# IMPROVED WATER MANAGEMENT OF COAL CONVERSION PROCESSES BY PRELIMINARY ABSORPTION OF HALIDES

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**Davy McKee**
ENGINEERS AND CONSTRUCTORS
CLEVELAND, OHIO

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SUMMARY

## 1.0 SUMMARY

Water management is an inherent consideration in coal conversion technology. Water supply is required for steam generation and may be required for process operations and evaporative cooling. Wastewater collection, processing, reuse and/or disposal must be provided for any process wastes and for storm runoff from coal handling and processing areas. For coal conversion processes that employ wet cleaning of gases, such as those supplying cooled gas to storage or to distribution systems, gas condensates are produced that contain various pollutants. Condensates from gases derived from coal may contain gas borne particulates, tars, oils or water solubles such as ammonia, chlorides, cyanides, fluorides, phenols and sulfides. The upgrading of contaminated gas condensates to service water quality by treatment oriented technology can entail a complex sequence of processing including desalination.

A modified wet gas cleaning system was proposed as offering promise of improving the practicality of reuse of gas condensates from coal gasification. This production oriented technology would replace the conventional single stage quench of the gases with a system incorporating a two-stage quench. The concept advocated the removal of particulates and the absorption of halides and other strong electrolytes in a controlled first stage quench that would condense a small fraction of the water vapor contained in the product gas. The low volume primary condensate would be purged from the system. The major fraction of condensate collected by the second stage quench would be relatively free of strong electrolytes and would be amenable to processing for reuse without resort to desalination technology. The volatile weak electrolytes in the secondary condensate would be removed by steam stripping and organic substances would be removed by biological treatment.

The objective of the study was to assess the technical, economic and environmental feasibility of the application of two stage gas quenching to a producer gas installation. The study approach consisted of a comparative analysis of the advantages and disadvantages of conventional single quenching versus the two stage quenching concept for hypothetical gasifier installations.

The gasifier selected for analysis was modeled after the pressurized, stirred, fixed-bed gasifier of the Morgantown Energy Technology Center. The hypothetical settings employed five 300 ton per day gasifiers for an operational capacity of 1500 tons per day. Carson City, Nevada was selected as a site representative of a Western installation, and Wheeling, West Virginia was selected as a site representative of an Eastern installation. The study analyzed air-blown and oxygen-blown gasifier operations on Illinois No. 6 and Montana Rosebud coals.

The Eastern and Western gasifier settings were selected as analogous to gasification facilities that might be utilized for the supply of low or medium Btu gas to major industries or industrial complexes that have fuel requirements that are not amenable to direct coal firing. Many industrial facilities in the Eastern and Midwestern United States are currently operating with the interruptable natural gas supplies and some industries are threatened with curtailment of natural gas supply. The gasifier settings were realistic and practical inasmuch as producer gas can replace natural gas for most industrial heating and firing operations with minor modification of the burner systems.

Alternative water management systems were conceptualized to maximize practical reuse of gas condensates and minimize blowdown requirements. Effluent recovery and disposal concepts were evaluated, including reverse osmosis, thermal evaporation, evaporation ponds, storage with barging to sea and storage with controlled discharge to watercourse. An engineering trade-off analysis was performed to provide a comparative technical and economic assessment of single stage gas quenching versus two stage gas quenching.

The thermodynamic analysis of mass transfer in the gas train indicated that halides and other strong electrolytes were absorbed to a high degree in a first stage quench condensing 10 gallons of water per ton of coal. Although a sieve tray absorber was employed to establish equilibrium conditions in the first stage quench, the study results indicated that an elementary absorption device, such as a venturi scrubber, in conjunction with a knockout drum and efficient mist eliminator, would suffice for absorption of halides. Alternative control schemes were devised to effect the condensation of a selected volume of gas moisture in the first stage quench. The concept for the engineering and operation of a two stage quench system was considered sound.

The secondary condensate was indicated to be amenable to recovery as a low dissolved solids service water after processing by steam stripping plus activated sludge treatment. The formation of thiocyanate in the secondary condensate is expected to produce some residual sulfate in the effluent from biological treatment, but halide concentrations are indicated to be extremely low. It is projected that the biotreated effluent could be processed by reverse osmosis to a permeate with quality approaching that of evaporator condensate.

The results indicated that the alternative of preliminary absorption of halides by application of a two stage quench system was a cost effective design procedure for operation of the selected gasifier installation on Illinois No. 6 coal containing 0.25 percent chloride and 0.012 percent fluoride. Subject to the assumptions employed in the study, the alternative of the two stage quench possessed a cost advantage over a single quench system of 0.99 $/ton and 1.07 $/ton respectively for gasifier operation air blown and oxygen blown.

The results indicated that a conventional single quench system was the cost effective design procedure for operation of the selected gasifier installation on Montana Rosebud coal containing 0.03 percent chloride and 0.003 percent fluoride. Subject to the assumptions employed in the study, the single quench alternative possessed a cost advantage over a two stage quench system of 0.95 $/ton and 1.26 $/ton respectively for gasifier operation air blown and oxygen blown.

The dominant variable affecting the cost effectiveness of the alternatives was indicated to be the halide content of the coal feed for situations involving substantial reuse of gas condensate in cooling tower circuits. Under the conditions of the study, the trade-off cost between single stage and two stage gas quenching systems was indicated to occur at coal total halide concentrations of 0.15 and 0.16 percent (expressed as chlorides) respectively for air blown and oxygen blown gasification. It was concluded that the two stage quench would be widely applicable to coal conversion processes employing wet cleaning of halide bearing gases such as are characteristic of many Eastern coals.

Additional studies were recommended for the assessment of thiocyanate formation and for the evaluation gas train materials of fabrication. Cost estimates were based on carbon steel decanters and vessels although the presence of halides in conjunction with low pH could impair the serviceability of the units. The application of pH adjustment and/or inhibitors warrant evaluation as well as alternate materials. Design revisions were suggested to improve the cost effectiveness of the two stage quench system but it is inherently more elaborate than a single quench system.

Incineration with disposal of salt cake to sea, using a two stage gas quench system in conjunction with desalination by reverse osmosis and thermal evaporation, was indicated as the preferred disposal concept for inorganic wastewater blowdown from a gasification plant operating on high halide coal at a location in the vicinity of Wheeling, West Virginia.

Pond evaporation, using a single gas quench system in conjunction with desalination by reverse osmosis, was considered as the preferred disposal concept for inorganic wastewater blowdown from a gasification plant operating on low halide coal near Carson City, Nevada.

The water management concepts developed in the study are consistent with the objective of convergence on "zero discharge." The low volume concentrates obtained from application of two stage gas quenching, or by desalination, are amenable to processing by disposal methods that produce dry residues.

## 2.0 INTRODUCTION

Water management is an inherent consideration in coal conversion technology. Water supply is required for steam generation and may be required for process operations and evaporative cooling. Wastewater collection, processing, reuse and/or disposal must be provided for any process wastes and for storm runoff from coal handling and processing areas. For coal conversion processes that employ wet cleaning of gases, such as those supplying cooled gas to storage or to distribution systems, gas condensates are produced that contain various pollutants. Condensates from gases derived from coal may contain gas borne particulates, tars, oils or water solubles such as ammonia, chlorides, cyanides, fluorides, phenols and sulfides. The upgrading of contaminated gas condensates to service water quality by treatment oriented technology can entail a complex sequence of processing including desalination.

A modified wet gas cleaning system was proposed by Davy McKee Corporation as offering promise of improving the practicality of reuse of gas condensates from coal gasification. This production oriented technology replaced the conventional single stage quench of the gases with a system incorporating a two-stage quench. The concept advocated the removal of particulates and the absorption of halides and other strong electrolytes in a controlled first stage quench that would condense a small fraction of the water vapor contained in the product gas. The low volume primary condensate would be purged from the system. The major fraction of condensate collected by the second stage quench would be relatively free of strong electrolytes and would be amenable to processing for reuse without resort to desalination technology. The volatile weak electrolytes in the secondary condensate would be removed by steam stripping and organic substances would be removed by biological treatment.

The two stage quench system holds promise for the improvement of water management from coal conversion operations employing wet cleaning of gases. The potential improvement could be manifested as water conservation, energy conservation or conservation of economic resources. A proposal for a study to evaluate the concept for a producer gas installation operating air-blown and oxygen-blown on Eastern and Western coals was accepted by the Division of Environmental Control Technology, DOE.

DESCRIPTION
AND SCOPE

## 3.0 PROJECT DESCRIPTION AND SCOPE OF WORK

The objective of the study was the assessment of the technical, economic and environmental feasibility of the application of two stage gas quenching to a producer gas installation. The study approach consisted of a comparative analysis of the advantages and disadvantages of conventional single quenching versus the two stage quenching concept for hypothetical gasifier installations.

The gasifier selected for analysis was modeled after the pressurized, stirred, fixed-bed gasifier of the Morgantown Energy Technology Center. The hypothetical settings employed five 300 ton per day gasifiers for an operational capacity of 1500 tons per day. Carson City, Nevada was selected as a site representative of a Western installation and Wheeling, West Virginia was selected as a site representative of an Eastern installation. The study analyzed air-blown and oxygen-blown gasifier operations on Illinois No. 6 and Montana Rosebud coals.

The Eastern and Western gasifier settings were selected as analogous to gasification facilities that might be utilized for the supply of low or medium Btu gas to major industries or industrial complexes that have fuel requirements that are not amenable to direct coal firing. Many industrial facilities in the Eastern and Midwestern United States are currently operating with interruptable natural gas supplies and some industries are threatened with curtailment of natural gas supply. The gasifier settings were realistic and practical inasmuch as producer gas can replace natural gas for most industrial heating and firing operations with minor modification of the burner systems.

The selection of the Morgantown Energy Technology Center gasifier provided the study with an operational data base relative to a fixed bed gasifier that has a demonstrated capability of being able to operate on any rank of coal. It is a fixed bed gasifier with a high carbon conversion efficiency and excellent turndown capability. The process falls within the classification of tar producing coal gasification technology for which there exists some background of wastewater processing experience. Thus the gasifier was selected for relevance to the setting and for study credibility.

The selection of sites and gasification system provided the project definition required to enable the synthesis of comparable water management situations employing single stage and two-stage gas quenching. Sufficient engineering of the systems was performed to develop water management systems, to predict the performance of component processes and to enable the preparation of comparative estimates of cost. The results obtained were indicative of the applicability and the relative feasibility of the two stage gas quench system.

The scope of work required the development of technical reference information on the systems germane to the study and the application of thermodynamic analyses for the projection of system performance. The computer projections were to be coordinated with appropriate field data. An engineering trade-off analysis was to be performed to provide a comparative technical and economic assessment of single stage gas quenching versus two stage gas quenching. Alternative water management systems were to be conceptualized to maximize practical reuse of effluent flows so as to minimize blowdown requirements from the coal conversion systems. Effluent disposal systems including fueled heat evaporation, evaporation ponds, storage with barging to sea, and storage with controlled discharge were to be conceptualized. Comparative cost estimates were to be tabulated for candidate water management alternatives employing single stage and two stage gas quenching systems. The boundaries of feasibility of single stage and two stage gas quenching systems were to be identified.

METHODOLOGY
TECHNICAL

## 4.0 METHODS - TECHNICAL ANALYSIS

The study sought analysis of the applicability of a two stage gas quench system for the improvement of water management from coal conversion processes. The study approach employed a comparative analysis of single quench and two stage gas quench systems attached to hypothetical 1500 ton per day producer gas facilities employing five 300 ton per day fixed-bed gasifiers. The gasifiers were modeled after the pressurized, stirred, fixed bed gasifier of the Morgantown Energy Technology Center and were operated air blown and oxygen blown. Hypothetical gasifier locations at Wheeling, West Virginia and Carson City, Nevada were evaluated to represent Eastern and Western site situations.

A schematic of a five gasifier installation with a capacity of 1500 tons of coal per day is given as Figure 4-1. The illustration is applicable to both single stage quench and two stage quench gasification systems selected for the study. The process flow schematic for the gas train employing the single stage gas quench is given in Figure 4-2. A plot plan of the gas plant selected as the single quench alternative is given in Figure 4-3. A block diagram is given as Figure 4-4 that illustrates the water management concept proposed for the condensate from the single stage quench alternative.

The process flow schematic for the gas train employing a two-stage gas quench is given in Figure 4-5. A plot plan for the gas plant selected as the two stage quench alternative is given in Figure 4-6. A block diagram is given as Figure 4-7 that illustrates the water management concept proposed for the condensates from the two stage quench alternative. The selection of a definitive hypothetical gasifier installation, connected to alternative single stage quench and two stage quench gas cleaning systems, provided the basis for an evaluation of the technical and economic feasibility of the alternatives.

