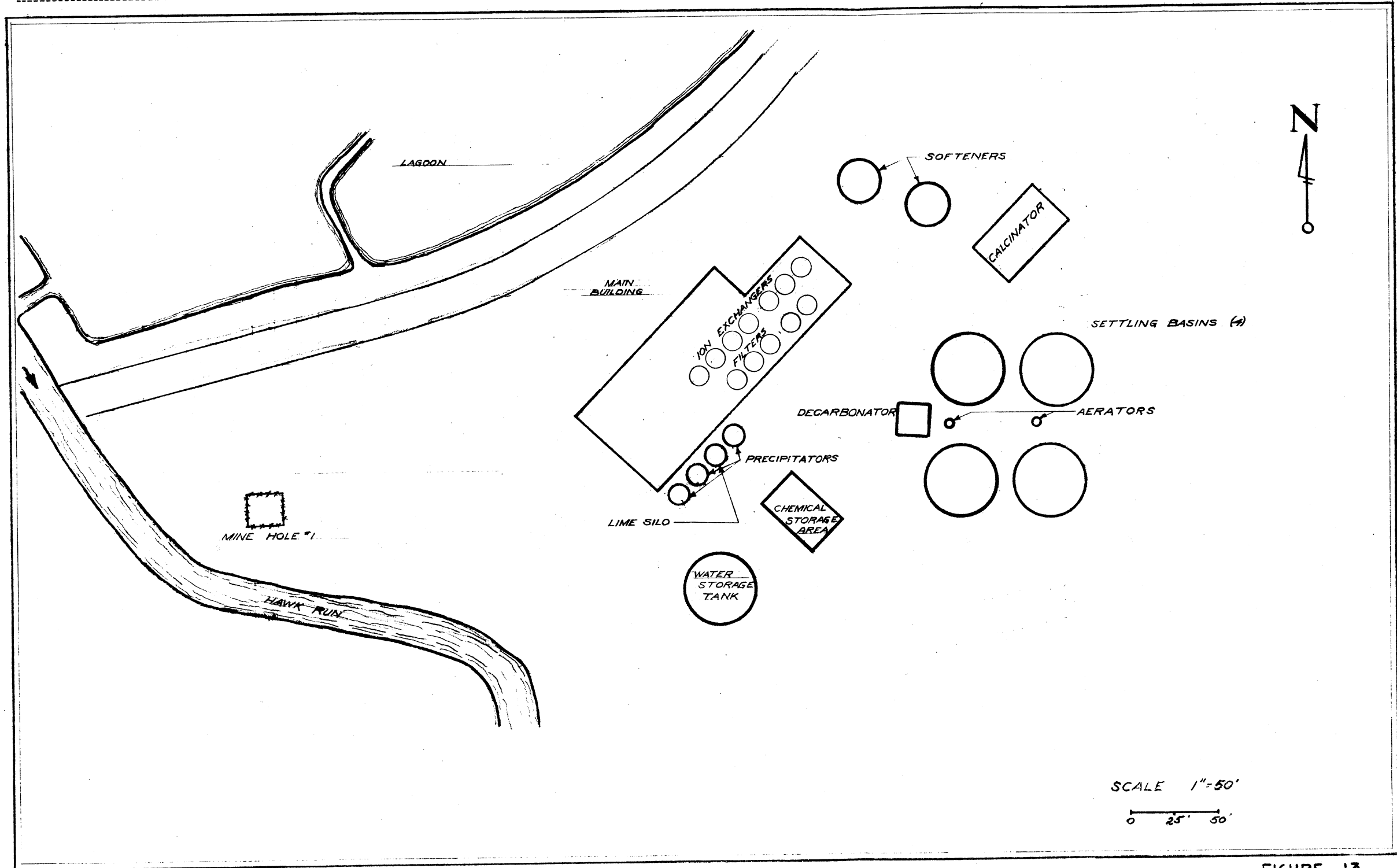


FIGURE 13

CAPITAL COST OF PROJECT

The capital cost of the proposed A.M.D. treatment plant is \$1,515,060. The capital costs are summarized in Table 5. A contingency of 10% is included in the above stated cost.

The capital costs are itemized in Table 6. Cost estimated for the proposed plant were obtained by contacting and discussing with various manufacturers, the components of the plant and the extreme and unusual operating conditions required for the operation of the plant. Manufacturers then submitted budget proposals for the various components. Chemical costs were obtained by contacting chemical suppliers in the local area of the plant location. Plant site preparations, building and tank foundations, heating, plumbing and electrical power supply equipment were estimated by the Burns and Roe estimating department.

Table 6

Itemized Capital Cost

<u>Item</u> <u>No.</u>	<u>Description</u>	<u>Cost</u>	
1	- Site Preparation		
	Building Foundation	\$ 33,400	
	Sump and Tank Foundations	42,400	
	Lagoon, Dam and Dike	30,000	
	Roads and Parking Area	17,800	
	Subtotal		\$ 123,600
2	- Building and Accessories		
	Building	65,000	
	Interior	25,000	
	Heating, Ventilating, Plumbing, Electrical	169,200	
	Subtotal		259,200
3	- Laboratory Equipment and Supplies	15,000	
	Subtotal		15,000
4	- Plant Chemical Inventory - One Month		
	Quicklime	1,920	
	Ammonia	360	
	Carbon Dioxide	5,580	
	Fuel Oil	720	
	Subtotal		8,580
5	- Ion Exchange Resin - \$85.00/cu ft	193,800	
	Subtotal		193,800
6	- Water Treatment Plant		
	Water Treatment Plant (Vendor Estimated)	306,200	
	Instrumentation, Special	20,000	
	Pumps - Raw Water, Service, Back- wash, Recovery, CaSO ₄ , Slurry	36,000	
	Chemical Storage Tanks, Fuel Oil Tank	11,350	
	Chemical Precipitators	18,000	
	Filter, Plate and Frame	34,300	
	Ammonia Stripper and Auxiliaries	38,000	
	Regenerant Storage Tank	3,800	
	Lime Slaker, Feeders and Silo	24,500	
	Waste Solids Conveyor	5,000	
	Piping, Plant and Interconnecting	30,000	
	Subtotal		527,150
7	- Water Treatment Plant Installation Costs	250,000	
	Subtotal		250,000
8	- Subtotal Plant Cost		1,377,330
9	- Contingency - 10%		<u>137,730</u>
10	- Total Plant Cost		\$1,515,060

DIRECT OPERATING COSTS OF TREATMENT PLANT

The expected chemical and power costs for purifying Hawk Run acid mine drainage water are shown in Table 7. The initial plant design utilizes ammonia recovery but does not include lime or carbon dioxide recovery. The expected chemical operating costs for the initial plant design is \$207.92 per day or \$75,890.80 per year. The expected power operating cost for the initial plant design is \$75.23 per day or \$27,459.00 per year. Thus, the operating cost of the treatment plant will be \$283.00 per day or \$103,351.00 per year.

When the capacity of the initial plant is increased to 1.0 million gpd, recovery of lime and carbon dioxide can be incorporated by the calcination of the calcium carbonate sludge. If we assume a 70 percent recovery of lime from the lime softener sludge and a 50 percent carbon dioxide recovery, the plant operating cost will be \$224.14 per day or \$71,811.10 per year. The chemical material cost is estimated to be \$119.00 per day or \$43,541.00 per year. The power operating cost is estimated to be \$104.00 per day or \$37,858.00 per year. The percentage of chemical recovery is based on data published on lime recovery at the Lake Tahoe sewage treatment plants. With some plant operating experience, the process efficiency should be further improved.

Table 7

Summary

Direct Operating Cost of Treatment Plant
Ammonia Recovery Only

	<u>Cost/Day</u>	<u>Cost/Year</u>
Chemicals	\$207.92	\$ 75,890.80
Fuel	52.50	19,162.50
Power	<u>22.73</u>	<u>8,297.55</u>
Total Operating Cost	\$283.15	\$103,350.85

CHEMICAL AND OPERATING COST ESTIMATE

1: Chemical Cost - 95% Ammonia Recovery, No Lime or Carbon Dioxide Recovery

<u>Material</u>	<u>Makeup (Lb/Day)</u>	<u>Unit Cost (\$/Lb)</u>	<u>Makeup Cost (\$/Day)</u>
Ammonia	160	\$0.030	\$ 4.80
Carbon Dioxide	6,180	\$0.0225	139.05
Lime - 95% CaO	6,430	\$0.00995	64.07
Total - 1 Day			\$207.92

2: Chemical Cost with Ammonia, Lime and Carbon Dioxide Recovery. (95% Ammonia Recovery; 70% Lime Recovery; 50% Carbon Dioxide Recovery)

	<u>Makeup (Lb/Day)</u>	<u>Recovery Makeup (Lb/Day)</u>	<u>Unit Cost (\$/Lb)</u>	<u>Makeup Cost (\$/Day)</u>
Ammonia	160	160	\$0.030	\$ 4.80
Carbon Dioxide	6,180	3,090	\$0.0225	69.52
Lime - 95% CaO	6,430	4,520	\$0.00995	44.97
Total - 1 Day				\$119.29

3: Chemical Cost with and without Lime and Carbon Dioxide Recovery

	<u>Without Recovery</u>	<u>With Recovery</u>
Cost per day	\$ 207.92	\$ 119.29
Cost per year	\$75,890.80	\$43,541.00

4: Steam Heat Fuel Cost

	<u>Btu/Day</u>	
Carbon Dioxide Vaporization - 1 day	1,620,000	
Ammonia Vaporization - 1 day	93,800	
Ammonia Stripper Still - 1 day	50,000,000	
Building Heat Load - 1 day	822,000	
Total Heat per Day		52,535,800
Fuel Oil, #2 - 150,000 Btu/gallon	350 gal/day	
Cost per day - \$0.15/gal	\$ 52.50/day	
Cost per year	\$19,162.50/year	

5: Electrical Power Cost

Motor Efficiency = 0.75 Power Cost = \$0.01 per kwhr

A. Without Lime and Carbon Dioxide Recovery

$$\text{Kwhr/day} = \frac{1705}{0.75} = 2273.3 \text{ kwhr/day}$$

$$\begin{aligned} \$/\text{day} &= 2273.3 \text{ kwhr/day} \quad \$0.01/\text{kwhr} = \$22.73/\text{day} \\ &= \$8,297.55/\text{year} \end{aligned}$$

B. With Lime and Carbon Dioxide Recovery (Assumed Power Rate \$0.01/kwhr)

Calcination Power:	2849 kwhr x \$0.01/kwhr = \$28.49/day
Plant Power :	22.73/day
Total Power :	\$51.22/day
	\$18,695.30/year

CHEMICAL AND OPERATING COST ESTIMATE (continued)

6: Total Chemical, Fuel and Power Cost

	<u>Ammonia Recovery Only</u>		<u>Ammonia, Lime, Carbon Dioxide Recovery</u>	
	<u>One Day</u>	<u>One Year</u>	<u>One Day</u>	<u>One Year</u>
Chemical	\$207.92	\$ 75,890.80	\$119.29	\$43,541.00
Power	22.73	8,297.55	51.22	18,695.30
Fuel	52.50	19,162.50	52.50	19,162.50
Total	\$283.15	\$103,350.85	\$222.91	\$81,398.80