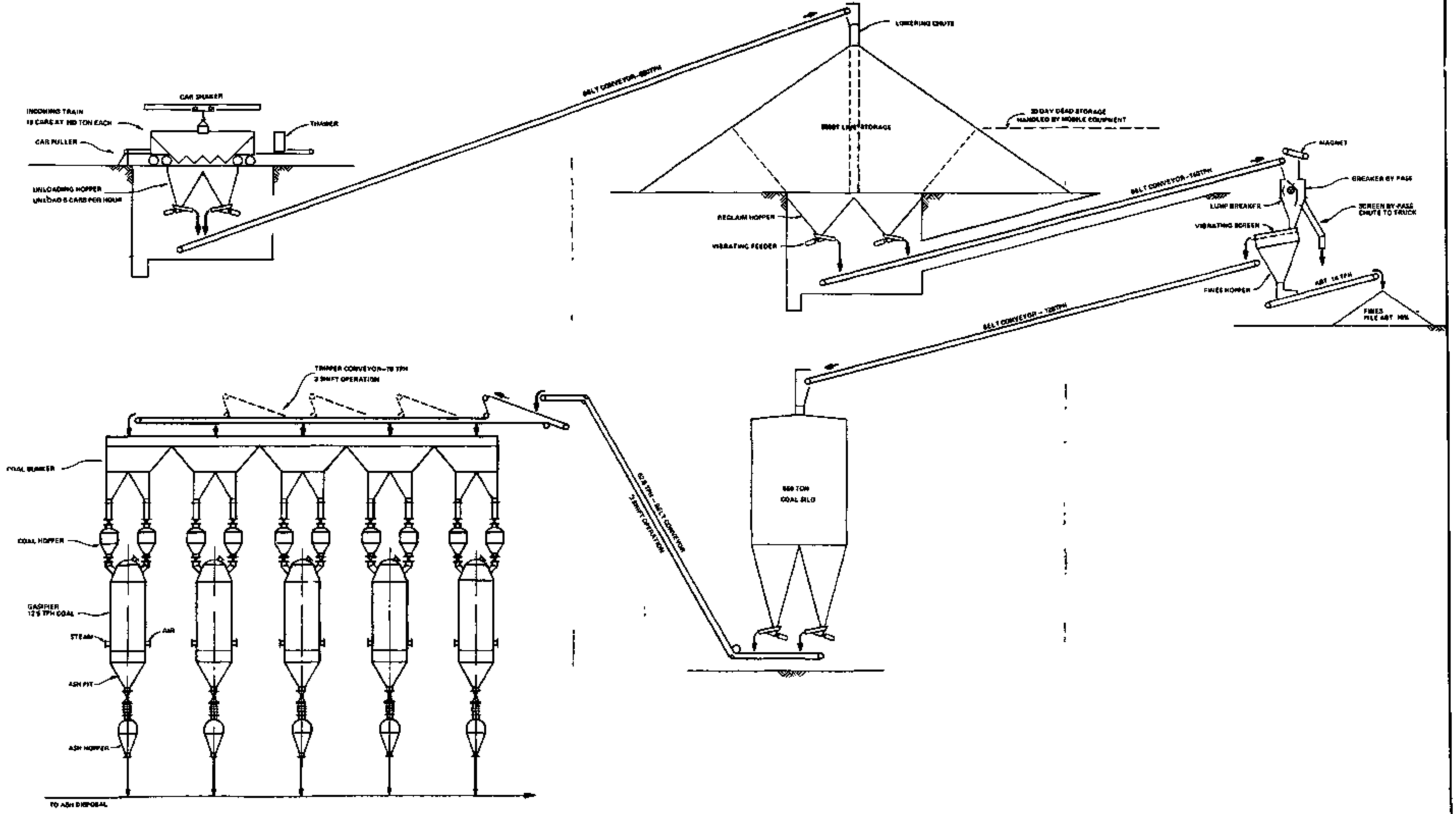

FIGURE 4-1 COAL HANDLING AND GASIFICATION SYSTEM

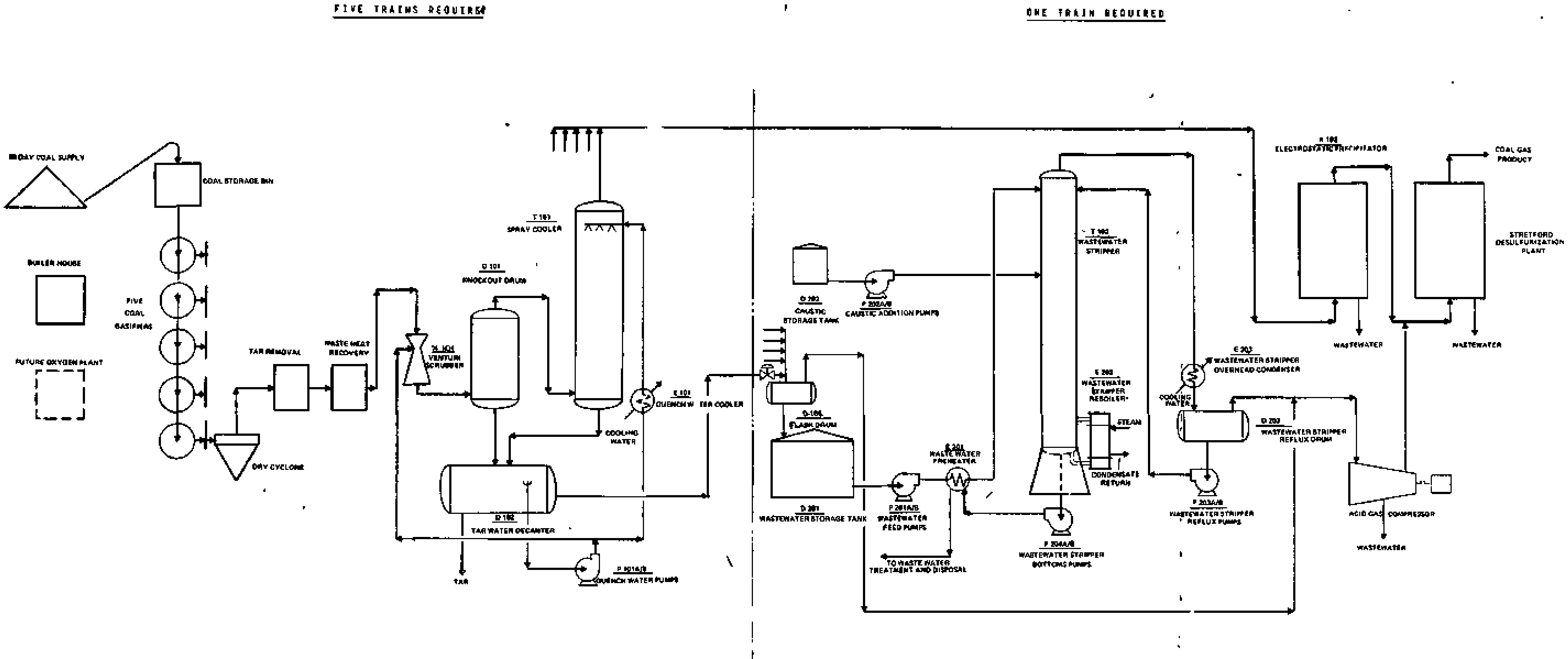

FIGURE 4-2. PROCESS FLOW–CONVENTIONAL SINGLE STAGE GAS QUENCH

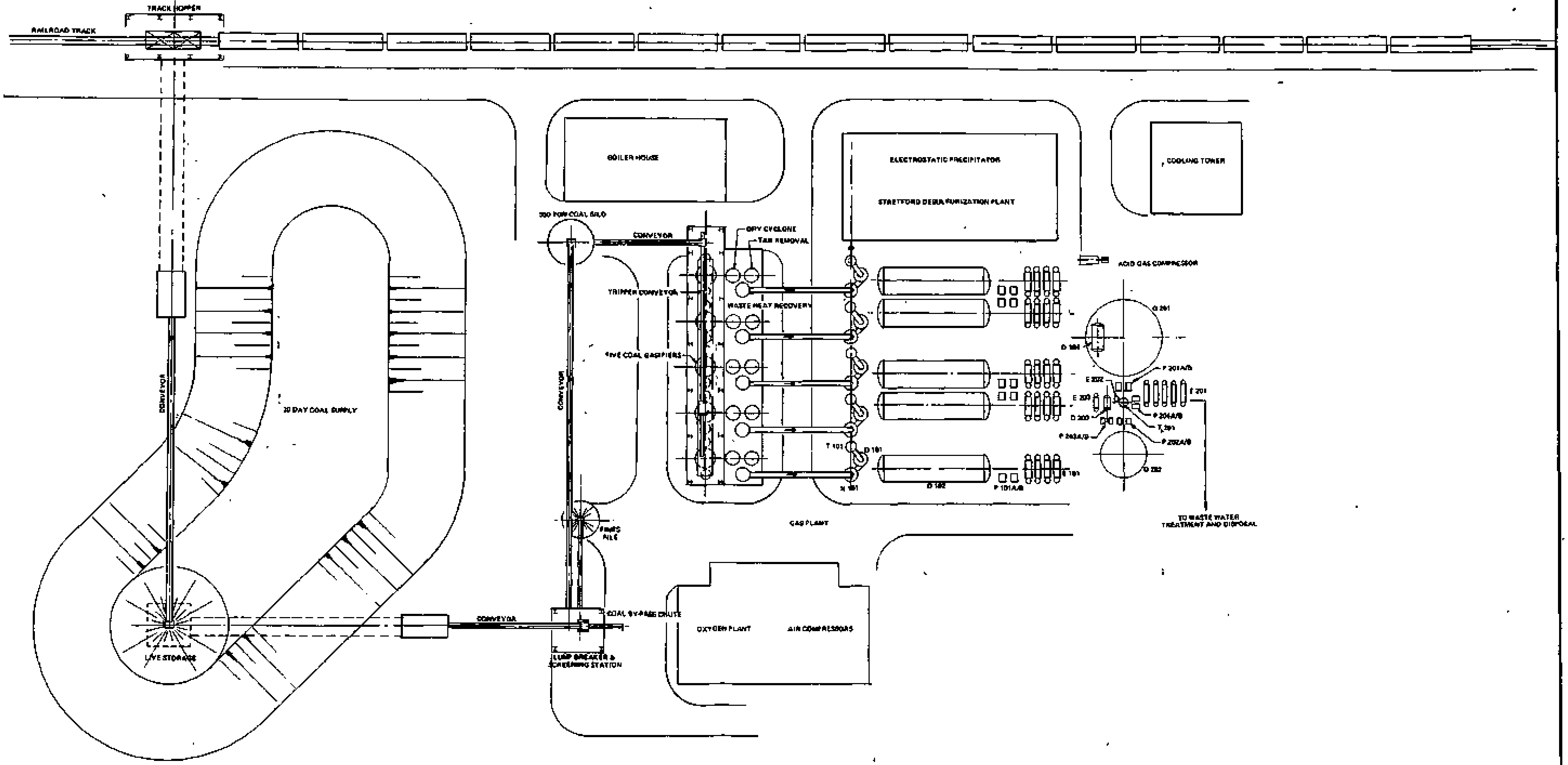

FIGURE 4-3 PLOT PLAN – SINGLE STAGE GAS QUENCH

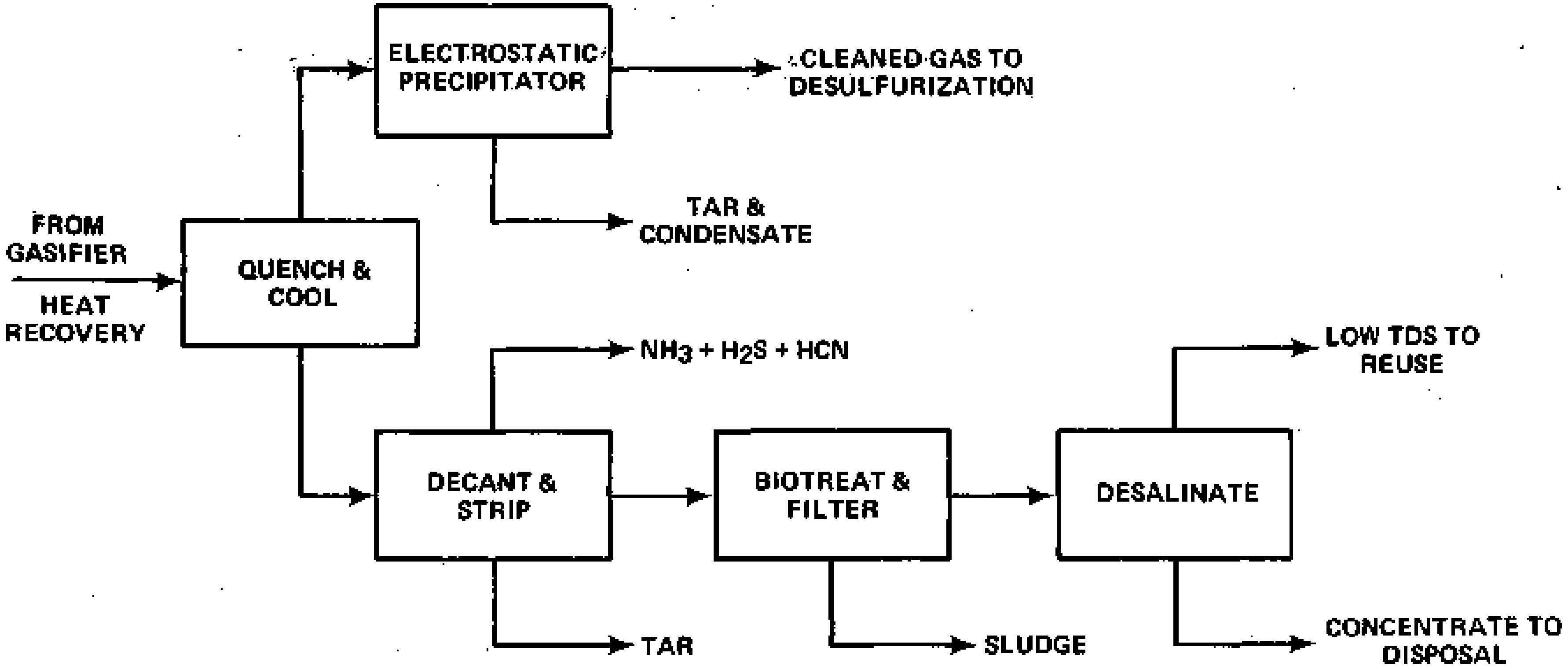

FIGURE 4-4: CONVENTIONAL GAS CLEANING AND CONDENSATE TREATMENT

**Davy McKee**

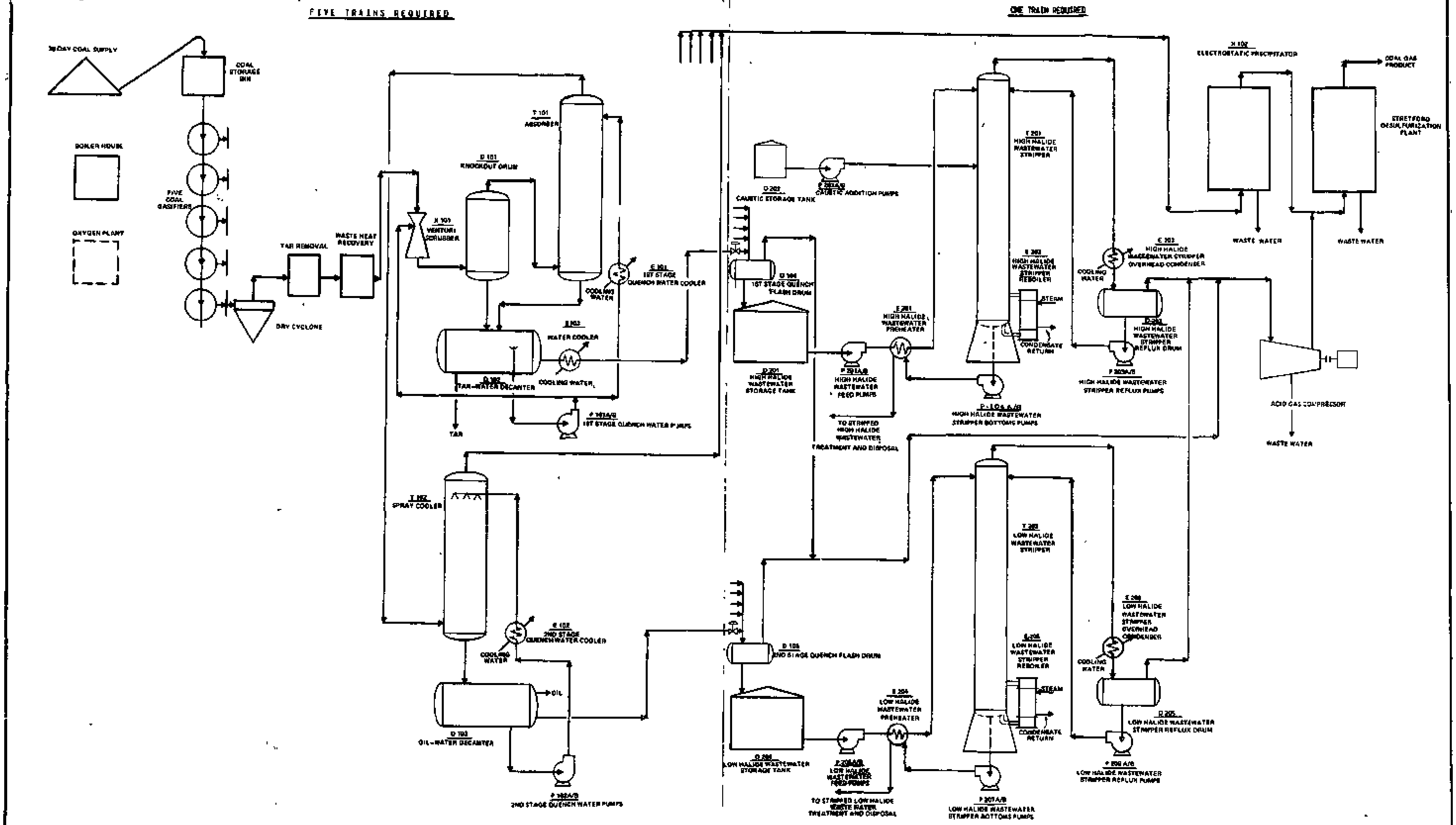

FIGURE 4-5 PROCESS FLOW – TWO STAGE QUENCH

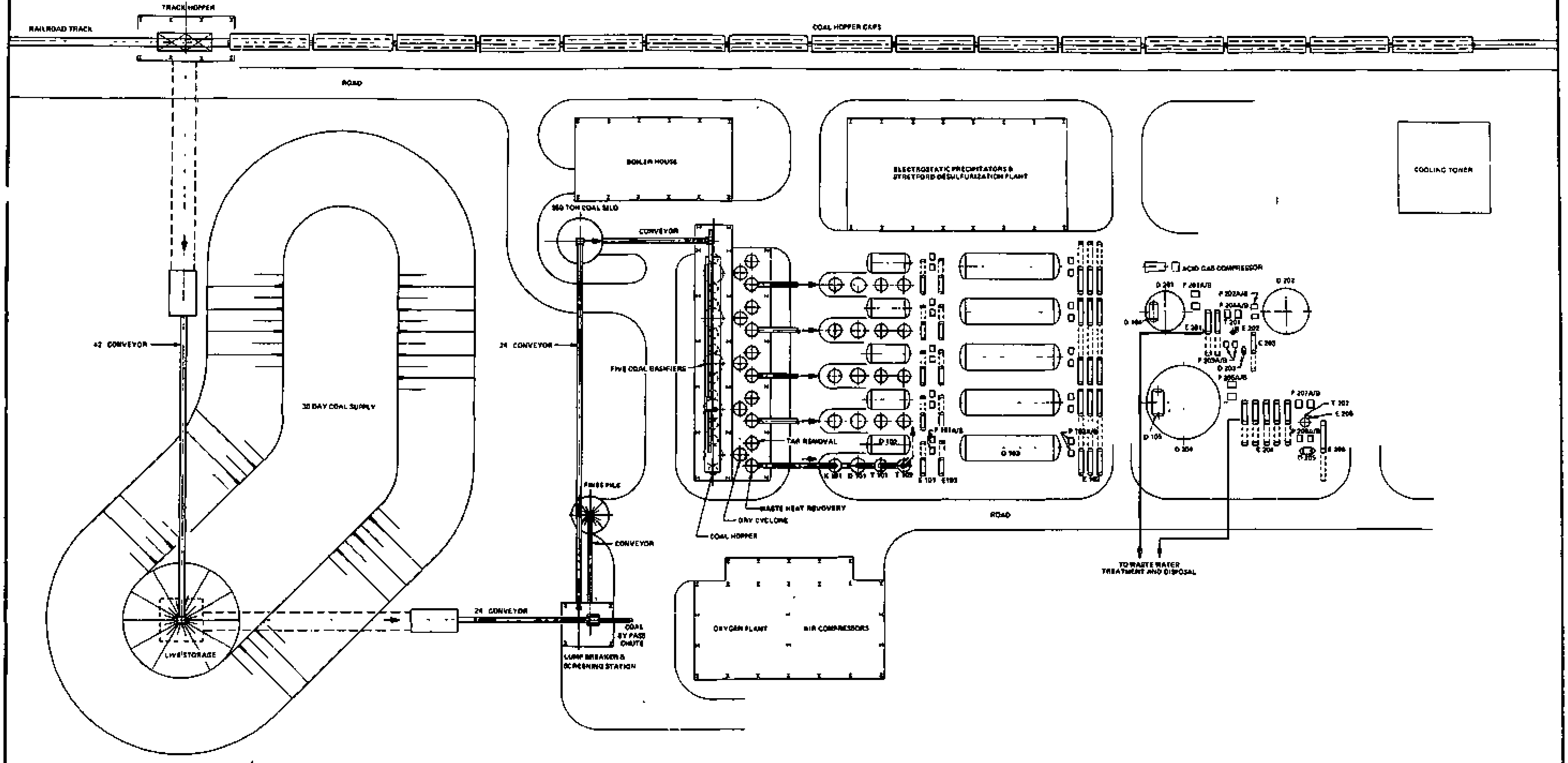

FIGURE 4-6: PLOT PLAN—TWO STAGE GAS QUENCH

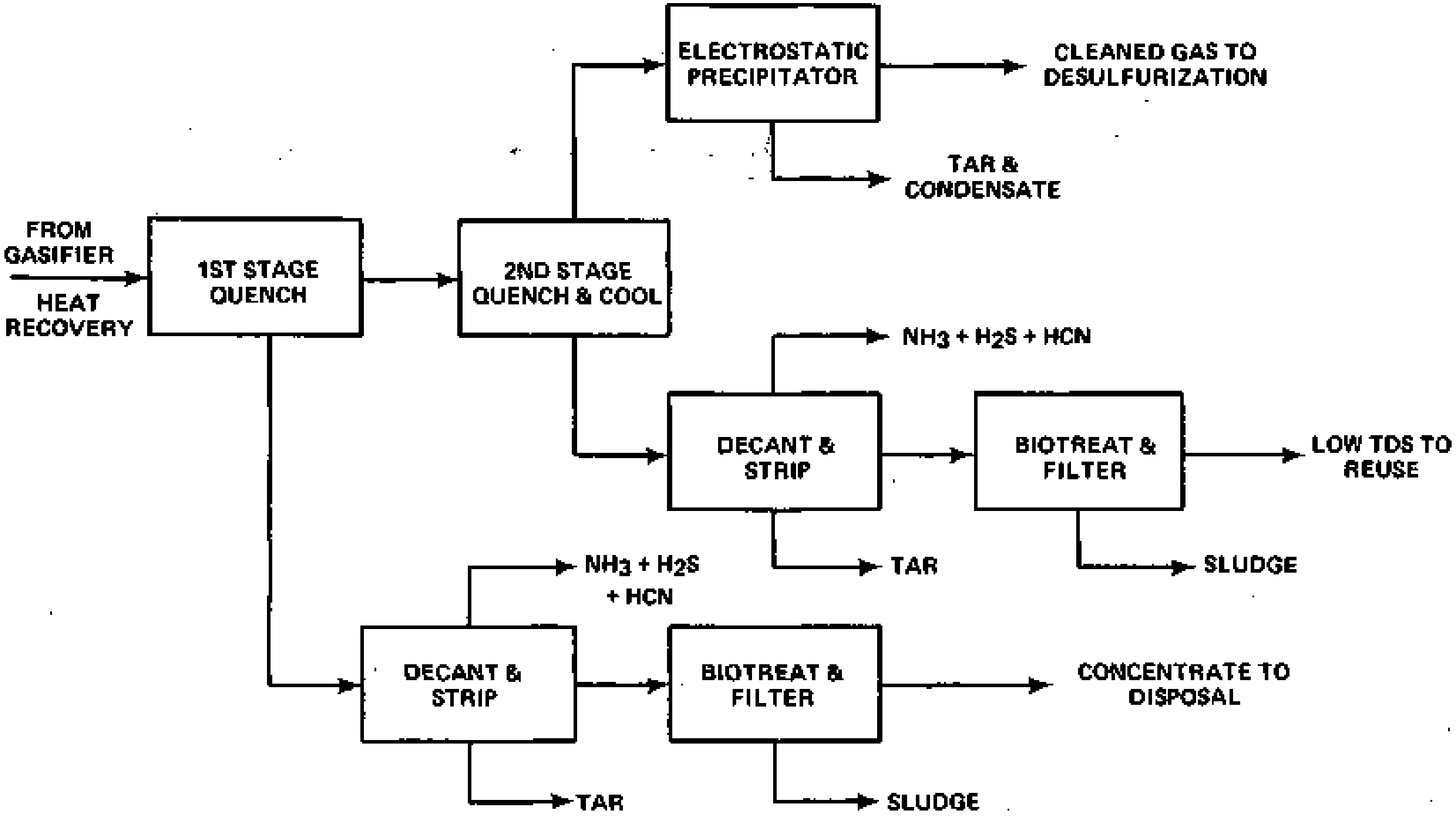

FIGURE 4-7: TWO STAGE GAS CLEANING AND CONDENSATE TREATMENT

## 4.1 Gasifier Material and Energy Balances

The material and energy balance calculations for the proposed commercial gasifiers are based on the operating data of the Morgantown Energy Technology Center (METC) pilot plant (1). Illinois No. 6 and Montana Rosebud coals were chosen for the study and both air blown and oxygen blown gasifiers were considered. The proximate analysis for each coal is shown in Table 4-1.

The overall material and energy balances for the gasifier are shown in Table 4-2 for the Illinois No. 6 coal and in Table 4-3 for the Montana Rosebud. Energy balances around the METC pilot plant gasifier indicate heat losses which are approximately 15% of the total heat input. The METC gasifier considered for this study is specified with a steam jacket and heat losses are 2% of the total heat input into the gasifier. Because of the reduction in percentage of heat lost to the surroundings, modifications in the gasifier steam requirements were necessary. The steam input requirements are less for the steam jacketed case on a per weight of coal basis. The predicted steam requirements are very close to steam input data reported for the Lurgi unit in Westfield, Scotland (2). As a result, the predicted gasification product distribution is based on both the Lurgi and METC operating data as well as on literature characterizing coal properties.

Typical operating conditions for the METC gasifier are maximum bed temperatures of 2500°F and outlet gas temperature values between 900°F and 1200°F. The gasifier operating pressure for all the cases considered was fixed at 103 psia. The product gas flow rates and compositions were calculated at these conditions for each case assuming 95% carbon conversion. The quantities of gases and liquids produced are given in Tables 4-2 and 4-3. The composition of the product gas for each case considered is shown in Tables 4-4 and 4-5.

## Table 4-1: Analysis of Coal Feeds

| | Illinois No. 6 Coal | Montana Rosebud Coal |
|---|---|---|
| | wt.% | |
| C | 65.70 | 50.56 |
| H | 4.80 | 3.18 |
| S | 3.70 | 1.09 |
| N | 1.10 | 0.90 |
| O | 8.04 | 9.81 |
| Cl | 0.25 | 0.03 |
| F | 120 ppmw | 30 ppmw |
| Ash | 11.20 | 9.73 |
| $H_2O$ | 5.20 | 24.70 |
| | 100.00 | 100.00 |
| Higher Heating Value, Btu/lb | 11,750 | 8,611 |

ppmw = parts per million by weight

## Table 4-2: Material and Energy Balances for Gasifier Operation on Illinois No. 6 Coal

### Illinois No. 6 Coal, Airblown

| INPUT | tons/day | Million Btu/day |
|---|---|---|
| Coal | 1,500 | 35,250 |
| Steam | 2,400 | 5,615 |
| Air | 4,013 | 468+ |
|  | 7,913 | 41,333 |

+Sensible heat

| OUTPUT | tons/day | Million Btu/day |
|---|---|---|
| Product Gas | 7,610 |  |
| Heat of Combustion |  | 26,499 |
| Sensible Heat |  | 9,523 |
| Tar, Oil & Phenols | 112 | 3,705* |
| Dust | 15 | 345* |
| Ash | 176 | 450* |
| Steam Jacket & Heat Losses | - | 811 |
|  | 7,913 | 41,333 |

*Includes heat of combustion, latent heat and sensible heat

### Illinois No. 6 Coal, Oxygen Blown

| INPUT | tons/day | Million Btu/day |
|---|---|---|
| Coal | 1,500 | 32,250 |
| Steam | 3,300 | 7,720 |
| $O_2$ | 886 | 93+ |
|  | 5,686 | 43,063 |

+Sensible heat

| OUTPUT | tons/day | Million Btu/day |
|---|---|---|
| Product Gas | 5,383 |  |
| Heat of Combustion |  | 26,930 |
| Sensible Heat |  | 10,822 |
| Tar, Oil & Phenols | 112 | 3,705* |
| Dust | 15 | 345* |
| Ash | 176 | 450* |
| Steam Jacket & Heat Losses | - | 811 |
|  | 5,686 | 43,063 |

*Includes heat of combustion, latent heat and sensible heat

## Table 4-3: Material and Energy Balances for Operation on Montana Rosebud Coal

### Montana Rosebud Coal, Airblown

| INPUT | tons/day | Million Btu/day |
|---|---|---|
| Coal | 1,500 | 25,833 |
| Steam | 900 | 2,105 |
| Air | 2,833 | 330+ |
| | 5,233 | 28,268 |

+Sensible heat

| OUTPUT | tons/day | Million Btu/day |
|---|---|---|
| Product Gas | 4,965 | |
| Heat of Combustion | | 18,400 |
| Sensible Heat | | 5,121 |
| Tar, Oil & Phenols | 104 | 3,511* |
| Dust | 15 | 281* |
| Ash | 149 | 382* |
| Steam Jacket & Heat Losses | - | 573 |
| | 5,233 | 28,268 |

*Includes heat of combustion, latent heat and sensible heat

### Montana Rosebud Coal, Oxygen Blown

| INPUT | tons/day | Million Btu/day |
|---|---|---|
| Coal | 1,500 | 25,833 |
| Steam | 1,500 | 3,509 |
| $O_2$ | 611 | 64+ |
| | 3,611 | 29,406 |

+Sensible heat

| OUTPUT | tons/day | Million Btu/day |
|---|---|---|
| Product Gas | 3,343 | |
| Heat of Combustion | | 19,011 |
| Sensible Heat | | 5,648 |
| Tar, Oil & Phenols | 104 | 3,511* |
| Dust | 15 | 281* |
| Ash | 176 | 282* |
| Steam Jacket & Heat Losses | - | 573* |
| | 3,611 | 29,406 |

*Includes heat of combustion, latent heat and sensible heat

Of particular interest for purposes of this study, are the concentration levels of halides in the product stream. Examination of Tables 4-4 and 4-5 reveals that the concentration of hydrogen-chloride in the product gases is much higher for the Illinois No. 6 case than the Montana Rosebud case. When considering the concentration of hydrogen fluorides in the product gases, higher values are also derived from the Illinois No. 6 coal. One should note that the difference is not an order of magnitude in value as it is for hydrogen chloride. It should also be noted that for each case the halide concentration in the outlet gas is higher for the oxygen blown case than the air blown case. This reflects the presence of nitrogen in product gases from air blown gasifiers.

Other contaminants also play a role in the design of the gas quench stage. It can be seen that appreciable quantities of ammonia and hydrogen cyanide are present in the outlet gas stream for both coals. Sulfur compounds ($H_2S$, COS) can also be found in substantial quantities and concentration levels in the product gases. Concentrations of sulfur compounds are at least twice as great for the high sulfur containing Illinois No. 6 coal as for the Montana Rosebud coal.

## 4.2 Process Design of the Gas Cleaning Systems

Process design schemes were prepared for the single stage quench system (Fig. 4-2) and the two stage quench system (Fig. 4-5). It was assumed that each of the five gasifiers would feed into a separate gas treating train. This was done in anticipation of periodic system shutdowns for system maintenance. With five trains, any one gasifier could be shutdown without interrupting operation of the other units. The only gas clean-up unit servicing full flow from all the gasifiers would be the electrostatic precipitator.

## Table 4-4: Producer Off Gas Analysis - Illinois No. 6 Coal

| | Air Blown | | Oxygen Blown | |
|---|---|---|---|---|
| | Wet Gas | Dry Gas | Wet Gas | Dry Gas |
| $CO_2$ | 9.56 | 13.79 | 13.28 | 28.74 |
| CO | 9.90 | 14.30 | 9.93 | 21.49 |
| $H_2$ | 15.01 | 21.67 | 18.96 | 41.03 |
| $H_2O$ | 30.73 | - | 53.80 | - |
| $CH_4$ | 1.59 | 2.29 | 2.74 | 5.93 |
| $C_2H_6$ | 0.14 | 0.20 | 0.17 | 0.37 |
| $H_2S$ + COS | 0.47 | 0.68 | 0.58 | 1.26 |
| $N_2$ + A | 32.38 | 46.75 | 0.27 | 0.59 |
| $NH_3$ | 0.17 | 0.25 | 0.21 | 0.46 |
| HCl | 311 ppmv | 449 ppmv | 383 ppmv | 829 ppmv |
| HF | 28 ppmv | 40 ppmv | 34 ppmv | 74 ppmv |
| HCN | 173 ppmv | 250 ppmv | 214 ppmv | 0.05 |
| | 100.00 | 100.00 | 100.00 | 100.00 |
| Higher Heating Value, Btu/scf | 102.9 | 148.6 | 128.9 | 279.0 |
| Gas Outlet Temperature °F | 1050 | | 1070 | |
| Operating Pressure psia | 103 | | 103 | |
| Gas Flow, scfm | 178800 | | 145100 | |

Steam fed to the gasifier @ 110 psi, 350°F .

## Table 4-5: Producer Off Gas Analysis - Montana Rosebud Coal

| | Air Blown | | Oxygen Blown | |
|---|---|---|---|---|
| | Wet Gas | Dry Gas | Wet Gas | Dry Gas |
| $CO_2$ | 11.72 | 14.88 | 17.36 | 30.76 |
| CO | 11.88 | 15.08 | 11.99 | 21.24 |
| $H_2$ | 16.32 | 20.71 | 22.80 | 40.41 |
| $H_2O$ | 21.23 | - | 43.56 | - |
| $CH_4$ | 1.69 | 2.14 | 3.07 | 5.43 |
| $C_2H_6$ | 0.17 | 0.21 | 0.22 | 0.38 |
| $H_2S$ + COS | 0.22 | 0.28 | 0.28 | 0.50 |
| $N_2$ + A | 36.47 | 46.31 | 0.33 | 0.58 |
| $NH_3$ | 0.27 | 0.35 | 0.35 | 0.62 |
| HCl | 60 ppmv | 76 ppmv | 77 ppmv | 136 ppmv |
| HF | 11 ppmv | 14 ppmv | 14 ppmv | 26 ppmv |
| HCN | 272 ppmv | 345 ppmv | 348 ppmv | 617 ppmv |
| | 100.00 | 100.00 | 100.00 | 100.00 |
| Higher Heating Value, Btu/scf | 114.0 | 144.8 | 151.0 | 267.6 |
| Gas Outlet Temperature °F | 1050 | | 1040 | |
| Operating Pressure psia | 103 | | 103 | |
| Gas Flow, scfm | 112100 | | 87400 | |

In each processing scheme considered, the product gases first enter a tar removal unit and then a waste heat recovery system. Following this, the gas enters a venturi scrubber at 400°F and 95 psia. The gas is cooled as it undergoes close turbulent contact with water recycled from a decanter.

At this point the gas and quench water are assumed to be at their adiabatic saturation temperature (about 290°F). The mixture then enters a knockout drum in which it is dewatered. The tar, oil, solids and water are drained from the knockout drum into a decanter for separation. The gaseous products pass through the knockout drum and beyond this point, the processing scheme differs for the single quench and two stage quench gas cleaning systems considered.

### 4.2.1 Single Quench Gas Cleaning System

In the single stage quench gas cleaning system, illustrated in Figure 4-2, a countercurrent spray cooler (T-101) condenses most of the water vapor contained in the product gas stream. The gas then leaves the countercurrent spray cooler at a temperature of 110°F. The liquid stream exits the bottom of the spray cooler at a temperature of 130°F. Some light oils which remain in the vapor state at the 290°F exit temperature of the venturi scrubber are condensed in the spray cooler and are removed with the condensed phase. In addition to light oils, residual particulate matter is also removed in this effluent stream. This emulsion and any contained particulate matter enters a decanter unit (D-102) along with the liquid effluent discharged from the knockout drum (D-101). The gas stream leaving the spray cooler is sent to an electrostatic precipitator (X-102). This unit is designed to remove water and tar mists, as well as any entrained particulates contained in the product gas streams coming from all five gasifiers.

The mixture in the decanter is separated into a water fraction and a waste fraction consisting of tar, oil and solids. The water stream from the decanter is recycled as cooling water to the spray cooler and venturi scrubber by means of a quench water pump (P-101 A/B). The water recycled to the spray cooler is cooled to 105°F in a heat exchanger (E-101) unit contained in the recirculation loop. The tars, oils and solids are removed from the decanter in a separate stream and are discharged to disposal.

Any accumulation of condensate in the decanter is bled off and sent to a flash drum (D-104). The flash drum is operated at a reduced pressure which results in the evolution of gases dissolved in the condensate. The gases coming off the flash drum are sent to the acid gas compressor and combined with the product gas stream coming from the electrostatic precipitator. This gas then undergoes desulfurization in a conventional Stretford plant.

The liquid stream coming off the flash drum is sent to a wastewater storage tank (D-201) where it is held for further processing. Additional required treatment consists of steam stripping the flash drum effluent in the stripper-reboiler unit (E-202). This is done to remove $NH_3$, HCN, $CO_2$, and $H_2S$. Sodium hydroxide is added to the stripper to free fixed ammonia.

The bottoms from the stripper column are removed and used to preheat the stripper feed in the wastewater preheater (E-201). The effluent is then treated for the removal of cyanides, thiocyanates, phenols and other organics as per Figure 4-4. This is accomplished by employing the activated sludge process. The effluent is then filtered by granular media filtration for capture of fugitive suspended solids.

### 4.2.2 Description of Two Stage Quench Cleaning System

The process scheme for the two stage quench system is very similar to that of the single stage system. The proposed two stage quench system consists of an absorbing column and a spray cooler connected in series as per Figure 4-5 and each of these units is followed by a separate wastewater treatment train. Separate wastewater processing trains are necessary to isolate the high halide containing effluent from the reuseable low halide containing effluent as illustrated in Figure 4-7.

As in the single stage quench process, the effluent stream from the venturi scrubber enters a knockout drum (D-101) where it is dewatered. The gas stream then enters a sieve tray absorber (T-101) rather than a spray cooler as in the single stage quench system. The absorber column provides the means for attainment of equilibrium between gaseous and liquid phases at minimal condensation rates. This procedure produces a low volume flow of condensate having a high halide content. An efficient mist eliminator is incorporated in the absorber.

The absorber column (T-101) has been designed to condense from 5 to 30 percent of the water vapor that is contained in the venturi scrubber effluent gas stream. As shown in the absorber control scheme, Figure 4-8, the condensation rate in the absorber is fixed by means of a set point controller according to prevailing gasifier operating conditions.

The high halide condensate stream from the absorber is circulated through a variable volume decanter unit to separate tar condensed from the gas. Any accumulation of water in the decanter is bled off and cooled to 180°F in the water cooler (E-103). The stream is sent to an atmospheric flash drum (D-104) and then to the high halide wastewater storage facility (D-201). The remaining decanter water is recycled to the absorber and to the venturi scrubber after it has been cooled to approximately 215°F in the quench water cooler (E-101).

The wastewater contained in the high halide storage tank is processed in a manner similar to that described for the single stage quench system. The capacity requirements for the downstream units were smaller for this case since the maximum amount of condensate to be handled ranged from 5 to 30 percent of the total amount of water vapor contained in the gasifier product gas.

The gas and water vapor mixture coming off the top of the absorber column are sent to a spray cooler (T-102). As described for the single stage quench system, the spray cooler condenses most of the water vapor contained in the absorber effluent gas stream. The gas stream from the spray cooler is sent to an electrostatic precipitator (X-120) for removal of entrained water, tar and particulate matter. Following this, the gas is treated in a Stretford desulfurization unit.

The liquid effluent coming off the spray cooler is also processed in much the same way as in the single stage quench process. Unlike the single quench system, the spray cooler condensate stream contains only a trace quantity of halides. For this reason, the wastewater can be recycled after treatment for use in the gasification circuit without going through a desalination step. The addition of sodium hydroxide to the stripper column is necessary for fixed ammonia removal since most strong acids will be removed in the absorber.

### 4.2.3 First Stage Quench Control Scheme

Figure 4-8 is a schematic of the proposed first stage quench control system. The first stage quench water cooler is designed for the quench water cooling duty necessary to achieve 5 to 30 percent condensation of water vapor contained in the absorber inlet gas stream. The rate of condensation is controlled by adjusting the temperature of the quench water going into the absorber. Changes in the temperature of the quench water are accomplished by adjusting the fraction of quench water that actually goes through the quench water cooler (E-101). The total quench water flow rate is held constant as well as the cooling water flow to the exchanger (E-103) and the quench water flow rate to the venturi scrubber. The control scheme consists of a feedforward system operating in conjunction with primary and secondary feedback loops. The primary feedback loop adjusts the blowdown from the decanter as a function of the level in the decanter. The secondary feedback loop adjusts the temperature of the quench water (i.e. the volume of condensate produced) as a function of the blowdown flow from the decanter. Consequently, the average blowdown rate is determined by the set point on the blowdown flow controller. The set point on the blowdown flow controller is adjusted feedforward from the air/oxygen input to the gasifier. Thus the control scheme is capable of delivering a desired unit flow of blowdown from the first stage quench operation.

The study also developed an alternate control scheme for the single stage quench that possesses the attribute of simplicity but may require additional decanter capacity. In the alternate scheme, the blowdown from the first stage decanter is maintained at a rate proportional to the air/oxygen feed rate to the gasifier. This provides essentially a fixed rate of blowdown for a given gasifier production rate.

**Davy McKee**

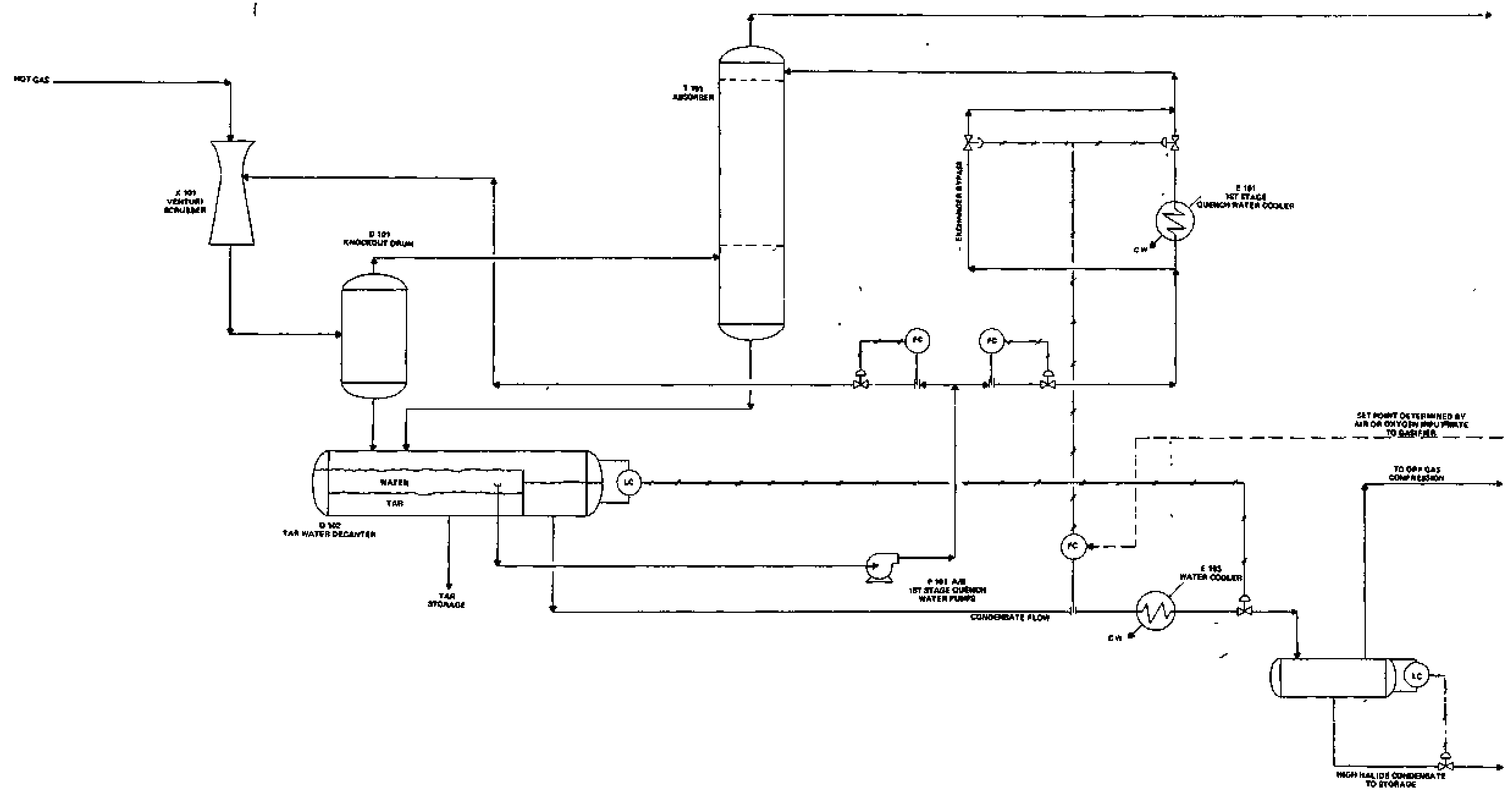

FIGURE 4-8 CONTROL SCHEME – FIRST STAGE QUENCH

The alternate control scheme regulates the temperature of the quench water with a feedback control loop from the level sensor in the decanter to the control valve on the quench water cooler bypass line. When the decanter level drops, the control valve closes to send cooler water into the quench so as to condense more water and thereby fill the decanter. When the decanter level rises, the control valve opens to send warmer water into the quench with the result that less water is condensed. Thus the operation of the decanter as a variable level device enables the maintenance of a fixed blowdown rate from the system for a given gas throughput. Low level and high level override sensors would be installed in the decanter to prevent excessive level fluctuations.

### 4.2.4 Design Basis for Equipment Common to the Single Stage Quench and Two Stage Quench Systems

#### Venturi Scrubbers (X-101)

All the venturi scrubbers were sized to handle the total amount of gas flow to the vessel. The gas and water leaving the vessel were assumed to be at their adiabatic saturation temperature. Typically, 95 percent of the product gas cooling duty is provided by vaporization of the incoming quench water. Additional cooling is achieved by heating the quench water to the adiabatic saturation temperature. The venturi scrubbers were of the variable throat type to provide turndown capability.

#### Knockout Drums (D-101)

The knockout drum was specified to be a vertical, cylindrical vessel with a tangential inlet nozzle. This nozzle configuration provides a cyclone separation effect to separate water, tar, oil and solids from the vapor. The drum's cross-sectional area is sized on the basis of the required entrainment velocity and is calculated using the Souders-Brown expression:

$$V_e = 0.157 \sqrt{\frac{\rho_L - \rho_V}{\rho_V}}$$

$V_e$ = Design entrainment velocity, ft/sec.

$\rho_L$ = Liquid density at conditions, $lb/ft^3$

$\rho_V$ = Vapor density at conditions, $lb/ft^3$

0.157 = Empirical design factor (in-house design practice), ft/sec

## Pumps

All pumps are designed to handle 110 percent of normal flow.

## Electrostatic Precipitator (X-102)

The electrostatic precipitators were designed to handle the flow from all five gasifiers and to operate at conditions of 110°F and 95 psia. The inlet particle loading was specified as 5 $grains/ft^3$ and the removal efficiency at 99 percent. The unit was specified to consist of 300-8" pipes. The electrostatic precipitators were designed for cleaning by solvent and spindle brush, but extensive fouling from the clean gas is not anticipated (1).

### Tar Fouling

Fouling of gas cleaning equipment by deposition of tar is a potential problem associated with tar producing coal conversion processes. The tar handling problem has been investigated in test runs with the METC pilot plant fixed bed gasifier (1). The tests with a recycle quench system condensed dry tar at 375°F. The test results reported that tar discharge through the steam traced tar lock hopper and discharge valve was not difficult. The remaining tar was collected with the quench water and was delivered to a pressurized decanter that operated at 120-140°F. The decanter was reported to function efficiently as a primary separator of particulate solids and low boiling tars.

The conceptual gas trains for the prototype single quench and double quench systems at the present study employed a tar scrubber to collect high boiling tar prior to heat recovery and humidification. The high temperature components of the gas train were insulated to minimize tar deposition on cold surfaces. It is recommended that consideration be given in detailed engineering to the provision of a hot tar washout for units that are indicated as subject to tar fouling. A hot tar washout has been demonstrated as effective for removal of tar deposits at by-product coke plants.

## 4.2.5 Design Basis for Equipment for the Single Stage Quench System

### Spray Coolers (T-101)

It is assumed that the gas leaving the spray cooler will be at a temperature of 110°F. The total cooling duty of the spray cooler consists of cooling gas from its inlet temperature to 110°F and condensing the water vapor from the gas stream.

The spray cooler quench water temperature was set at 105° at the inlet and 130°F at the outlet. These temperatures provide a reasonable heat transfer driving force and quench water flow rate.

The single stage spray cooler is designed to condense out all the water vapor contained in the product gas. Design equations for the spray coolers were obtained from an article by Fair (3).

### Tar Water Decanter (D-102)

The tar decanter in the single stage quench system is designed to handle the total flow of recycle and condensables from the knockout drum and the spray cooler. Decanter volumes are based on the maximum water throughput and liquid residence times of 5 minutes. The decanter is to operate 70 percent full and is designed to operate at conditions of 180°F and 100 psi.

### Flash Drum (D-104)

The flash drum is designed to handle the condensate flow from all five gasifiers. The design temperature and pressure of the vessel are 180°F and 20 psig respectively. The volume of the flash drum is determined by the maximum condensate flow expected and a five minute residence time.

### Single Stage Quench Water Cooler (E-101)

The quench water cooler for the single stage quench was designed on the basis that all the inlet water vapor to the spray cooler is condensed in that unit. The overall heat transfer coefficient specified for this unit was 90 Btu/hr-ft$^2$-°F. Based on METC pilot plant performance, unmanageable fouling of the heat exchanger is not anticipated (1).

### 4.2.6 Design Basis for Equipment for the Two Stage Quench System

Spray Cooler (T-102)

The design criteria for the two stage quench spray cooler is based on a 5 percent water vapor condensation in the absorber. All other design criteria are the same as for the single stage spray cooler (T-101) except that the gas is somewhat cooler at the inlet (280°F).

Tar Water Decanter (D-102)

This vessel is designed to operate at higher temperatures than the single quench system tar-water decanter since it is downstream from the absorber rather than a cooling spray cooler which condenses and cools the incoming water vapor. The vessel is sized to handle 5 to 30 percent range of condensation in the absorber. As was the case for the single stage quench decanter vessel, this unit is also designed for a 5 minute recycle plus condensate residence time and to operate at 70 percent of total capacity.

Tar Water Decanter (D-103)

The design basis for the second stage decanter is the same as for the single stage quench system decanter except for the total volume requirement. Since a portion of the water vapor contained in the product gas is condensed in the absorber, the capacity requirements are a bit lower for this unit.

### Flash Drums (D-104, D-105)

Flash drum D-104 is sized to handle the high halide condensate waste stream consisting of up to 30 percent of gas moisture.

Flash drum D-105 is sized to handle the high volume flow of the reuseable low halide condensate stream.

### Heat Exchangers

#### First Stage Quench Water Cooler (E-101)

The first stage quench water cooler was sized on the basis that 5 to 30 percent of the water vapor going into the absorber is to be condensed. The total heat transfer requirement is that which is necessary to cool the mixture to its dew point and to remove the latent heat released by the condensing water vapor. The quench water flow rate to the absorber is fixed to satisfy the mass transfer requirements necessary to remove the halides. Since this operating parameter is fixed, the quench water temperature is adjusted to meet the absorber cooling requirements.

The quench water cooler is a one shell pass, two tube pass unit with the cooling water specified to flow on the shell side. An overall heat transfer coefficient was estimated to be 100 Btu/hr-ft$^2$-°F (1). This includes an allowance for surface fouling on the tube side. Multiple units were specified. Unmanageable fouling of the heat exchanger is not anticipated.

#### Second Stage Quench Cooler (E-102)

This unit is designed for a cooling duty that is necessary to condense out all the water vapor contained in the spray cooler inlet gas stream when the absorber is operating at 5 percent condensation.

Cooling water is assumed to be at 85°F at the exchanger inlet and 100° at the outlet. An overall heat transfer coefficient of 125 Btu/hr-ft$^2$-°F was used to determine the exchanger surface area. Unmanageable fouling of the heat exchanger is not anticipated.

## Water Cooler E-103

The condensate cooler is sized for 30 percent condensation in the first stage. This exchanger cools the absorber condensate down to 180°F to prevent flashing and emulsion formation.

## Absorbers (T-101)

The absorption towers remove most of the halides contained in the inlet gas stream with minimal condensation of the water vapor. The solubilities of HF and especially HCl are so great that almost complete removal from the halides can be expected in one theoretical stage. For example, at the specified operating conditions in the absorber, HF and HCl have relative volatilities of about 1/100 and 1/100,000 that of $NH_3$ respectively. In order to assume the predicted equilibrium absorption of HCl and HF, the number of theoretical trays in the column are determined for the absorption of $NH_3$. The theoretical stages were stepped off on an X-Y plot for $NH_3$ after assuming a reasonable liquid to gas ratio (about 6). For all column designs, 3 to 4 theoretical trays were required to achieve thorough removal of $NH_3$. An overall tray efficiency of 50% was used in determining the number of actual trays.

The solubility of ammonia in the condensate, or X-Y plot, was developed using the computer program described in Section 4.3.2 by varying the concentration of ammonia over the narrow range of interest while holding all other component concentrations constant.

The absorbers were designed for 5 percent condensation in the first stage quench which resulted in the highest liquid and vapor loadings. Absorber trays were specified to be the sieve tray type. The column diameter was based on the equation given in the F.R.I. manual (4).

## 4.3 Method for Calculating the Quench Unit Material Balances

Material balances for the gas cleaning systems were analyzed for four producer gas compositions derived from the gasification of Illinois No. 6 and Montana Rosebud coals with air or oxygen. All of the gas compositions were considered for the design of both the two stage and single stage quench gas cleaning systems. For the two stage quench design, 5, 10, 20 and 30 percent condensation was considered in the absorber to determine the effect of the amount of condensation on the removal of halides. In each of the proposed designs, the gas sent to the electrostatic precipitator is at a temperature of 110°F.

### 4.3.1 Theoretical Basis for the Material Balance Calculations

The aqueous solution in the quench unit contains a variety of molecular and ionic species. The molecular species considered in the computer program used in this study were $NH_3$, $CO_2$, $H_2S$, HCN, HCl and HF. This system is similar to the weak electrolytic solution modeled by Edwards (5) for prediction of vapor-liquid equilibrium. In the study by Edwards, a computer program (WAVES) was written for computation of the vapor phase composition based on knowledge of the liquid phase composition for the system described. The model proposed by Edwards yields satisfactory results for the temperature range 0 to 100° and for weak electrolyte concentrations not exceeding two molal. Since the quench units in the two gas cleaning systems considered operate above 100°C, modifications were necessary in the WAVES program to adapt it to this study. Also, since sulfur dioxide does not exist in the proposed process, it was eliminated from the program. Since one of the main objectives of this study was to determine the absorption of HCl and HF in the condensate, these molecular species were incorporated into the modified WAVES program.

To solve a multicomponent vapor-liquid equilibrium problem of this type, four different sets of equations must be solved simultaneously as described by Edwards. These include component material balances, a liquid phase charge balance, chemical equilibria, and phase equilibria. For the liquid phase, seventeen chemical species are considered in addition to $H_2O$: $NH_3$ (molecular), $NH_4^{++}$, $CO_2$ (molecular), $HCO_3^-$, $CO_3^=$, $H_2S$ (molecular), $HS^-$, $S^=$, HCN (molecular), $CN^-$, HCl (molecular), $Cl^-$, HF (molecular), $F^-$, $NH_2COO^-$, $H^+$, $OH^-$. The vapor phase contains molecular $NH_3$, $CO_2$, $H_2S$, HCN, HCl, HF and $H_2O$. Vapor phase dissociation of these compounds is not considered since it requires substantially higher temperatures. The vapor phase also contains other products of gasification which contribute to the total system pressure; as a simplifying assumption, these other constituents are considered to be inert and not to enter into the liquid phase.

Material balances can be written for each of the molecular species in the liquid phase.

$$NH_3 = m_{NH_3} + m_{NH_4^+} + m_{NH_2COO^-} \quad (1)$$

$$CO_2 = m_{CO_2} + m_{HCO_3^-} + m_{CO_3^=} + m_{NH_2COO^-} \quad (2)$$

$$H_2S = m_{H_2S} + m_{HS^-} + m_{S^=} \quad (3)$$

$$HCN = m_{HCN} + m_{CN^-} \quad (4)$$

$$HCl = m_{HCl} + m_{Cl^-} \quad (5)$$

$$HF = m_{HF} + m_{F^-} \quad (6)$$

The bracketed species represent the total molality of each molecular component in the liquid phase; and $m_i$, $m_{i+}$ and $m_{i-}$ represent the molecular concentration of each particular form of the species present in the liquid phase.

The second characteristic phenomenon of the system requires electroneutrality of the liquid phase. The total negative charge and positive charge for the ionic species in the quench unit liquid phase must satisfy the following equation:

$$m_{NH_4^+} + m_{H+} = m_{HCO_3^-} + 2m_{CO_3^=} + m_{HS^-} + 2m_{S^=} + m_{CN^-} + m_{Cl^-} +$$
$$m_{F^-} + m_{NH_2COO^-} + m_{OH^-} \quad (7)$$

Each one of these molecular species and water contribute to the overall pressure of the quench units (spray cooler or absorber). The ionic species in the liquid phase do not exert a vapor pressure but do affect the partial pressures exerted by the molecular components in the liquid phase. Since the ionic species do not exert a vapor pressure, it is necessary to know to what extent each molecular species dissociates in the liquid phase. The chemical equilibrium between the dissociated and undissociated species in the liquid phase is expressed in terms of the dissociation equilibrium constant:

$$K_{NH_3} = \frac{(a_{NH_4^+})(a_{OH^-})}{a_{NH_3}} \qquad (8)$$

$$K_{CO_2} = \frac{(a_{HCO_3^-})(a_{H^+})}{a_{CO_2}} \qquad (9)$$

$$K_{HCO_3^-} = \frac{(a_{CO_3^=})(a_{H^+})}{a_{HCO_3^-}} \qquad (10)$$

$$K_{H_2S} = \frac{(a_{HS^-})(a_{H^+})}{a_{H_2S}} \qquad (11)$$

$$K_{HS} = \frac{(a_{S^=})(a_{H^+})}{a_{HS^-}} \qquad (12)$$

$$K_{HCN} = \frac{(a_{CN^-})(a_{H^+})}{a_{HCN}} \qquad (13)$$

$$H_{HCl} = \frac{(a_{Cl^-})(a_{H^+})}{a_{HCl}} \qquad (14)$$

$$K_{HF} = \frac{(a_{F^-})(a_{H^+})}{a_{HF}} \qquad (15)$$

$$K_{NH_2COO^-} = \frac{(a_{HCO_3^-})(a_{NH_3})}{a_{NH_2COO^-}} \qquad (16)$$

$$K_{H_2O} = \frac{(a_{H^+})(a_{OH^-})}{a_{H_2O}} \qquad (17)$$

where $a_i$ represents the activity of species i and is assumed to be 1 for $H_2O$. Ionization constants as a function of temperature for hydrogen chloride and hydrogen fluoride were obtained from Clark (6) and data for other species was taken from Edwards, Mourer, Newman and Prausnitz (7).

The activity of species i, ($a_i$), is related to its molality ($m_i$), through the activity coefficient ($\gamma_i$):

$$a_i = m_i \gamma_i$$

where $\gamma_i \rightarrow 1$ as $\sum_{j=1}^{n} m_j \rightarrow 0$, and subscript j refers to all solute species. The activity coefficient of each species in the liquid phase is obtained from the following expression based on the extended Debye-Huckel Theory (6) (8):

$$\ln\gamma_i = \frac{-aZ_i^2 \sqrt{I}}{1 + \sqrt{I}} + 2\sum_{k \neq w} \beta_{ik} m_k$$

where:

W = Represents water
a = Debye-Huckel proportionality factor
$Z_i$ = Ionic charge of species i

$\beta_{ik}$ = Specific interaction parameter for species i and k
$m_k$ = Molality species k
$I^i$ = Ionic strength of solution defined by:

$$I = 1/2 \sum_{j=1}^{n} Z_j^2 m_j$$

The ik term represents the short range (van der Waals) interactions between species i and k. These interactions are divided into three classes: molecule-molecule, molecule-ion, and ion-ion. Interaction parameters for the species resulting from $NH_3$, $CO_2$, $H_2S$, and HCN were obtained from Edward's program. The ion-ion interaction parameter for $Cl^-$ and $F^-$ were obtained from an article by Bromley (9). Due to a lack of information, the molecule-molecule and molecule-ion interaction parameters for HCl and HF were set equal to zero.

Equilibrium between the vapor and liquid phase for the solutes can be expressed by the following equation:

$$Y_a Ø_a P = m_a \gamma_a H_a$$

where $Ø_a$ is the molecular fugacity coefficient, $Y_a$ is the vapor phase mole fraction, P is the total system pressure, $m_a$ is the molality, $\gamma_a$ is the activity coefficient, and $H_a$ is Henry's constant. For typical operating conditions for the quench units, it was assumed that the fugacity coefficient, $Ø_a$ was equal to one.

The dissociated forms of the weak electrolytes in the liquid phase do not contribute to the vapor pressure of the undissociated electrolytes. For this reason, the vapor-liquid equilibrium expression is written only for the molecular form of the species considered. Thus, the following equations completely describe the vapor-liquid equilibria for the electrolytic species considered:

$$P_{NH} = \phi \, H_{NH_3} \quad (18)$$

$$P_{CO_2} = m_{CO_2} \, \gamma_{CO_2} \, H_{CO_2} \quad (19)$$

$$P_{H_2S} = m_{H_2S} \, \gamma_{H_2S} \, H_{H_2S} \quad (20)$$

$$P_{HCN} = m_{HCN} \, \gamma_{HCN} \, H_{HCN} \quad (21)$$

$$P_{HCl} = m_{HCl} \, \gamma_{HCl} \, H_{HCl} \quad (22)$$

$$P_{HF} = m_{HF} \, \gamma_{HF} \, H_{HF} \quad (23)$$

Henry's constants as a function of temperature for equations (18) through (23) were obtained from the literature (7).

Partial vapor pressure for aqueous hydrogen chloride solutions were taken from the literature (10). Using the ionization constant of HCl and calculating the mean ionic activity coefficient ($\gamma \pm$) by a method described by Bromley (9), the vapor pressure of HCl was correlated to the molecular (non-ionized) concentration of HCl in the liquid as a function of temperature. This molecular Henry's constant for HCl was developed for the temperature range of about 0°-150°C.

Partial vapor pressures for aqueous hydrogen fluoride were also found from the literature (11). The molecular Henry's constant for HF was developed for the temperature range of about 0° to 150°C by using the ionization constant and calculating $\gamma\pm$.

Since water exists as the principal process component, even in the concentrated electrolyte solution exiting the absorber, it was assumed that liquid phase non-ideality effects on the partial pressure of water are small. For this reason, Raoult's Law was used rather than Henry's Law to express the partial pressure of the water present in the liquid phase. According to Raoult's Law, the partial pressure of water ($P_{H_2O}$) may be expressed as:

$$P_{H_2O} = X_{H_2O} \quad P_{H_2O}$$

where $X_{H_2O}$ = mole fraction of $H_2O$ in the liquid phase

$P_{H_2O}$ = vapor pressure of water

Additional complexity is introduced into the system due to the presence of tars and phenols. The quantity of tar present in each quench was determined from distillation curves of a stored tar mixture (12). The molecular weight of the tar was taken from a typical tar analysis (12). It was assumed that the tar exists as a separate liquid rather than as an emulsion with the condensed water.

Calculations were done to determine the amount of phenol present in the organic liquids, the water condensate, and in the vapor phase. These calculations were based on the total number of moles of condensed tar, the dissociation constant for phenol, Henry's constant for phenol at system temperature, the hydrogen ion concentration, the total amount of liquid water present, the total number of moles in the vapor phase and the total system pressure. The ionization constant for phenol and the phenol-water vapor pressure data were found in the literature (13, 14, 15). A molecular form of Henry's constant was developed for phenol for the temperature range 0 to 150°C based on these data.

No attempt was made to estimate the degree to which the organic liquid phase forms an emulsion with the water phase. It was assumed that the organic liquid and water layers existed as separate fluids in the quench units.

### 4.3.2 Computer Algorithm for Executing the Material Balance Calculations

As mentioned in Section 4.3.1, several modifications in the WAVES computer program were necessary to improve convergence and to more closely simulate the environment of the quench units. Other changes were necessary besides the addition of components HCl and HF and the removal of $SO_2$. Correlations were developed for the ionization constants and Henry's constants for a temperature range of 0° to 170°C. This was necessary since the operating temperature of the quench units exceeded the 0 to 100°C temperature range which the constants had been correlated for in the original program.

Also included in the modified WAVES program were correlations to account for the vapor-liquid equilibria exhibited by phenols and tars over this temperature range.

Equations (7) through (23), plus associated expressions for activity coefficients, comprised the 34 independent equations necessary to solve the molality and activity of each species in the liquid phase for a known vapor phase. If the total molality of the liquid phase is known and the vapor phase is to be calculated, then equations (18) through (23) are replaced by equations (1) through (6).

The computer algorithm block diagram is shown in Figure 4-9. As shown in the flow chart, the required computer input includes: system temperature and pressure, total moles in the gas phase after absorption, weight of water in the liquid phase, and total moles of $NH_3$, $CO_2$, $H_3S$, HCN, HCl and HF before absorption. From this input, the equilibrium liquid and vapor split of each component was computed. With this information, the mole fraction of water in the liquid phase was calculated.

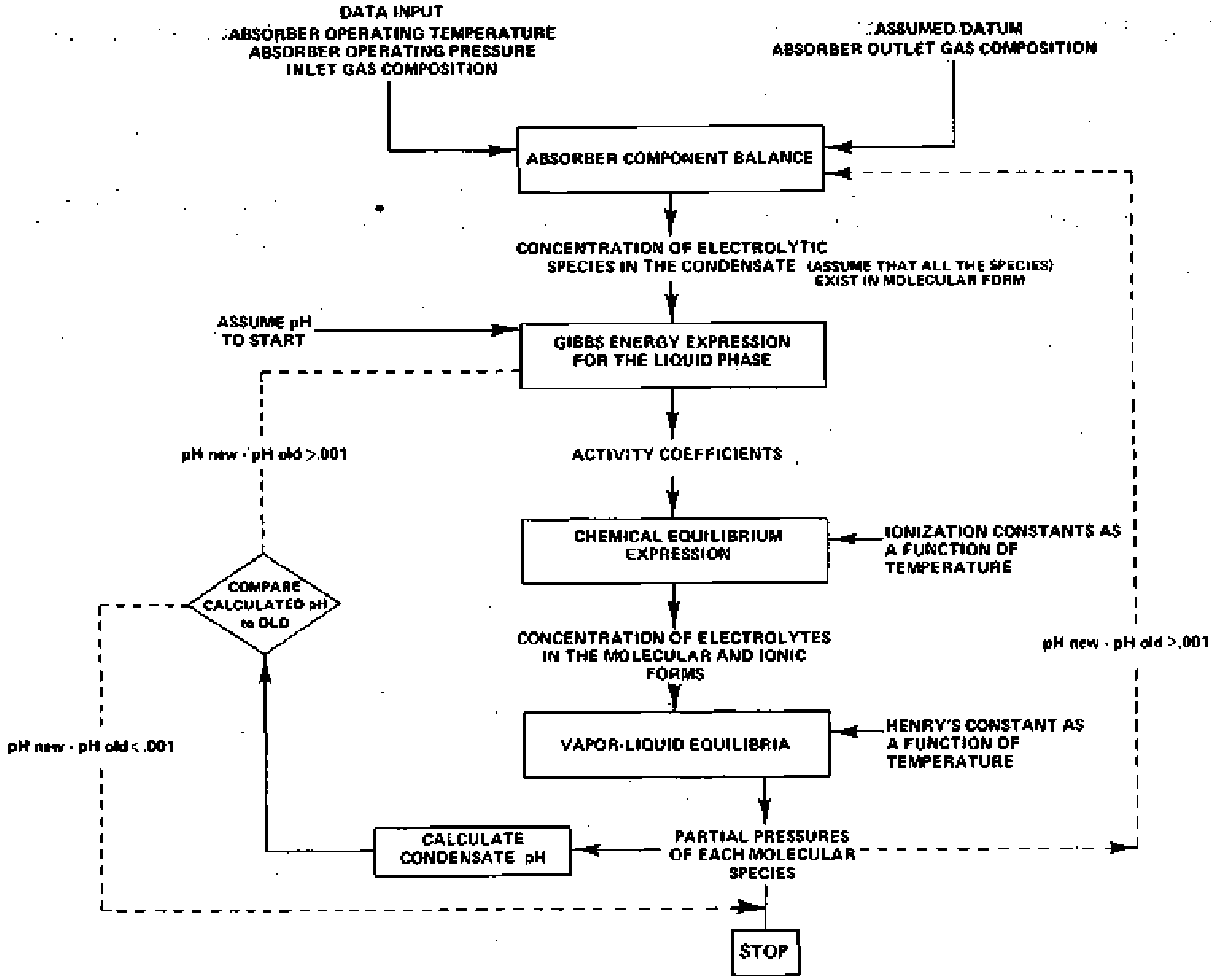

FIGURE 4-9: COMPUTER FLOW CHART FOR CALCULATING VAPOR-LIQUID EQUILIBRIA

One may refer to Figure 4-9 to follow the technique used to solve the given set of equations. All of the ionic species in the charge balance equation (Eq. 7) can be expressed in terms of the hydrogen ion concentration $m_{H+}$, the ionization and Henry's constants, the partial pressures and activity coefficients of the pertinent molecular species. This results in a cubic equation in terms of the hydrogen ion concentration. For a given temperature, ionization and Henry's constants are calculated using the proper correlations. Initial estimates of partial pressures are given and activity coefficients are initially set equal to one. Using a Newton-Raphson convergence technique, the hydrogen ion concentration is determined. From this hydrogen ion concentration, all activity coefficients and partial pressures are evaluated. The partial pressures are determined by an expression based on the total moles of each component existing in each phase, the total pressure of the system, the total moles in the gas phases, and the total water in the liquid phase (all of these are specified as computer input). The newly evaluated activity coefficients and partial pressures are substituted back into the cubic equation for recalculation of the hydrogen ion concentration. The iteration process is continued until convergence of the partial pressure values is attained.

## 4.4 Engineering of Condensate Stripper Modules

### 4.4.1 Description of the Single Stage Quench Wastewater Stripper (Cases III & IV)

The total condensate output from all five gasifiers is collected in the wastewater storage tank (D-201). As shown in Figure 4-10 the condensate is pumped through a preheater (E-201) before entering the stripper column (T-102).

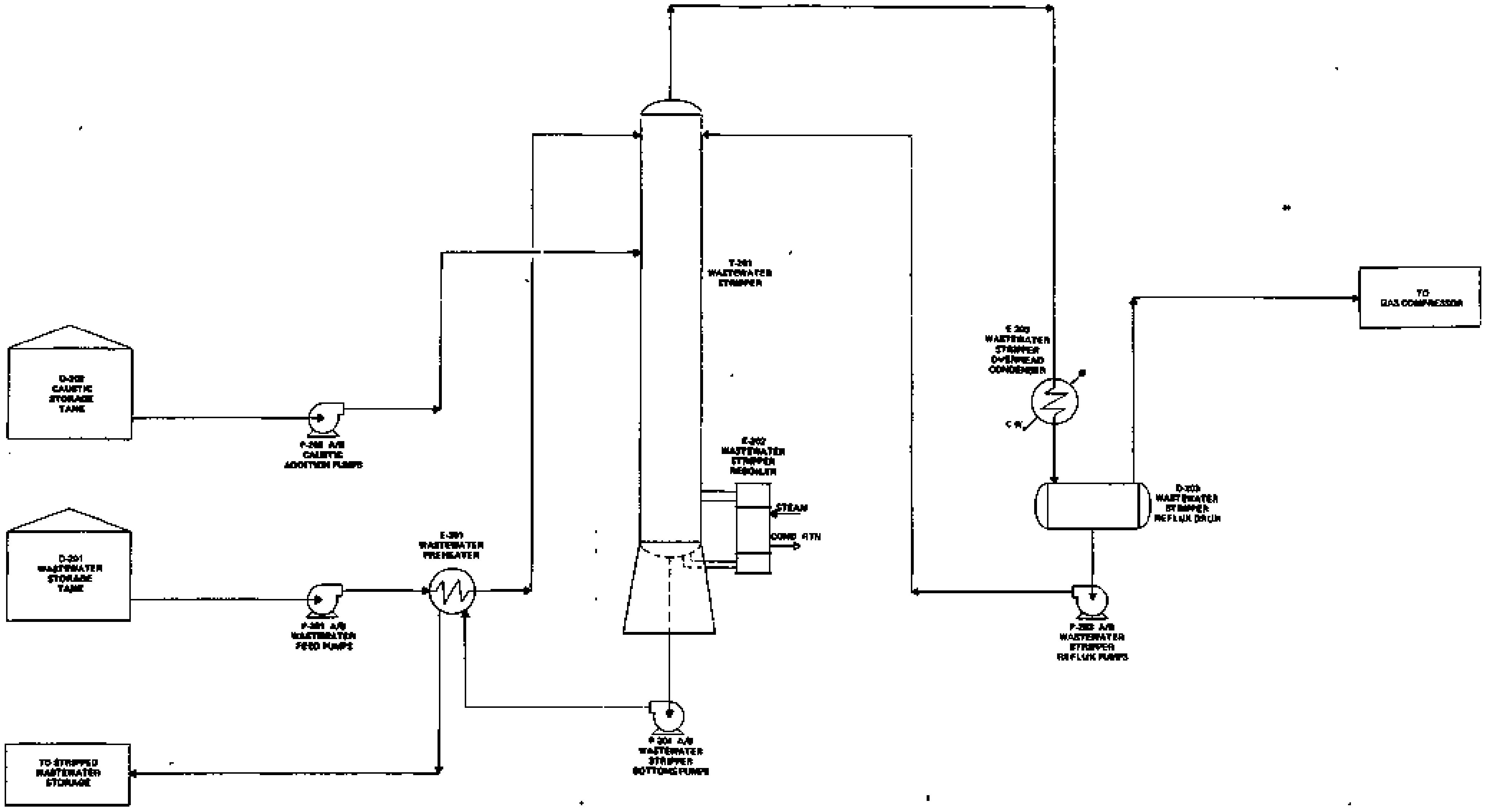

FIGURE 4-10 CONDENSATE STRIPPER – SINGLE STAGE QUENCH SYSTEM

The condensate entering the stripping column contains a considerable amount of volatile acid gases. These contaminants are removed from the condensate stream as it passes down the column and makes contact with a counter-current flow of steam. The stripping steam is provided by vaporizing a portion of the wastewater with the wastewater stripper reboiler (E-202). At some point in the column, the acid gases are reduced to a concentration such that the ammonia may be removed by the addition of caustic. The tray at which injection of caustic takes place was determined by equilibrium calculations done for the column.

By the time the wastewater stream leaves the stripper, the ammonia and other contaminant concentration levels have been reduced substantially. Estimates of these concentrations are shown in the condensate stripper result section. The stripped wastewater stream is then sent to a heat exchanger (E-201) to preheat the incoming condensate. Following this, the wastewater undergoes further processing in the biological treatment plant.

The vapor composition leaving the stripping column passes through an overhead condenser (E-203) to remove the water vapor. This condensate stream, concentrated with $NH_3$, $CO_2$ and $H_2S$, is recycled to the feed tray of the column. The vapors leaving the reflux drum (D-203) are sent to a compressor where they are compressed and injected back into the gas product stream leaving the electrostatic precipitator.

### 4.4.2 Description of the Two Stage Quench Wastewater Stripper (Cases I & II)

The wastewater stripper processing scheme (see Fig. 4-11) for the two stage quench system is almost identical to that of the single stage quench system. The difference being the need for

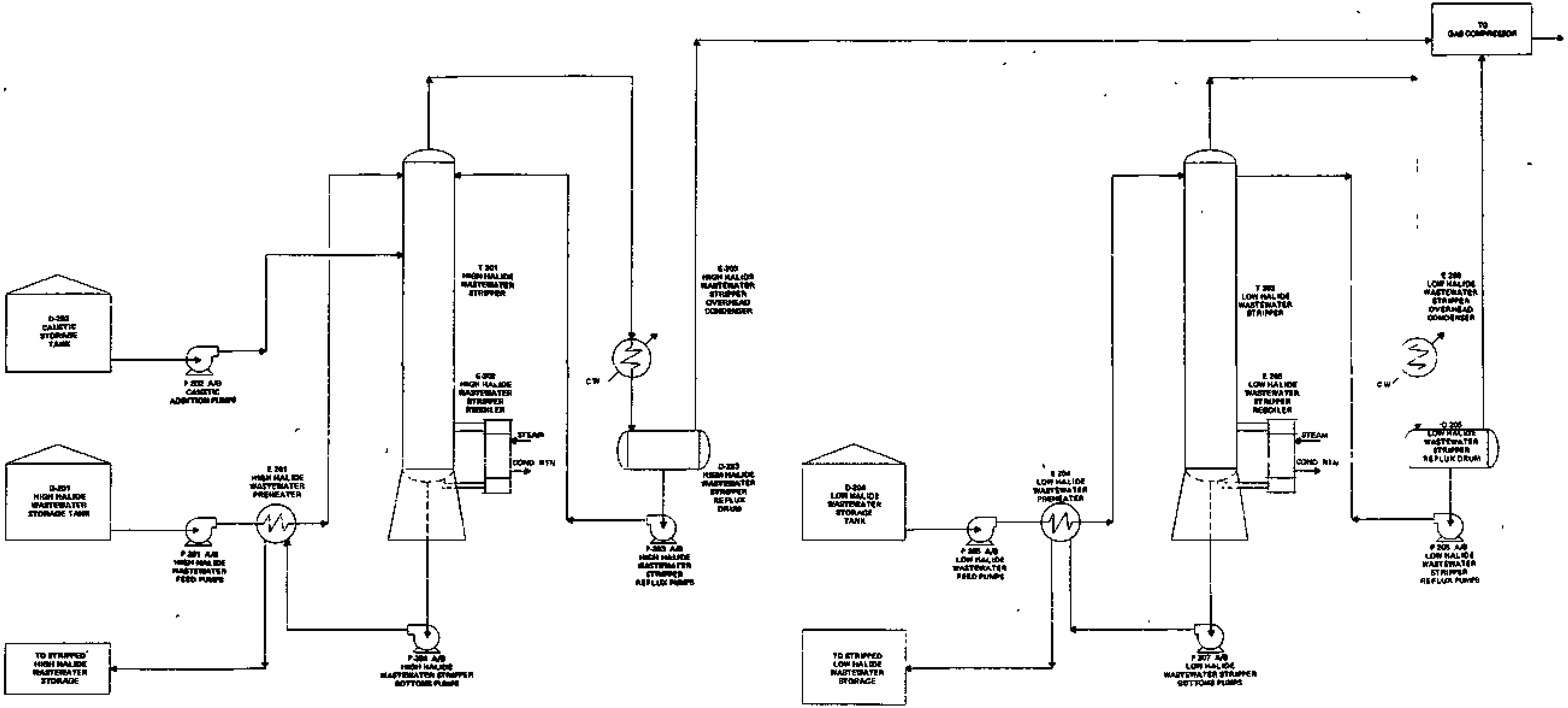

FIGURE 4-11 CONDENSATE STRIPPERS – TWO STAGE QUENCH SYSTEM

two process units; one stripper column to handle the high halide condensate and the other to handle the low halide condensate stream. One additional difference arises in the low halide condensate stripper unit. Based on prediction of acid concentrations in the second stage condensate, it appears that the injection of caustic is not necessary for the low halide wastewater stripper.

### 4.4.3 Wastewater Stripper Design

The wastewater stripper column design for the single stage quench process was based on the total condensate flow from the five gasifiers. For the case of the two stage quench system, the high halide wastewater stripper design was based on 10 percent condensation in the first stage quench. The low halide wastewater scrubber design was based on the condensate flow from the second stage quench when 10 percent condensation takes place in the first stage quench.

Based on similar stripper designs, 0.12 pounds of stripping steam per pound of wastewater feed was used for all the stripper designs. The total steam demand for the stripper column was determined by the stripping steam and heating steam requirements. The heating steam requirements were based on the amount of preheating necessary to raise the column feed to the operating temperature of the feed tray. The column feed includes the wastewater feed and the reflux. It assumed that all of the heating steam condenses on the feed tray. Thus, the liquid flow down the column consists of the wastewater feed, the reflux stream, and the condensed heating steam.

Column diameters for the stripper were sized based on the sieve tray correlation given in the F.R.I. manual (4). The overhead condenser pressure of all strippers was set at 18 psia and 0.2 psi pressure drop per theoretical tray was assumed. For each design, a 60 percent tray efficiency was used based on sieve tray efficiency correlations (4) and other sources as part of the design bases, tray temperatures were assumed to be that of saturated steam at the prevailing pressure.

To specify the stripper design requirements, a different version of the modified WAVES computer program is used. With the modified version, the vapor phase is specified and the composition of the corresponding liquid phase is predicted. In determining the specifications for the stripper units, the following constraints are assumed for the stripper bottom contaminants:

$NH_3$ < 50 PPM
$H_2S$ < 5 PPM
HCN < 2 PPM

Knowing this, the total number of moles of $NH_3$, $CO_2$, $H_2S$ and HCN leaving the reflux drum were estimated by means of a component balance. The operating conditions of the reflux were set at 170°F and 17 psia so that most of the water vapor condensed out of the overhead. The amount of liquid water in the reflux drum and water vapor leaving the drum were determined from the stripping steam rate, moles of $NH_3$, $CO_2$, $H_2S$, and HCN leaving as vapor, vapor pressure of water at 170°F, and total system pressure of 17 psia. The computer program was used to predict the equilibrium concentrations of $NH_3$, $CO_2$, $H_2S$, and HCN in the reflux for the vapor leaving the reflux drum.

The sum of the reflux drum vapor stream components and the liquid effluent stream components comprise the overhead vapor stream from the stripper column. The wastewater feed stream to the stripper and the stripper reflux stream constitute the total input flow to the stripper feed tray. Using this information, a component balance was calculated for the feed tray to determine the composition of the vapor coming from the second tray. Knowing the vapor composition, the composition of the liquid stream leaving the second tray was determined by using the vapor-liquid equilibrium computer program. A component balance was then calculated for the second tray to determine the composition of the vapor entering the second tray. The equilibrium calculation was repeated for this tray and the composition of the liquid leaving the tray was determined. This procedure continued until the composition of the liquid product stream from any given tray met the constraints initially set in the program.

Tray-by-tray calculations were done for one low halide and one high halide stripper based on the condensate obtained from the Illinois No. 6 case - air blown. The condensate properties of all other cases were similar enough that additional tray-by-tray calculations were not necessary.

### 4.4.4 Design Basis for Equipment

Storage Tanks (Wastewater D-201, D-204; Caustic D-202)

All wastewater stroage tanks were sized for a 16 hour capacity. The caustic storage tank was sized for a 30 day supply of a 20 weight percent caustic solution.

Pumps

Wastewater Feed Pumps P-201A/B, P-205A/B
Stripper Bottoms Pumps P-204A/B, P-205A/B
Caustic Addition Pump P-202A/B
Reflux Pumps P-203A/B, P-206A/B

All pumps contained in the stripper section were sized for 110% of normal flow and each pump was specified with a spare.

## Centrifugal Compressors

A two-stage centrifugal compressor complete with a cooling unit after the first stage was specified to handle all the vapors coming from the stripper reflux drum. The unit is designed to compress the vapors from 17 psia to 75 psia for injection into the gas product line.

## Reflux Drums (D-203, D-205)

All reflux drums were specified to be horizontal and to have a five minute liquid residence time at operating conditions of one-half full.

## Heat Exchangers

### Stripper Column Preheaters (E-201, E-204)

The feed preheaters for the low halide wastewater stripper feed stream (E-204) and the single stage quench stripper feed stream (E-201) are designed to recover heat from the stripper bottoms stream. The wastewater feed is heated from about 110°F to 210°F as the stripper bottoms are cooled from about 236°F to 130°F.

The high halide wastewater preheater (E-201) was also designed to recover heat from the stripper bottoms stream. The high halide feed is preheated from about 170°F to 210°F as the stripper bottoms are cooled from about 235°F to 190°F.

The first two exchangers mentioned (E-204, E-201) were specified to be one shell pass two-tube pass types, and five exchangers in series were used to prevent temperature crossing. For the third exchanger mentioned (E-201) two exchangers were connected in series to prevent temperature crossing. For all the preheater designs, the surface area was determined on the basis of a 100 Btu/hr ft$^2$°F heat transfer coefficient. For each case, the bottoms liquid was specified to be on the shell side.

Reboilers (E-202, E-205)

All the stripper reboilers were specified as the vertical thermosyphon type. It was assumed that 50 psig saturated steam would be available. An overall heat transfer coefficient of 200 Btu/hr ft$^2$°F was used for determining the overall heat transfer area.

Overhead Condensers (E-203, E-206)

The overhead condensers were designed to partially condense and cool the stripper overhead vapors from about 212°F to 170°F. A one shell pass-two tube pass exchanger was specified with a heat transfer coefficient of 150 Btu/hr ft$^2$°F for determining the surface area. Cooling water is contained on the shell side with an inlet temperature of 85°F and an outlet temperature of 110°F.

## 4.5 Desulfurization and Utilities

All gasifier alternatives were presumed to employ a Stretford plant for desulfurization of the cooled gas. The Stretford plant provided for hydrolysis of COS and provided for containment of purges from the cyanide scrubber and from the sulfide oxidizer (sorbent).

## 4.6 Activated Sludge

A wastewater treatment concept was developed to biologically treat the bottoms from the ammonia stripper by the activated sludge process. The process flowsheet illustrated in Figure 11-B1 (Section 11, Appendix B) included storage of inflow, cooling, aeration, settling and dual media filtration. The food to microorganism ratio was 0.05 pounds of phenol per day per pound of mixed liquor volatile suspended solids. The aeration tank temperature was maintained between 75 and 90°F. Excess sludge from the bioreactors and solids from filter backwash were concentrated by dissolved air flotation prior to dewatering on belt presses. The pressed sludge cake was spread on the coal pile for feed to the gasifier. The performance of the activated sludge plant was estimated from correlations with waste treatment results at coke plants and gasification plants.

Levels of hydrogen sulfide and phenol in activated sludge effluent were taken as 0.1 mg/l based on representative process performance at coke plants. Process performance is less predictable in regard to cyanide and thiocyanate. For cyanide inputs of over 2.0 mg/l, effluent residuals were taken as 2.0 percent of input values or 0.7 mg/l, whichever was higher. For cyanide inputs of less than 2.0 mg/l, effluent residuals were taken as 70 percent of input values or 0.2 mg/l, whichever was higher. Effluent thiocyanate residuals were estimated as 10 mg/l for input levels above 50 mg/l. For input levels of less than 50 mg/l, effluent thiocyanate levels were estimated as 20 percent of input values or 2.0 mg/l, whichever was higher. Sulfur in converted hydrogen sulfide and converted thiocyanate was assumed to report to the effluent as sulfate.

The fate of ammonia in biological treatment is difficult to evaluate because it may be generated in the process and there are several pathways by which it may leave the process. Some ammonia is converted to organic nitrogen as a component of cellular growth. Thiocyanate, cyanide or amines may be degraded with release of ammonia to the mixed liquor. Some ammonia will be volatilized and escape with the offgases. Ammonia may undergo nitrification and oxidized nitrogen may undergo denitrification with release of nitrogen to the offgases.

For purposes of performance projection it was assumed that the ammonia synthesized to organic nitrogen was equivalent to 0.014 times the phenol in the feed (16). A 20 percent loss of residual free ammonia to nitrification or to volatilization was assumed. In addition it was assumed that 70 mole percent of the thiocyanate degraded was recovered as ammonia. The relation for estimation of effluent ammonia became:

$$\text{Effluent } NH_3 = 0.8\ (\text{Influent } NH_3 - 0.014 \times \text{phenol}) + 0.21\ (\text{SCN degraded})$$

## 4.7 Disposal Concepts

### 4.7.1 Store and Discharge

A concept was prepared to receive treated effluent from the activated sludge process for disposal by store and discharge. The store and discharge concept illustrated in Figure 11-C1 (Section 11, Appendix C) provided for storage of the waste in lined lagoons for a period of up to 4 months so as to enable avoidance of discharge during periods of low flow or when the chloride concentration in the receiving stream was above 250 mg/l. The concept would be inapplicable in the area of the Carson City site and would be of uncertain applicability at the Wheeling site - depending upon regulatory negotiations.

### 4.7.2 Store and Barge

A concept was prepared to receive treated effluent from the activated sludge process for disposal by store and barging. The concept illustrated in Figure 11-D1 (Section 11, Appendix D) provided for storage of the waste in lined steel tankage for a period of up to 28 days to accommodate barge scheduling during worst conditions. The barge would transport the waste from the Wheeling site downriver to the open sea for disposal. Although the disposal method is favored by logic, the applicability of the concept would be conditional to regulatory negotiations. The concept would be inapplicable to Carson City.

### 4.7.3 Deep Well Injection

A discharge concept was prepared for treated activated sludge effluent that featured storage, filtration and deep well injection. From the geological standpoint it was determined that the concept is more applicable to the Wheeling area than to the Carson City area. The applicability of the concept would be subject to regulatory negotiation. The study determined that representative cost estimates were unobtainable, therefore, the economic evaluation of the concept was discarded.

### 4.7.4 Pond Evaporation

A concept was prepared for disposal of treated activated sludge effluent in the Carson City area by pond evaporation. Disposal by pond evaporation is infeasible in the Wheeling area. The Climate Atlas of the U.S. reports for the Carson City area in annual lake evaporation of 40 inches and an annual precipitation of 8.43 inches for a net evaporation of 40-8.43=31.6 inches per year. The evaporation ponds were lined to restrict percolation and were designed to provide a net evaporation of 1.25 times the inflow plus storage for 5 months.

### 4.7.5 Thermal Evaporation

Evaporation concepts were developed for wastewater concentration for wastewater disposal. On concept employed a 14-effect vertical tube evaporator to concentrate feeds to a battery of 10 percent dissolved solids for disposal.

The second evaporator concept employed a kettle evaporator to produce a bottom of 60 percent dissolved solids for feed to a wastewater incinerator for disposal. The incinerator produced a dry salt product (ash).

### 4.7.6 Reverse Osmosis

A concept was prepared to desalinate treated activated sludge effluent by reverse osmosis as illustrated in Figure 11-F1 (Section 11, Appendix F). The concept included storage, chlorination, cartridge filtration, dechlorination, acidification and membrane filtration. The performance of the system was estimated from correlation to operating results reported for cooling tower blowdown (17). The permeate was reused as service water and the concentrate was sent to disposal.

## 4.8 Cooling Towers

Concepts were prepared for evaporative cooling tower systems. The cooling tower analyses employed a drift loss of 0.01 percent and took evaporation as 0.08 percent per °F of range (18).

The cooling towers were operated in accordance with water quality criteria that limited halide levels to 600 mg/l as chloride and total strong electrolyte levels to 1700 mg/l as sodium chloride.

## 4.9 Water Quality Criteria

The establishment of water quality criteria for cooling system recycle circuits involves consideration of costs of replacement of system components and materials of fabrication versus costs of water/blowdown treatment. Consequently definitive criteria applicable to all situations are unavailable. However, for guideline purposes, "normal" conditions for cooling tower operation have been defined as a circulating water with a pH between 6 and 8, a chloride content below 750 mg/l as NaCl, a total dissolved solids content below 1500 mg/l and a maximum temperature of 130°F (18). In addition, the water quality criteria presumed adequate water treatment to minimize corrosion and scaling.

Water quality guidelines advanced by Kunz et al (19) for cooling tower operation specified maximum total dissolved solids at 2500 mg/l and maximum conductivity as 4000 micromhos/cm. For purposes of the present study, water quality criteria for cooling tower circuits were adopted limiting concentration of halides to 600 mg/l as chloride and total strong electrolytes to 1700 mg/l as sodium chloride.

Cooling towers and pipe systems designed for sea water service can accommodate electrolyte concentrations up to the point of scaling. These systems often utilize fiberglass components and polypropylene fill. Metallic components such as heat exchangers would be fabricated in titanium or comparable material.

Applicable water quality criteria for the Ohio River in the Wheeling area were taken from the amended Stream Quality Criteria adopted by the Ohio River Valley Water Sanitation Commission on September 9, 1977. The following stream quantity limits were deemed applicable to the evaluation of a store and discharge concept.

| | |
|---|---|
| pH | 6 to 9 |
| Dissolved Oxygen | 5.0 mg/l |
| Dissolved solids | 500 mg/l |
| Chlorides | 250 mg/l |
| Ammonia, un-ionized | 0.05 mg/l |
| Cyanide | 0.025 mg/l |
| Fluoride | 1.0 |
| Phenol | 0.01 |

For control purposes the minimum 7-day flow once in ten years, the "critical" flow, is taken as 6600 cfs at Wheeling, West Virginia.

In addition, Pollution Control Standard No. 1-70 of the Ohio River (November 30, 1970) Valley Water Sanitation Commission specifies a limiting concentration of 0.2 mg/l of cyanide in industrial and domestic effluents.

A store and discharge concept was deemed infeasible for the Carson City area.

## 4.10 Equalization and Comparative Economic Analysis

A cost comparison between single and two-stage quench systems was desired. The cooling tower circuits in conjunction with water quality criteria were employed to establish a comparable basis for a cost comparison. The procedure involved the use of reverse osmosis treated activated sludge effluent from single quenching and activated sludge treated second stage condensate for makeup to similar cooling tower circuits. The cooling tower circuits were operated at the limits of the water quality criteria (600 mg/l halides as Cl, or 1700 mg/l total electrolyte as NaCl). Additional side stream reverse osmosis capacity was added to the lower performance system until equal blowdown flows from the cooling tower circuits were achieved. This situation provided the basis for a comparable cost comparison biologically treated secondary condensate and single stage condensate treated by activated sludge and reverse osmosis.

The projected performance of the reverse osmosis system in fluenced the blowdown rate from the cooling tower systems. Based on published values of reverse osmosis system performance (17), (20), the following values of percent rejection were employed as representative of expected performance of reverse osmosis on components of gas condensates: $Na^+$ - 85%, $Cl^-$ - 83%, $SO_4^=$ - 98.8%.

The establishment of equalized performance situations between single stage and two stage gas cleaning systems enabled the development of a realistic cost comparison between the systems. The capital and operational cost estimates for the alternative systems, including gas cleaning, wastewater treatment, and supplemental reverse osmosis treatment, were determined on a difference basis.

Annualized capital cost differences were determined from an incremental difference in capital cost using a return on investment of 15 percent and a productive life of 20 years. The annualized capital cost differences were added to the yearly operating cost differences to obtain an indication of the relative economic advantage of the alternative single quench and double quench systems. An illustration of the methodology involved in the analysis is presented in Appendix G.

METHODOLOGY
ECONOMIC

## 5.0 METHODS - COST ESTIMATES

The objective of the economic analysis was to obtain order of magnitude cost estimates relative to the selected producer gas systems that would compare costs associated with single stage gas quenching with two stage gas quenching. Several procedures of estimation were employed in the attainment of the objective. All estimates were based on a mid-1978 price level.

Equipment costs were obtained from the Davy McKee Equipment Cost Library, from budget type estimates from suppliers, and from estimating reference documents (21). Costs associated with civil structures and transmission piping were based on quantities taken from engineering sketches and in-house cost information. Costs of erection, electrical service, instrumentation, field expenses and project services were estimated by factoring from experience at average installations. The methodology employed was generally similar to that proposed by Guthrie (22). The estimating procedure was amenable to adjustment for alternative materials of fabrication.

Engineering flow sketches and sized equipment lists were prepared for gas train components and the ammonia stripper systems. Eight module estimates were prepared for various facility situations involving gas quenching and ammonia stripping.

Estimates for wastewater treatment by the activated sludge process were prepared from engineering flow sketches and equipment lists for three flow and organic loading situations. For estimation purposes the components of the activated sludge system were segregated into cost modules correlated to input flow or input phenol loading. Estimates were prepared for flow/phenol situations pertaining to 15 gpm/370 lb per day, 62 gpm/1600 lb per day and 600 gpm/11000 lb per day. Components allocated to flow included input wastewater storage, clarifiers, deep bed filters, wastewater pumping facilities and control buildings. Components allocated to phenol loading included aeration tanks and mechanisms, as well as sludge dewatering and disposal facilities. The segregated cost modules facilitated the factoring of estimates of activated sludge systems to intermediate flow and loading situations.

Estimates for processing effluent from the activated sludge system by reverse osmosis, by evaporation and by incineration were prepared from engineering flow sketches and equipment lists for 20 gpm, 100 gpm and 600 gpm flow situations. The cost of process equipment required for the estimates was obtained primarily from solicited budget type quotations from vendors. Estimates for intermediate flow situations were obtained from correlations of installed cost to flow.

Estimates for disposal of effluent from the activated sludge system by pond evaporation, by store and discharge to river, and by store and barge to sea were prepared from engineering flow sketches and equipment lists for three flow situations. The pond evaporation estimates pertained to flows of 16 gpm, 80 gpm and 480 gpm. The store and discharge and the store and barge estimates pertained to flows of 20 gpm, 100 gpm and 600 gpm. The estimates involved primarily civil work and structures. Quantities were taken from the sketches and estimated at in-house unit prices. The direct costs were factored to a total installed cost basis. Estimates for intermediate flow situations were obtained from correlations of installed cost to flow.

Estimates for cooling tower facilities were obtained from conceptual engineering cooling system analyses coupled with solicited budget type vendor quotations of installed costs.

The estimated difference in the cost of operation of alternate facilities was determined from values repeated in the literature or by synthesis from the sum of the aggregate estimated cost for payroll (including fringe costs), utilities and other services (e.g. scavenger), chemicals and other supplies, overhead (management, accounting, purchasing, legal, etc.), with allowance of credit for reusable or marketable by-products. Payroll costs and overhead were estimated from staff requirements selected on the basis of the complexity and size of the systems involved. The costs for utilities, services, chemicals, supplies and allowances for credits were estimated from considerations relative to material balances and energy requirements. Maintenance costs were estimated as a percentage of capital cost, with the percentage factor weighted to the characteristics (mechanical, electrical, component serviceability, etc.) of the system. Thus the estimates were synthesized on the basis of research, analysis, judgment and experience.

Some comment is appropriate regarding the quality of the estimates generated in this study. The Contract called for the generation of a large number of estimates in a short time and on the basis of limited field study or analysis of definitive engineering drawings. The estimates were therefore generated primarily from flow sheet engineering concepts which provide a reasonable cost basis for only a portion of an estimated cost. The remaining portion of the cost estimate was based on conceptual design, average conditions and judgements. Two principal cost areas involving extensive application of judgement and experience were piping and electrical. Without an engineered layout or definitive piping system information, it was necessary to employ judgment estimates of the magnitude and complexity of the piping systems. In the case of electrical service components, it was necessary to make judgments, based on power requirements and other considerations associated with the concepts, as to the inclusion of transformers and other power supply costs. Major mechanical facilities such as cooling towers, pumps, thickeners, filters, etc., were based upon price information obtained from equipment suppliers or the Contractor's file information, adjusted to the specific situation. Routine pile foundations were assumed for heavy facilities - simple spread foundations were assumed for light construction. Calculated estimates were extended by factoring in certain applicable situations. In some instances, facilities such as buildings, were estimated as modules instead of by component cost classification. The estimates are intended to apply to the mid-1978 cost level and include construction management and a contingency factor of 10 to 15 percent.

The estimating procedures employed in this study fall primarily within the classifications listed in Perry (23) for "study estimates" (±30 percent) and "order of magnitude estimates" (±40 percent). The Contractor used his best efforts to provide estimates well within the stated ranges.

RESULTS

## 6.0 RESULTS

The complete results of the gas train material balance studies are presented in Section 11.0, Appendix A.

The higher halide and sulfur contents of the selected Illinois No. 6 coal described in Table 4-4 as compared to the selected Montana Rosebud coal yielded higher concentrations of HCl and HF in the gasifier raw gas as indicated in Table 4-5. Residual concentrations of HCl and HF in the cooled gas leaving the gas train were less than 0.0001 percent of gasifier gas concentrations. Thus halides were projected to be substantially removed by effective wet gas cleaning. For a given coal, the cooled gas leaving the electrostatic precipitator had essentially the same composition whether processed by single stage or two-stage quenching.

### 6.1 Gas Cleaning Systems

Figure 6-1 illustrates the predicted relationship for HCl absorption as a function of the water condensed in the first stage quench for an air blown gasifier. The combination of the venturi scrubber (15 in wg. pressure drop) plus the sieve tray absorber with 9 trays was indicated to remove 99.9999 percent of the chlorides from Illinois No. 6 producer gas at a condensation level of 20 gallons per ton of coal fed. This constituted essentially complete removal of chlorides in the first 20 gallons (7 percent) of a total of 290 gallons of condensate produced per ton of coal fed. Somewhat higher absorption of HCl was projected for operation with Montana Rosebud due to increased alkalinity of the condensate.

The predicted removal of HF as a function of volume of condensate for air blown gasifier is presented as Figure 6-2. With producer gas from Illinois No. 6 it was projected that a first stage quench of 30 gallons per ton would absorb over 98 percent of the HF from

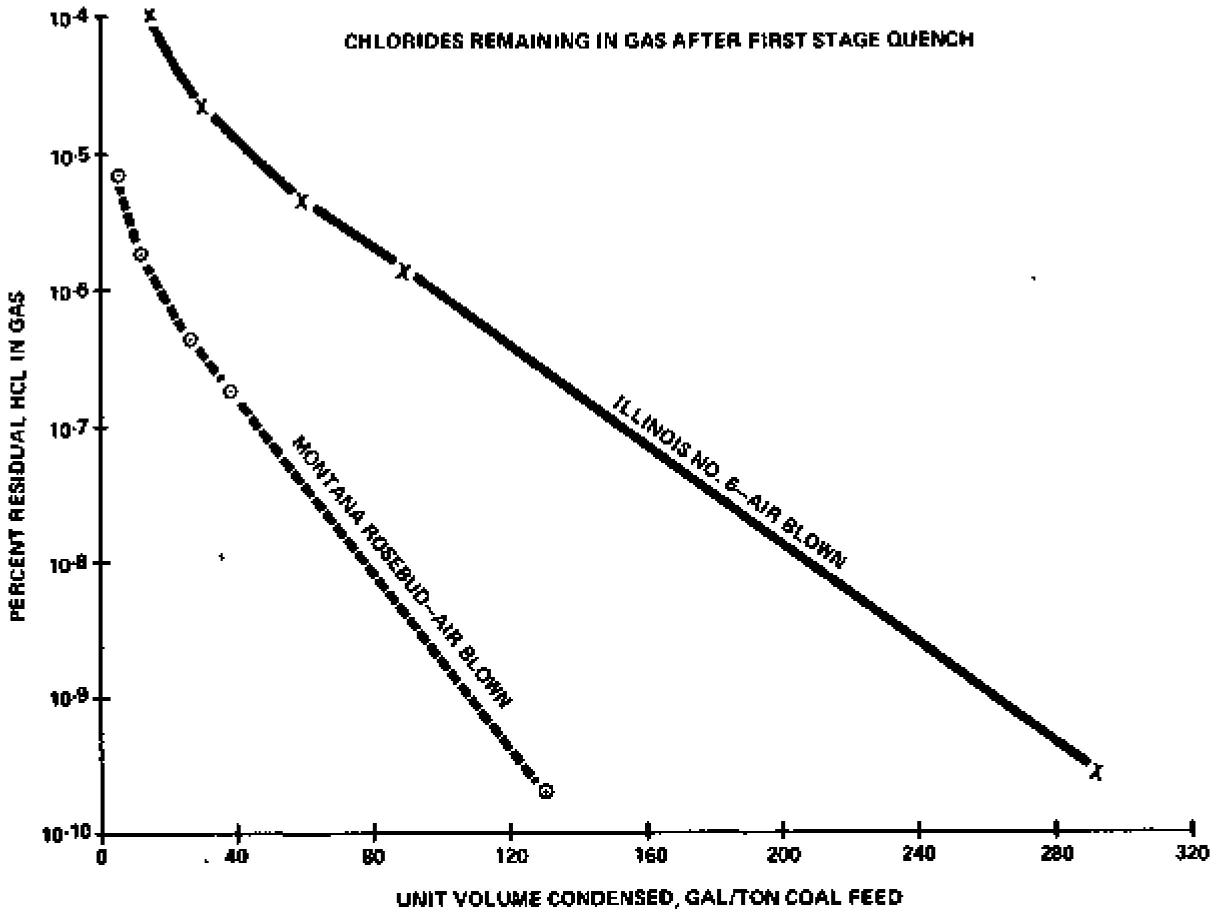

FIGURE 6-1: EFFECT OF UNIT VOLUME CONDENSED ON ABSORPTION OF CHLORIDES

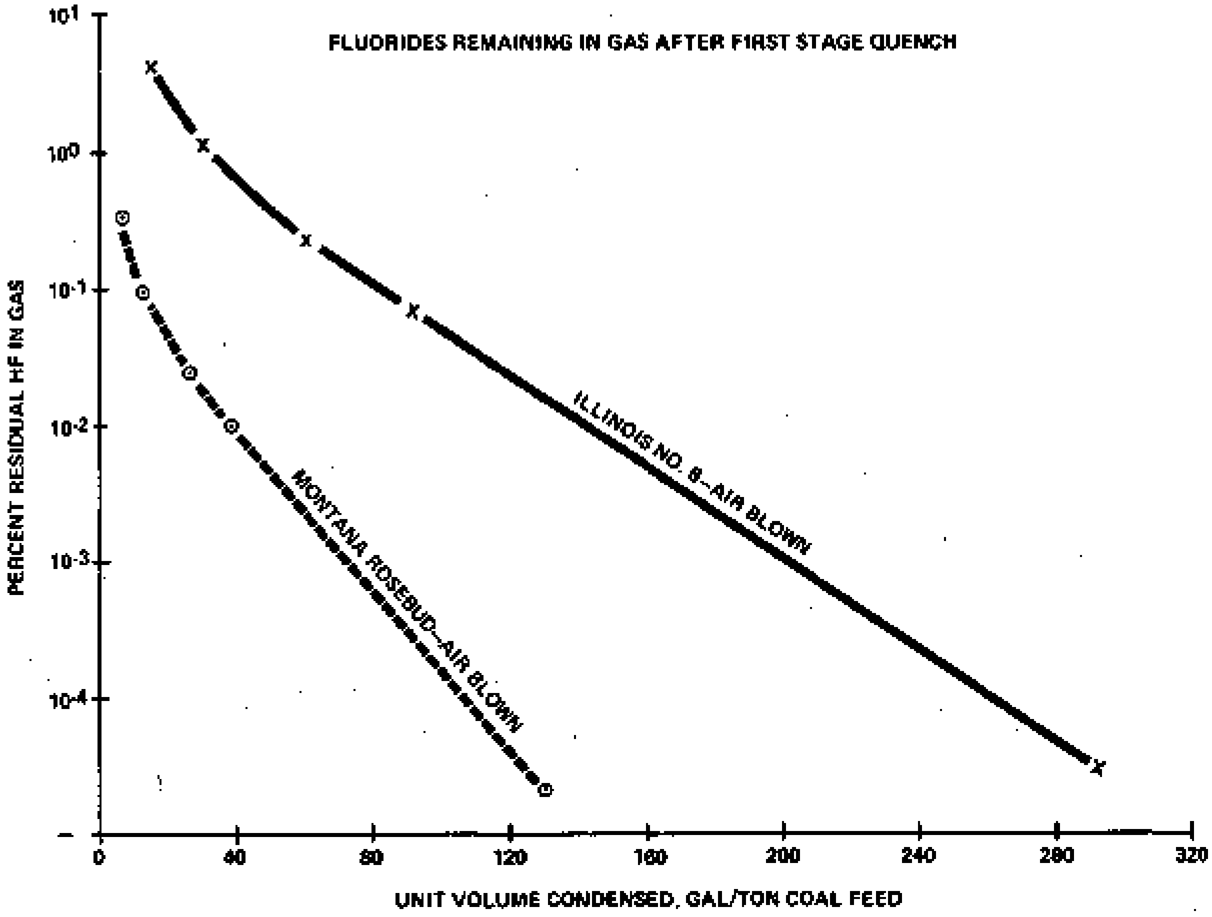

FIGURE 6-2: EFFECT OF UNIT VOLUME CONDENSED ON ABSORPTION OF FLUORIDES

the gas. This amounted to 98 percent removal of fluoride in the first 10 percent of the total condensate. Fluoride removal from gas produced from Montana Rosebud was predicted to be higher because of increased alkalinity in the condensate. Thus the concept of preliminary absorption of halides was theoretically viable.

The removal of HCl in the first stage quench as a function of condensate volume per ton of coal feed is shown in Figure 6-3 for an oxygen blown gasifier. For producer gas from Illinois No. 6 coal, calculations indicate that 99.9997 percent of the chlorides are removed by quenching 5% of the water vapor contained in the absorber inlet flow stream. Expressed as an absolute quantity, the 5 percent quench corresponds to 21.5 gallons of condensate per ton of coal feed. As shown in Figure 6-3, increasing the amount of condensate does result in greater removal of chlorides, but the removal attained for a 5 percent quench is quite adequate.

As indicated for the air blown case, the higher alkalinity of the condensate for the oxygen blown, Montana Rosebud case increases the removal efficiency of HCl.

The prediction of HF removal for the oxygen blown case also yields favorable results as shown in Figure 6-4. A 20 percent quench achieves a 99.2 percent removal of HF for the producer gas from Illinois No. 6 coal. At a 30 percent level of condensation, 99.72 percent of the HF contained in the product gas is removed. A 20 percent quench value corresponds to 85.86 gallons of water condensed per ton of coal feed. The 30 percent quench value corresponds to 129 gallons of water condensed per ton of coal feed.

For the producer gas from the Montana Rosebud Coal, 99.63 percent of the HF is removed by employing a 5 percent quench. For a 10 percent quench, 99.895 percent of the HF is removed from the product gas stream. The 10 percent quench corresponds to 11 gallons of water condensed per ton of coal feed.

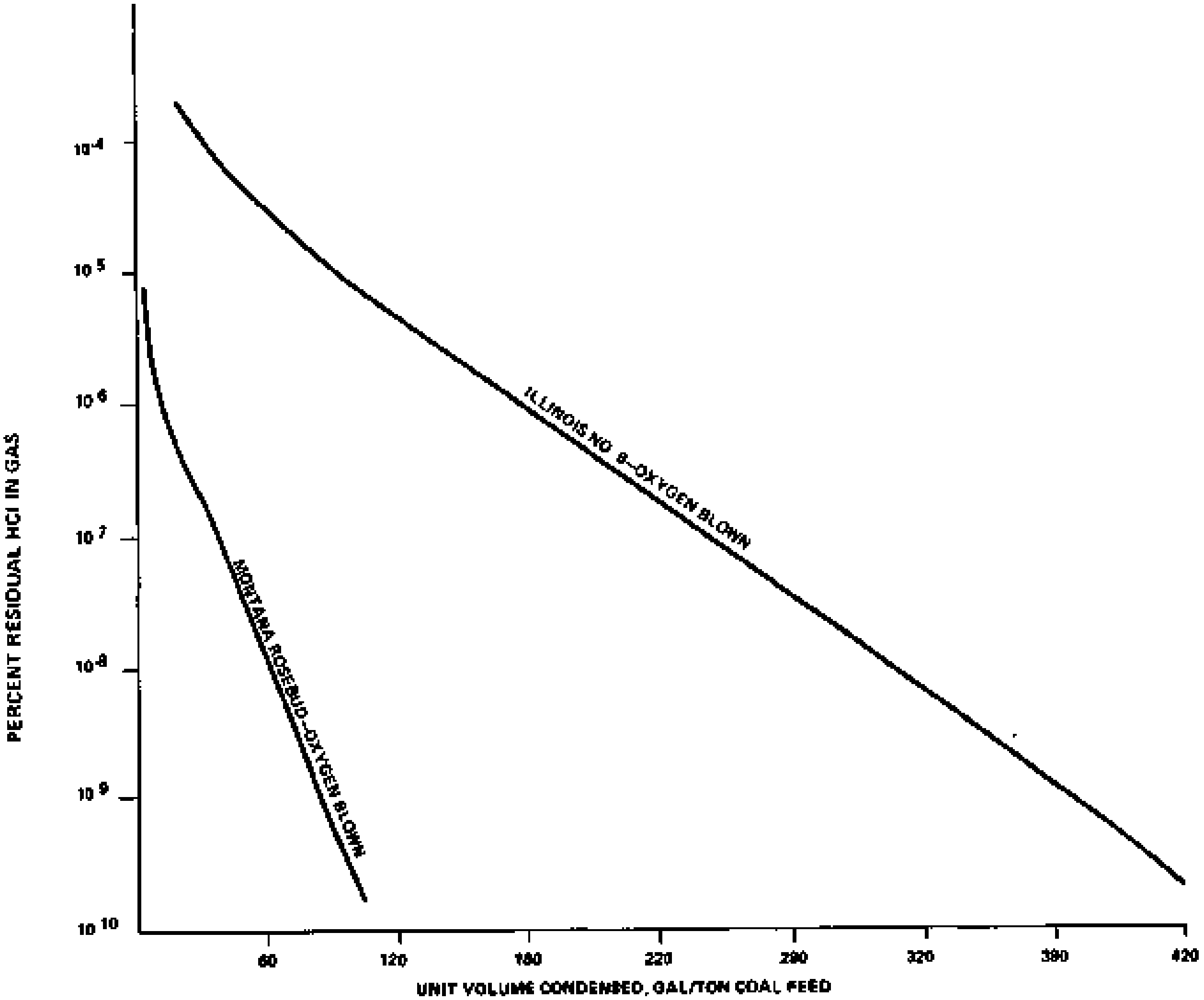

FIGURE 6-3: EFFECT OF UNIT VOLUME WATER CONDENSED ON ABSORPTION OF CHLORIDES–OXYGEN BLOWN CASE

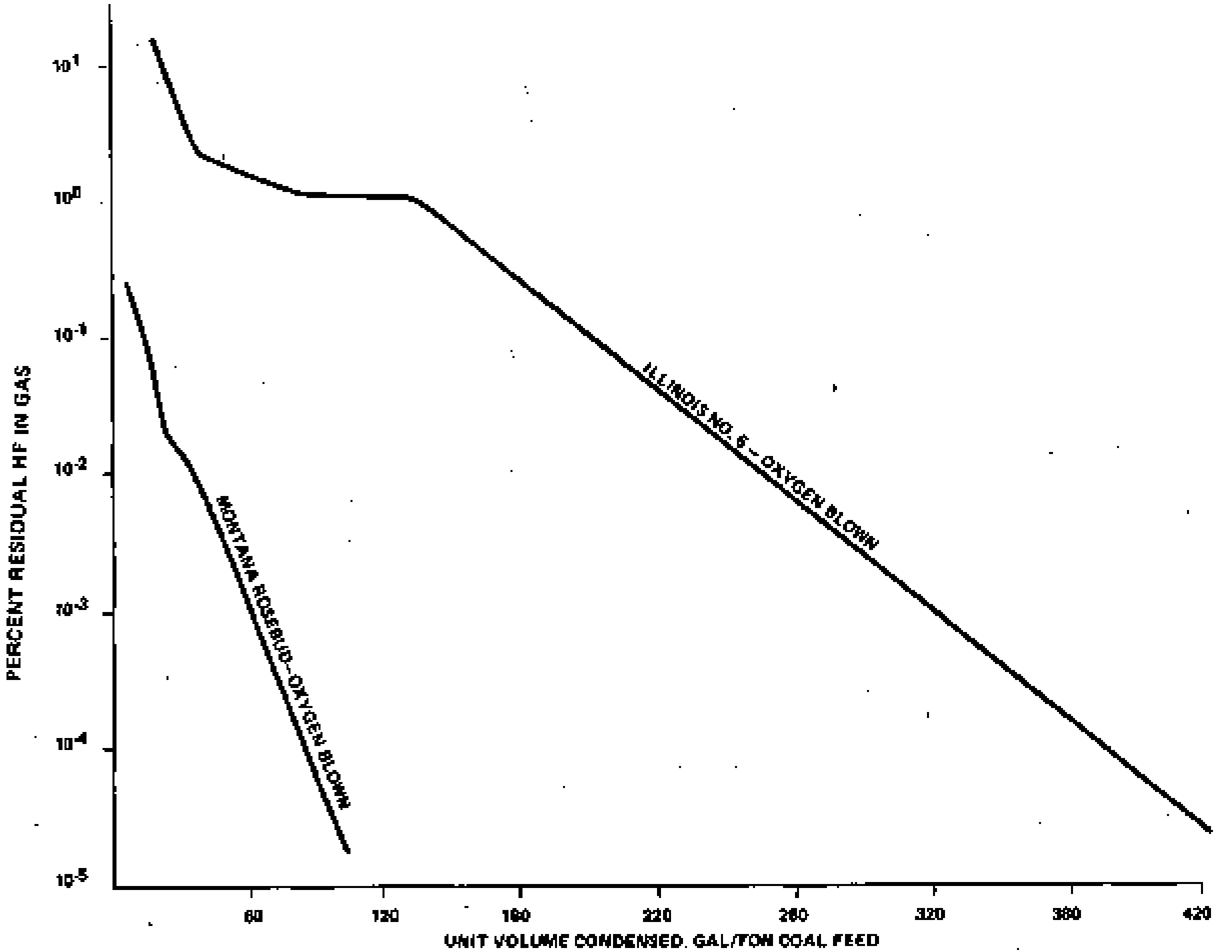

FIGURE 6-4: EFFECT OF UNIT VOLUME WATER CONDENSED ON ABSORPTION OF FLUORIDES–OXYGEN BLOWN CASE

The final design specifications for the absorber units call for nine trays for each of the Montana Rosebud gasification cases. For the producer gas from the Illinois No. 6 coal nine trays are required for ammonia equilibrium for the air blown case and seven trays for oxygen blown case.

## 6.2 Condensate Strippers Performance

The results from the modified WAVES program were compared to data from a study of sour water stripping (24) and reasonable correlation was obtained. Steam stripping was indicated to effectively remove volatile weak electrolytes such as $NH_3$, $CO_2$ and $H_2S$. Free cyanide would also be quantitatively removed but it forms stable complexes of low volatility with metals such as iron and it reacts through various mechanisms with sulfur compounds during cooling of the gas to form thiocyanate - a strong electroylte. Thiocyanate and metal cyanide complexes are refractory to the steam stripping process and therefore go with the stripper bottoms to biological treatment.

The predicted cyanide removal for all strippers was based on a study done by API on sour water strippers (25). Their findings indicate that the average cyanide removal was only 37% by stripping. For this study it was assumed that 40 percent of the total cyanides were removed for all stripper cases. The ratio of cyanide to thiocyanate was based on operational data.

The removal of phenol from the wastewater in the stripper was also based on operational data obtained from literature (24). The data indicated 20 percent removal of phenol when the stripper is operating with total reflux.

The results of the condensate stripper performance are shown in Tables 6-1 and 6-2 for the two air blown cases and in Tables 6-3 and 6-4 for the oxygen blown cases. One can observe the excellent removal of $NH_3$, $CO_2$ and $H_2S$ from the condensate with the specified strippers.

For the low halide stripper design, it was found that all of HCN and $CO_2$ could be removed after 8 theoretical trays. The $NH_3$ and $H_2S$ could be reduced to less than 50 and 5 ppm respectively using 22 theoretical trays.

For the high halide stripper designs, it was found that the presence of the strong acids, HCl and HF, fixed the ammonia in solution and made the weaker acid gases, $CO_2$, $H_2S$ and HCN more volatile. As a result, nearly all of the acid gases were stripped out of solution after about 4 theoretical trays. At this point, the injection of caustic was specified to neutralize the HCl and HF and free the fixed ammonia. For an assumed pH of approximately 8, calculations predict that the free ammonia could be stripped down to 50 ppm after about 20 theoretical trays for the oxygen blown cases. For the nitrogen abundant air blown cases, the $NH_3$ concentration was reduced to 52 mg/l and 51 mg/l for Illinois No. 6 and Montana Rosebud respectively.

The design specifications for the wastewater strippers contained in the single stage quench process train are the same as those for the low halide strippers. The final design calculations specify 40 sieve trays for each of these strippers. For the high halide wastewater strippers, 35 sieve trays were required to achieve the desired removal of volatile weak electrolytes. For actual operating conditions, the stripping steam rates and caustic addition rates, could be altered to achieve the desired bottoms concentrations for the specified number of trays.

## Table 6-1: Effect of Treatment on Condensate Characteristics

### Illinois No. 6 - Air Blown

| | Raw Condensate | | | Condensate After Stripping | | | Biotreated Condensate | | |
|---|---|---|---|---|---|---|---|---|---|
| Component | Single Quench | 10% Quench | Final Quench | Single Quench | 10% Quench | Final Quench | Single Quench | 10% Quench | Final Quench |
| $H_2O$ gal/ton coal | 292 | 30.1 | 262 | 293 | 33 | 261 | 294 | 34 | 259 |
| $NH_3$, mg/l | 5310 | 10880 | 4671 | 50 | 52 | 50 | 44 | 24 | 42 |
| $CO_2$, mg/l | 10740 | 360 | 11790 | 5 | 5 | 5 | - | - | - |
| $H_2S$, mg/l | 350 | 44 | 63 | 5 | 5 | 5 | 0.1 | 0.1 | 0.1 |
| HCl, mg/l | 2110 | 20480 | 1 | - | - | 1 | - | - | - |
| HF, mg/l | 104 | 1000 | 1.3 | - | - | 1.3 | - | - | - |
| CN, mg/l | 50 | 0.61 | 51 | 15 | 0.15 | 15 | 3.0 | 0.2 | 3.0 |
| SCN, mg/l | 260 | 3.2 | 275 | 171 | 1.9 | 177 | 10 | 2 | 10 |
| Phenol, mg/l | 3329 | 2209 | 3640 | 2640 | 1630 | 2920 | 0.1 | 0.1 | 0.1 |
| pH | 7.1 | 5.8 | 7.2 | 8-10 | 8-10 | 7-9 | 7.7 | 7.8 | 7.1 |
| NaCl, mg/l | - | - | - | 3460 | 29100 | - | 3380 | 29200 | 1 |
| NaF, mg/l | - | - | - | 230 | 1860 | - | 219 | 1870 | 2 |
| $SO_4$, mg/l | - | - | - | - | - | - | 282 | 14 | 291 |
| TSS, mg/l | - | - | - | - | - | - | 25 | 25 | 25 |
| Total Equivalent Strong Electrolyte, mg/l as NaCl | | | | | | | 4028 | 31800 | 359 |
| Concentration Ratio | | | | | | | 11.3 | 89 | 1.0 |

Table 6-2: Effect of Treatment on Condensate Characteristics

Illinois No. 6 - Oxygen Blown

| | Raw Condensate | | | Condensate After Stripping | | | Biotreated Condensate | | |
|---|---|---|---|---|---|---|---|---|---|
| Component | Single Quench | 10% Quench | Final Quench | Single Quench | 10% Quench | Final Quench | Single Quench | 10% Quench | Final Quench |
| $H_2O$ gal/ton coal | 422 | 42.8 | 379 | 424 | 45.5 | 378.5 | 424 | 46.0 | 378 |
| $NH_3$, mg/l | 3724 | 7673 | 3280 | 50 | 50 | 50 | 66 | 30 | 66 |
| $CO_2$, mg/l | 8925 | 442 | 9657 | 5 | 5 | 5 | - | - | - |
| $H_2S$, mg/l | 236 | 47 | 254 | 5 | 5 | 5 | 0.1 | 0.1 | 0.1 |
| HCl, mg/l | 1457 | 14400 | 1 | - | - | 1 | - | - | - |
| HF, mg/l | 71.0 | 677 | 2.8 | - | - | 2.8 | - | - | - |
| CN, mg/l | 68.7 | 0.085 | 71.7 | 20.5 | 0.02 | 21.5 | 4.1 | 0.2 | 4.3 |
| SCN, mg/l | 360 | 0.44 | 376 | 236 | 0.27 | 248 | 10 | 2 | 10 |
| Phenol, mg/l | 2402 | 1481 | 2605 | 1919 | 1095 | 2093 | 0.1 | 0.1 | 0.1 |
| pH | 6.7 | 5.5 | 6.7 | 8-10 | - | 7-9 | 7.7 | 7.8 | 7.1 |
| NaCl, mg/l | - | - | - | 2329 | 21680 | - | 2330 | 21550 | 1 |
| NaF, mg/l | - | - | - | 149 | 1335 | - | 149 | 1327 | 3 |
| $SO_4$, mg/l | - | - | - | - | - | - | 390 | 27 | 408 |
| TSS, mg/l | - | - | - | - | - | - | - | - | - |
| Total Equivalent Strong Electrolyte, mg/l as NaCl | | | | | | | 3018 | 23430 | 503 |
| Concentration Ratio | | | | | | | - | - | - |

## Table 6-3: Effect of Treatment on Condensate Characteristics

### Montana Rosebud - Air Blown

| Component | Raw Condensate | | | Condensate After Stripping | | | Biotreated Condensate | | |
|---|---|---|---|---|---|---|---|---|---|
| | Single Quench | 10% Quench | Final Quench | Single Quench | 10% Quench | Final Quench | Single Quench | 10% Quench | Final Quench |
| $H_2O$ gal/ton | 124 | 13.0 | 111 | 123 | 13.4 | 109 | 123 | 13.8 | 109 |
| $NH_3$, mg/l | 11990 | 4866 | 12730 | 50 | 51 | 50 | 40 | 4 | 31 |
| $CO_2$, mg/l | 27110 | 1888 | 29270 | 5 | 5 | 5 | - | - | - |
| $H_2S$, mg/l | 199 | 75 | 225 | 5 | 5 | 5 | 0.1 | 0.1 | 0.1 |
| HCl, mg/l | 601 | 5718 | 1 | - | - | 1 | - | - | - |
| HF, mg/l | 60 | 574 | 1 | - | - | 0.07 | - | - | - |
| CN, mg/l | 78 | 3.2 | 80 | 24 | 0.93 | 24 | 4.8 | 0.7 | 4.8 |
| SCN, mg/l | 410 | 16 | - | 270 | 11 | 279 | 10 | 2 | 10 |
| Phenol, mg/l | 6299 | 4675 | 7239 | 5056 | 3637 | 5824 | 0.1 | 0.1 | 0.1 |
| pH | 7.5 | 6.8 | 7.6 | 8-10 | 8-10 | 7-9 | 7.5 | 7.6 | 7.1 |
| NaCl, mg/l | - | - | - | 966 | 8619 | - | 976 | 8702 | 1 |
| NaF, mg/l | - | - | - | 127 | 1134 | - | 128 | 1145 | 1 |
| $SO_4$, mg/l | - | - | - | - | - | - | 447 | 43 | 458 |
| TSS, mg/l | - | - | - | - | - | - | 25 | 25 | 25 |
| Total Equivalent Strong Electrolyte, mg/l as NaCl | | | | | | | 1649 | 10350 | 561 |
| Concentration Ratio | | | | | | | 2.94 | 18.4 | 1.0 |

## Table 6-4: Effect of Treatment on Condensate Charactertistics

### Montana Rosebud - Oxygen Blown

| | Raw Condensate | | | Condensate After Stripping | | | Biotreated Condensate | | |
|---|---|---|---|---|---|---|---|---|---|
| Component | Single Quench | 10% Quench | Final Quench | Single Quench | 10% Quench | Final Quench | Single Quench | 10% Quench | Final Quench |
| $H_2O$ gal/ton coal | 104 | 10.9 | 93.5 | 103 | 11.3 | 91.9 | 103 | 11.6 | 91.3 |
| $NH_3$, mg/l | 14590 | 5965 | 15500 | 50 | 50 | 50 | 44 | * | 34 |
| $CO_2$, mg/l | 33580 | 2630 | 36230 | 5 | 5 | 5 | - | - | - |
| $H_2S$, mg/l | 174 | 93.1 | 201 | 5 | 5 | 5 | 0.1 | 0.1 | 0.1 |
| HCl, mg/l | 715 | 6833 | 1 | - | - | 1 | - | - | - |
| HF, mg/l | 71.3 | 681 | 0.08 | - | - | 0.09 | - | - | - |
| CN, mg/l | 99.6 | 3.4 | 101 | 30.3 | 0.98 | 30.9 | 6.1 | 0.7 | 6.2 |
| SCN, mg/l | 523 | 17.7 | 532 | 348 | 11.3 | 356 | 10 | 2.2 | 10 |
| Phenol, mg/l | 7315 | 5501 | 8606 | 5917 | 4253 | 6997 | 0.1 | 0.1 | 0.1 |
| pH | 7.5 | 6.7 | - | 8-10 | 8-10 | 7-9 | 7.5 | 7.6 | 7.1 |
| NaCl, mg/l | - | - | - | 1160 | 10610 | - | 1165 | 10370 | 1 |
| NaF, mg/l | - | - | - | 152 | 1385 | - | 152 | 1356 | 1 |
| $SO_4$, mg/l | - | - | - | - | - | - | 569 | 47 | 584 |
| TSS, mg/l | - | - | - | - | - | - | - | - | - |
| Total Equivalent Strong Electrolyte, mg/l as NaCl | | | | | | | 2070 | 12320 | 715 |
| Concentration Ratio | | | | | | | - | - | - |

*Ammonia supplement required for bio-oxidation

### 6.3 Biological Treatment

Biological treatment of condensates is projected to achieve excellent conversion of phenol plus substantial conversion of other organics and thiocyanates. This can be seen by again referring to Tables 6-1 through 6-4. Some additional electrolyte is introduced into the second stage condensate stream by means of the degradation of thiocyanate to ammonia and sulfate. This increases the total electrolyte concentration of the biologically treated reuseable condensate.

Comparisons between the total electrolyte concentration of the single stage condensate to that of the two stage reuseable condensate indicate excellent results for the two-stage scheme. Table 6-1 shows the predicted performance of biological treatment on the condensate obtained from the air blown gasification of Illinois No. 6. Each strong electrolyte was expressed as an equivalent amount of NaCl for purposes of quantifying and comparing the condensate composition obtained from the proposed processing schemes. The total equivalent strong electrolyte concentration for the second stage quench was predicted to be just 9 percent of that contained in the single stage quench and 1.1 percent of the first stage quench of 34 gallons per ton of coal feed.

For the oxygen blown Illinois No. 6 coal case, the biotreated condensate strong electrolyte concentration for the first stage quench was predicted to be just 16.7 percent of that contained in the single stage quench. The strong electrolyte concentration of the second stage quench was predicted to be just 2.1 percent of the first stage condensate at a quench of 43 gallons of water per ton of coal feed.

For the gasification of Montana Rosebud coal cases, calculations also indicate that the two stage quench yields a condensate with a much more favorable composition than that of the single stage case.

For the air blown case, the total equivalent strong electrolyte concentration of the second stage quench was projected to be 34 percent of that contained in the single stage biotreated condensate. The strong electrolyte concentration of the second stage condensate was projected to be 5.4 percent the first stage quench condensate at a quench level of 13.8 gallons per ton of coal feed.

For the case of oxygen blown gasification of Montana Rosebud coal, the projected total equivalent strong electrolyte concentration of the second stage condensate is 34.5 percent of the single stage condensate electrolyte concentration. As indicated before, the superior performance of the selective absorption unit can be realized by noting that the total strong electrolyte concentration of the second stage quench is projected to be just 5.8 percent of that calculated for the first stage quench. This is stated for the case of a 10 percent quench in the absorber unit which is equivalent to 11 gallons of water condensed per ton of coal feed.

In review of the projected results, one can observe the lower relative performance of the two stage quench for the Montana Rosebud cases. This can be attributed to the lower halide content of this coal. One should also note the substantial removal of phenol, SCN, CN and $H_2S$ achieved with the biological treatment step. In particular, the high phenol content of the stripper condensate is reduced to a trace content in the biological treatment effluent.

## 6.4 Analysis of Cooling Circuits

A flow schematic for the recycle cooling circuits associated with the gasification systems is given as Figure 6-5. The recycle cooling system concept was used in conjunction with water quality criteria for recycle circuits to obtain a quantitative comparison between biologically treated secondary condensates and single stage condensates treated by activated sludge and reverse osmosis. The

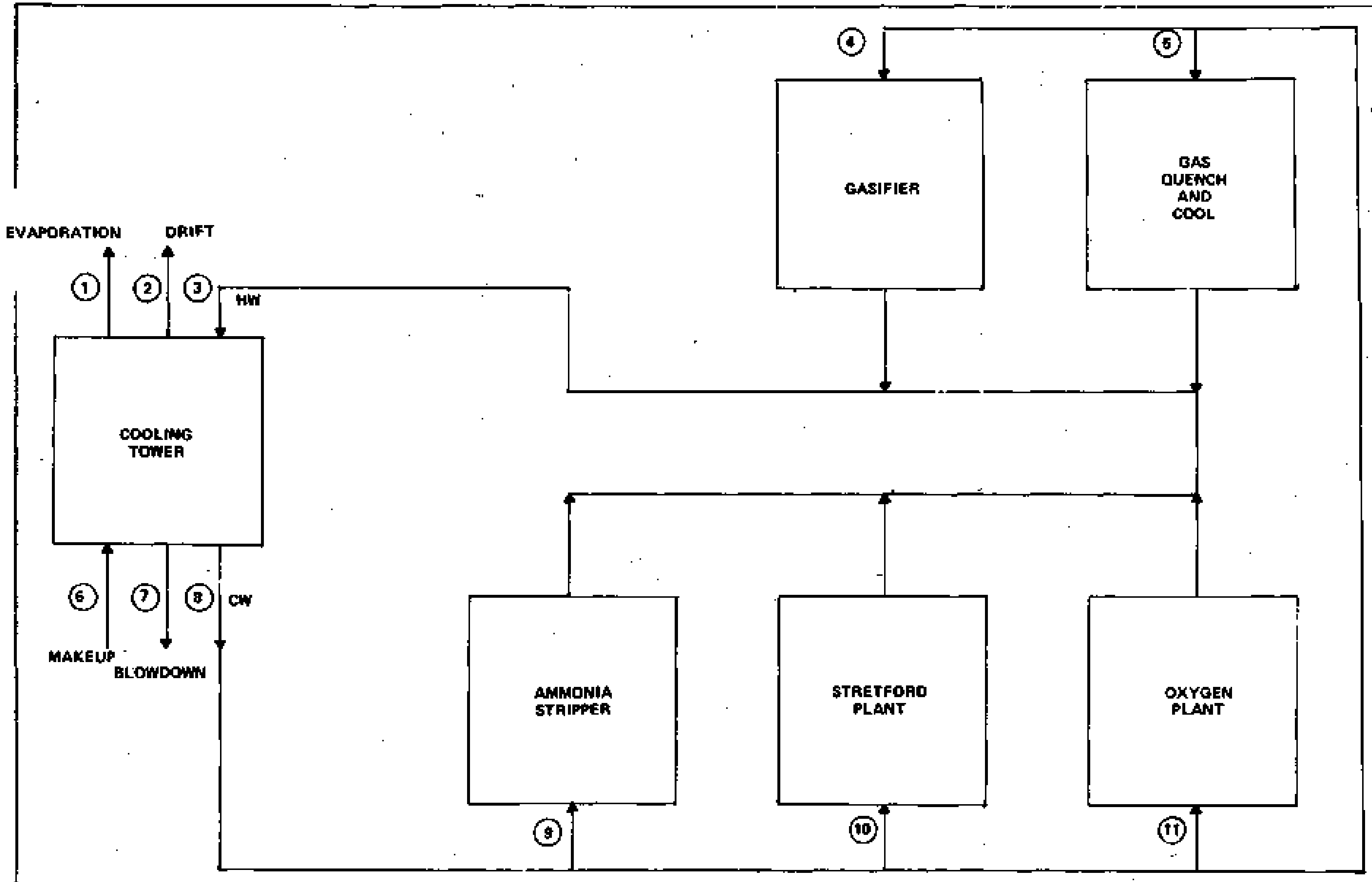

FIGURE 6-5: WATER BALANCE–COOLING CIRCUITS

condensate from the single quench was treated by activated sludge and reverse osmosis to yield a concentrate equal in volume in all cases to the alternate 10 percent first stage quench.

Tables 6-5 and 6-6 provide a comparison of the performance of treated condensates from single quench and two stage quench systems operating air blown on Illinois No. 6 coal. A comparison of the results revealed that the treated effluent from the single quench condensate required 696 gpm of blowdown whereas the blowdown from the treated secondary condensate was 105 gpm. The superior performance with the secondary condensate was attributable to the low halide content of secondary condensate as compared to the permeate from reverse osmosis.

Tables 6-7 and 6-8 provided a comparison of the performance of treated condensates from single quench and two stage quench systems operating oxygen blown on Illinois No. 6 coal. The results of the analysis of the cooling tower circuits showed that a blowdown of 394 gpm was required using treated single stage condensate as makeup whereas a blowdown of 214 gpm was adequate for operation on makeup consisting of treated secondary condensate. Halides were the water quality limitation for operation on reverse osmosis permeate as makeup; whereas total strong electrolyte was the water quality limitation using treated secondary condensate as makeup. The weaker performance indicated for the oxygen blown system operating on treated secondary condensate was attributable to the projected presence of increased thiocyanate in the secondary condensate as compared to airblown gasification.

Tables 6-9 and 6-10 present the results of analysis of cooling tower circuits operating on treated condensates from airblown gasification of Montana Rosebud. The blowdown from operation on treated condensate from single quench was indicated as 47 gpm as compared to 95 gpm for operation on treated secondary condensate.

### Table 6-5: Analysis of Cooling Tower Circuits for Air Blown Single Quench Using Illinois No. 6 Coal

| Stream Number and Description | Flow gpm | Temp F° | Total Halides as Cl mg/l | Total Strong Electrolyte as NaCl mg/l |
|---|---|---|---|---|
| 1. Cooling Tower Evaporation | 405 | - | - | - |
| 2. Cooling Tower Drift | 3.0 | - | 600 | 859 |
| 3. Cooling Tower Return Flow | 32650 | 100.5 | 600 | 859 |
| 4. Gasifier Cooling Water Inputs | 150 | 85 | 600 | 859 |
| 5. Cooling Water Input to Gas Cooling | 28200 | 85 | 600 | 859 |
| 6. Cooling Tower Makeup | 1104 | 75 | 380 | 544 |
| 7. Cooling Tower Blowdown | 696 | 85 | 600 | 859 |
| 8. Cooling Tower Recycle Flow | 32650 | 85 | 600 | 859 |
| 9. Cooling Water Input to Ammonia Stripper | 1500 | 85 | 600 | 859 |
| 10. Cooling Water Input to Stretford Plant | 2800 | 85 | 600 | 859 |
| 11. Cooling Water Input to Oxygen Plant | - | - | - | - |

NOTE: Stream numbers referenced to Figure 6-5.

## Table 6-6: Analysis of Cooling Tower Circuits for Air Blown Two Stage Quench* Using Ilinois No. 6 Coal

| Stream Number and Description | Flow gpm | Temp F° | Total Halides as Cl mg/l | Total Strong Electrolyte as NaCl mg/l |
|---|---|---|---|---|
| 1. Cooling Tower Evaporation | 405 | - | - | - |
| 2. Cooling Tower Drift | 3.0 | - | 11 | 1700 |
| 3. Cooling Tower Return Flow | 32650 | 100.5 | 11 | 1700 |
| 4. Gasifier Cooling Water Inputs | 150 | 85 | 11 | 1700 |
| 5. Cooling Water Input to Gas Cooling | 28200 | 85 | 11 | 1700 |
| 6. Cooling Tower Makeup | 513 | 75 | 2.3 | 357 |
| 7. Cooling Tower Blowdown | 105 | 85 | 11 | 1700 |
| 8. Cooling Tower Recycle Flow | 32650 | 85 | 11 | 1700 |
| 9. Cooling Water Input to Ammonia Stripper | 1500 | 85 | 11 | 1700 |
| 10. Cooling Water Input to Stretford Plant | 2800 | 85 | 11 | 1700 |
| 11. Cooling Water Input to Oxygen Plant | - | - | - | - |

*Ten percent of gas moisture condensed in primary quench

NOTE: Stream numbers referenced to Figure 6-5.

## Table 6-7: Analysis of Cooling Tower Circuits for Oxygen Blown Single Quench Using Illinois No. 6 Coal

| Stream Number and Description | Flow gpm | Temp F° | Total Halides as Cl mg/l | Total Strong Electrolyte as NaCl mg/l |
|---|---|---|---|---|
| 1. Cooling Tower Evaporation | 518 | - | - | - |
| 2. Cooling Tower Drift | 4 | - | - | - |
| 3. Cooling Tower Return Flow | 41500 | 100.6 | 600 | 944 |
| 4. Gasifier Cooling Water Inputs | 150 | 85 | 600 | 944 |
| 5. Cooling Water Input to Gas Cooling | 35705 | 85 | 600 | 944 |
| 6. Cooling Tower Makeup | 916 | 75 | 261 | 411 |
| 7. Cooling Tower Blowdown | 394 | 85 | 600 | 944 |
| 8. Cooling Tower Recycle Flow | 41500 | 85 | 600 | 944 |
| 9. Cooling Water Input to Ammonia Stripper | 2190 | 85 | 600 | 944 |
| 10. Cooling Water Input to Stretford Plant | 2771 | 85 | 600 | 944 |
| 11. Cooling Water Input to Oxygen Plant | 684 | 85 | 600 | 944 |

NOTE: Stream numbers referenced to Figure 6-5.

## Table 6-8: Analysis of Cooling Tower Circuits for Oxygen Blown Two Stage Quench*
## Using Illinois No. 6 Coal

| Stream Number and Description | Flow gpm | Temp F° | Total Halides as Cl mg/l | Total Strong Electrolyte as NaCl mg/l |
|---|---|---|---|---|
| 1. Cooling Tower Evaporation | 518 | - | - | - |
| 2. Cooling Tower Drift | 4 | - | 11 | 1700 |
| 3. Cooling Tower Return Flow | 41500 | 100.6 | 11 | 1700 |
| 4. Gasifier Cooling Water Inputs | 150 | 85 | 11 | 1700 |
| 5. Cooling Water Input to Gas Cooling | 35705 | 85 | 11 | 1700 |
| 6. Cooling Tower Makeup | 736 | 75 | 3 | 503 |
| 7. Cooling Tower Blowdown | 214 | 85 | 11 | 1700 |
| 8. Cooling Tower Recycle Flow | 41500 | 85 | 11 | 1700 |
| 9. Cooling Water Input to Ammonia Stripper | 2190 | 85 | 11 | 1700 |
| 10. Cooling Water Input to Stretford Plant | 2771 | 85 | 11 | 1700 |
| 11. Cooling Water Input to Oxygen Plant | 684 | 85 | 11 | 1700 |

*Ten percent of gas moisture condensed in primary quench

NOTE: Stream numbers referenced to Figure 6-5.

## Table 6-9: Analysis of Cooling Tower Circuits for Air Blown Single Quench Using Montana Rosebud Coal

| Stream Number and Description | | Flow gpm | Temp F° | Total Halides as Cl mg/l | Total Strong Electrolyte as NaCl mg/l |
|---|---|---|---|---|---|
| 1. | Cooling Tower Evaporation | 197 | - | - | - |
| 2. | Cooling Tower Drift | 2 | 85 | 600 | 958 |
| 3. | Cooling Tower Return Flow | 15977 | 100.4 | 600 | 958 |
| 4. | Gasifier Cooling Water Inputs | 150 | 85 | 600 | 958 |
| 5. | Cooling Water Input to Gas Cooling | 14388 | 85 | 600 | 958 |
| 6. | Cooling Tower Makeup | 246 | 75 | 119 | 190 |
| 7. | Cooling Tower Blowdown | 47 | 85 | 600 | 958 |
| 8. | Cooling Tower Recycle Flow | 15977 | 85 | 600 | 958 |
| 9. | Cooling Water Input to Ammonia Stripper | 616 | 85 | 600 | 958 |
| 10. | Cooling Water Input to Stretford Plant | 823 | 85 | 600 | 958 |
| 11. | Cooling Water Input to Oxygen Plant | - | - | - | - |

NOTE: Stream numbers referenced to Figure 6-5.

## Table 6-10: Analysis of Cooling Tower Circuits for Air Blown Two Stage Quench* Using Montana Rosebud Coal

| Stream Number and Description | Flow gpm | Temp F° | Total Halides as Cl mg/l | Total Strong Electrolyte as NaCl mg/l |
|---|---|---|---|---|
| 1. Cooling Tower Evaporation | 197 | - | - | - |
| 2. Cooling Tower Drift | 2 | - | 5 | 1700 |
| 3. Cooling Tower Return Flow | 15977 | 100.4 | 5 | 1700 |
| 4. Gasifier Cooling Water Inputs | 150 | 85 | 5 | 1700 |
| 5. Cooling Water Input to Gas Cooling | 14388 | 85 | 5 | 1700 |
| 6. Cooling Tower Makeup | 294 | 75 | 2 | 561 |
| 7. Cooling Tower Blowdown | 95 | 85 | 5 | 1700 |
| 8. Cooling Tower Recycle Flow | 15977 | 85 | 5 | 1700 |
| 9. Cooling Water Input to Ammonia Stripper | 616 | 85 | 5 | 1700 |
| 10. Cooling Water Input to Stretford Plant | 823 | 85 | 5 | 1700 |
| 11. Cooling Water Input to Oxygen Plant | - | - | - | - |

*Ten percent of gas moisture condensed in primary quench

NOTE: Stream numbers referenced to Figure 6-5.

The superior performance of the permeate as cooling system makeup was attributable to the high selectivity of reverse osmosis to the substantial mole percentage of sulfate in the Montana Rosebud gas condensate.

The results of analysis of cooling tower circuits for oxygen blown operation on Montana Rosebud coal are given in Tables 6-11 and 6-12. The blowdown flow indicated for cooling tower operation on treated condensate from single quench was 53 gpm as compared to 127 gpm for operation on treated secondary condensate. Thus the reverse osmosis process included in the treatment of the single quench condensate was designated as more effective in the particular case than the application of a two stage quench system.

## 6.5 Economic Analysis

Similarities and differences exist between gas trains incorporating single stage and two stage quench systems. It follows that an economic comparison between the systems can be constructed from consideration only of the costs attributable to components that are dissimilar. Tables 6-13 through 6-16 present estimates of capital and operating costs based on dissimilar components between "equalized" single stage and two stage quench systems. The analyses were based on carbon steel absorbers and decanters, and assumed similar loss of gas pressure through the alternate systems.

The results in Tables 6-13 and 6-14 pertaining to gasifier operation, air blown and oxygen blown, on Illinois No. 6 coal indicated a cost advantage for the two stage quench alternative. The implication was that the cost of the reverse osmosis facility attached to the single quench system exceeded the cost increase associated with the more complex two stage quench system. The results in Tables 6-15 and 6-16 pertaining to gasifier operation, air blown and oxygen blown, on Montana Rosebud coal suggested a cost advantage for the

### Table 6-11: Analysis of Cooling Tower Circuits for Oxygen Blown Single Quench Using Montana Rosebud Coal

| Stream Number and Description | Flow gpm | Temp F° | Total Halides as Cl mg/l | Total Strong Electrolyte as NaCl mg/l |
|---|---|---|---|---|
| 1. Cooling Tower Evaporation | 176 | - | - | - |
| 2. Cooling Tower Drift | 1 | - | - | - |
| 3. Cooling Tower Return Flow | 14282 | 100.4 | 600 | 960 |
| 4. Gasifier Cooling Water Inputs | 150 | 85 | 600 | 960 |
| 5. Cooling Water Input to Gas Cooling | 12330 | 85 | 600 | 960 |
| 6. Cooling Tower Makeup | 230 | 75 | 142 | 227 |
| 7. Cooling Tower Blowdown | 53 | 85 | 600 | 960 |
| 8. Cooling Tower Recycle Flow | 14282 | 85 | 600 | 960 |
| 9. Cooling Water Input to Ammonia Stripper | 512 | 85 | 600 | 960 |
| 10. Cooling Water Input to Stretford Plant | 818 | 85 | 600 | 960 |
| 11. Cooling Water Input to Oxygen Plant | 472 | 85 | 600 | 960 |

NOTE: Stream numbers referenced to Figure 6-5.

## Table 6-12: Analysis of Cooling Tower Circuits for Oxygen Blown Two Stage Quench* Using Montana Rosebud Coal

| Stream Number and Description | Flow gpm | Temp F° | Total Halides as Cl mg/l | Total Strong Electrolyte as NaCl mg/l |
|---|---|---|---|---|
| 1. Cooling Tower Evaporation | 176 | - | - | - |
| 2. Cooling Tower Drift | 1 | - | - | - |
| 3. Cooling Tower Return Flow | 14282 | 100.4 | 2 | 1700 |
| 4. Gasifier Cooling Water Inputs | 150 | 85 | 2 | 1700 |
| 5. Cooling Water Input to Gas Cooling | 12330 | 85 | 2 | 1700 |
| 6. Cooling Tower Makeup | 304 | 75 | 1 | 715 |
| 7. Cooling Tower Blowdown | 127 | 85 | 2 | 1700 |
| 8. Cooling Tower Recycle Flow | 14282 | 85 | 2 | 1700 |
| 9. Cooling Water Input to Ammonia Stripper | 512 | 85 | 2 | 1700 |
| 10. Cooling Water Input to Stretford Plant | 818 | 85 | 2 | 1700 |
| 11. Cooling Water Input to Oxygen Plant | 472 | 85 | 2 | 1700 |

*Ten percent of gas moisture condensed in primary quench

NOTE: Stream numbers referenced to Figure 6-5.

### Table 6-13: Comparative Cost Analysis for Single and Two Stage Quench Systems
### Illinois No. 6 - Air Blown

| | Single Quench | Two-Stage Quench |
|---|---|---|
| A. Capital Costs MM$ | | |
| Gas Cleaning and Cooling Module | 8.40 | 9.50 |
| Condensate Stripping and Biotreatment | 7.20 | 8.30 |
| Reverse Osmosis | 3.50 | -0- |
| Total Capital Cost, MM$ | 19.10 | 17.80 |
| B. Annualized Costs, MM$/yr. | | |
| Operation and Maintenance Difference | 0.32 | |

### Table 6-14: Comparative Cost Analysis for Single and Two Stage Quench Systems
### Illinois No. 6 - Oxygen Blown

| | Single Quench | Two-Stage Quench |
|---|---|---|
| A. Capital Costs MM$ | | |
| Gas Cleaning and Cooling Module | 9.20 | 10.20 |
| Condensate Stripping and Biotreatment | 7.90 | 9.12 |
| Reverse Osmosis | 3.60 | -0- |
| Total Capital Cost, MM$ | 20.70 | 19.32 |
| B. Annualized Costs, MM$/yr. | | |
| Operation and Maintenance Difference | 0.34 | |

Table 6-15: Comparative Cost Analysis for Single and Two Stage Quench Systems

Montana Rosebud - Air Blown

| | Single Quench | Two-Stage Quench |
|---|---|---|
| A. Capital Costs MM$ | | |
| Gas Cleaning and Cooling Module | 4.00 | 5.50 |
| Condensate Stripping and Biotreatment | 5.50 | 6.58 |
| Reverse Osmosis | 1.86 | 1.54 |
| Total Capital Cost, MM$ | 11.36 | 13.62 |
| B. Annualized Costs, MM$/yr. | | |
| Operation and Maintenance | | 0.14 |

Table 6-16: Comparative Cost Analysis for Single and Two Stage Quench Systems

Montana Rosebud - Oxygen Blown

| | Single Quench | Two-Stage Quench |
|---|---|---|
| A. Capital Cost MM$ | | |
| Gas Cleaning and Cooling Module | 3.60 | 4.60 |
| Condensate Stripping and Biotreatment | 5.35 | 7.43 |
| Reverse Osmosis | 1.70 | 1.17 |
| Total Capital Cost, MM$ | 10.65 | 13.20 |
| B. Annualized Costs, MM$/yr. | | |
| Operation and Maintenance | | 0.15 |

single quench system. The results were attributed to the low halide content of the condensate from Montana Rosebud and the superior rejection capability of reverse osmosis for sulfate as compared to chloride.

To facilitate quantification of the cost advantages, the results of Tables 6-13 to 6-16 were annualized in Table 6-17 by an incremental difference analysis outlined in Section 11, Appendix H. The unit cost advantage for the two stage quench alternative for gasifier operation on Illinois No. 6 coal, air blown and oxygen blown, was $0.99 and $1.07 per ton of coal gasified respectively. The unit cost advantage for the single quench alternative for gasifier operation, air blown and oxygen blown, on Montana Rosebud coal was $0.95 and $1.26 per ton of coal gasified respectively.

## 6.6 Disposal

The operation of coal gasification facilities results in the generation of residues consisting of ash and wastewater blowdowns. It follows that concepts for disposal of residues must be integrated into plans for installation of coal gasification facilities.

For purposes of the study it was presumed that landfilling would be employed for disposal of ash and minor amounts of other solid wastes, such as spent filter cartridges from reverse osmosis pretreatment. The solid wastes would be provided with chemical fixation, sealed encapsulation, or containment and treatment of leachates as required for the protection of surface or groundwater resources from pollution. The concept is considered as applicable to sites at Wheeling, West Virginia or Carson City, Nevada. In some cases solid waste residues can be used for backfill of mined out areas with implementation of necessary precautions against pollution of surface or groundwater resources.

Table 6-17: Incremental Difference Analysis of Annualized Cost of Single Stage and Two Stage Gas Cleaning Systems

| Most Economical Process | Illinois No. 6 Air Blown | Illinois No. 6 $O_2$ Blown | Montana Rosebud Air Blown | Montana Rosebud $O_2$ Blown |
|---|---|---|---|---|
| | Two Stage | Two Stage | Single Stage | Single Stage |
| Incremental Capital Cost Advantage | $1.3 MM | $1.38 MM | $2.26 MM | $2.55 MM |
| Incremental Operating Cost Advantage | $ .32 MM | $ .34 MM | $ .14 MM | $ .15 MM |
| Annualized Incremental Capital Cost Advantage | $ .20 MM | $ .22 MM | $ .36 MM | $ .41 MM |
| Total Incremental Annualized Cost Advantage over Alternative System | $ .52 MM | $ .56 MM | $ .50 MM | $ .66 MM |

| Unit Cost Advantage | Illinois No. 6 Air Blown | Illinois No. 6 $O_2$ Blown | Montana Rosebud Air Blown | Montana Rosebud $O_2$ Blown |
|---|---|---|---|---|
| Most Economical Process | Two Stage | Two Stage | Single Stage | Single Stage |
| $ per ton Coal Gasified | 0.99 | 1.07 | 0.95 | 1.26 |

The wastewater blowdowns could consist of reverse osmosis concentrates, boiler blowdown, spent ion exchange regenerant or cooling tower blowdown. These flows are characterized by substantial content of inorganic salts and low contents of organic matter and suspended solids. As such they should be considered as "inorganic residues" from wastewater treatment rather than as untreated wastewater.

Disposal concepts for wastewater blowdowns that were investigated in the study included store and discharge to river, store and barge to sea, pond evaporation and incineration to dry product. Since the cost of disposal is largely dependent upon the magnitude of the blowdown flow, it follows that an opportunity exists for a trade-off analysis between the cost of systems for reducing blowdown flow and the cost of disposal. Reverse osmosis and thermal evaporation are alternative desalination processes for the conversion of wastewater blowdowns into fractions of reuseable water and blowdown discharges of reduced volume and increased concentration. As a general guideline reverse osmosis is more practical for the concentration of blowdowns with a low solute content and thermal evaporation is more practical for concentration of blowdowns with a high solute content. With evaporation cost at $3.50 per 1000 lb steam and 1.2 gallons condensate per lb steam, the cost of steam exceeds the total cost of operation and maintenance for single stage desalination by reverse osmosis. For purposes of analysis, the following guideline limits were imposed upon reverse osmosis desalination operations:

a. maximum sulfate in concentrate = 6000 mg/l
b. maximum strong electrolyte in concentrate = 9000 mg/l as NaCl
c. maximum chlorides in permeate = 150 mg/l

The intent of the guideline limits was to provide control of membrane fouling by precipitates and to provide permeate that could be used as makeup for evaporative cooling systems operating at four concentrations. Operation at a strong electrolyte concentration of 9000 mg/l was visualized as roughly equivalent to a total dissolved solids content of 1.0 percent. Thus resort to evaporation was necessary for desalination of reverse osmosis concentrates or blowdowns with concentrations above the guideline limits. Reverse osmosis permeate was assigned a value of $0.10 per 1000 gallons at Wheeling and $1.50 per 1000 gallons at Carson City.

An analysis of the characteristics of the biotreated condensates presented in Tables 6-1 to 6-4 indicated that single stage reverse osmosis would produce permeates of acceptable quality for cooling system makeup except in the cases of the condensates from the 10 percent first stage quenches and the single quenches from operation on Illinois No. 6 Coal (Tables 6-1 and 6-2). Two stage reverse osmosis was applicable to the single quench condensates from Illinois No. 6 - condensates from the 10 percent first stage quenches were outside the guideline limits for reverse osmosis.

Multiple effect evaporation has potential application for concentration of condensates from first stage quenching and for increasing the concentration of reverse osmosis concentrates. The concentrations attainable in the bottoms from evaporators is limited by the precipitation of dissolved salts. The precipitation problem is controlled within limits by acidification and by crystal seeding. For purposes of the study limiting strong electrolyte concentrations in multiple effect evaporator bottoms were taken as 9.0 percent. The condensates from evaporation characteristically have dissolved solids contents of 2 to 3 mg/l. The superior quality of the condensates renders it suitable for reuse in boiler feedwater systems. The value assigned to evaporator condensate was $0.80 per 1000 gallons at Wheeling and $2.20 per 1000 gallons at Carson City.

Evaporation with incineration has potential application for conversion of multiple effect evaporator bottoms to a dry product. The system uses a pan evaporator to attain a concentrate of 60 percent solids content for feed to incineration. The system involves a relatively high commitment of resources in terms of energy and cost. Disposal of solid residue in the sea or in abandoned salt mines has been suggested.

### 6.6.1 Disposal at Wheeling Site

The disposal of wastewater blowdowns at the Wheeling, West Virginia site would be accomplished by discharge to surface waters or by deep well disposal. Climatic conditions at the Wheeling site are unfavorable for pond evaporation. The geology of the Wheeling area has been rated as generally favorable for deep well disposal (26), but it is anticipated that a deep well discharge proposal would be difficult to clear through regulatory authorities. Thus it was assumed that a surface discharge concept would be utilized. Alternative surface discharge concepts could involve controlled discharge to the Ohio River during periods of high flow or barging of the inorganic wastewater residue to the Gulf of Mexico.

It is the contractor's assessment that there is no means of disposal of inorganic wastewater residues that is generally acceptable - that is, there are no panaceas for disposal of wastewater residues to watercourses. Parenthetically, approval of concepts for disposal of wastewater residues to watercourses can often be negotiated on a case by case basis.

The essential elements of an acceptable concept for the disposal of inorganic residues are that regulatory guidelines be addressed, that the impacts on receiving waters and lower riparian users are quantifiable and minimal, and that the quality of the concept is superior to prevailing installations discharging similar wastes. In the Wheeling area treated coke plant effluents are discharged directly to the Ohio River, therefore a situation exists in which a logical case could be made for the disposal of gasification plant inorganic wastewater residues by concepts such as store and discharge to the river or store and barge to the Gulf of Mexico. The results of discharge permit negotiations are dependent upon the prevailing status of regulations and guidelines, community reactions, lower riparian reactions, and negotiator personalities. In many cases, the outcome of negotiations cannot be predicted with certainty.

The results of an analysis of present value unit costs are given in Table 6-18 for disposal at Wheeling, West Virginia of wastewater blowdowns by store and discharge, store and barge to sea and evaporation/incineration. The results pertain to gasifier operation air blown and oxygen blown on Illinois No. 6 coal. The results indicated that direct disposal by store and discharge (i.e. sans intermediate concentration) was the least expensive disposal alternative for all situations tabulated.

Table 6-18: Optimized Present Value Unit Costs for Disposal of Wastewater Blowdowns at Wheeling, West Virginia Site

| | Store and Discharge ($/Ton) | Store and Barge ($/Ton) | Evaporate & Incinerate ($/Ton) |
|---|---|---|---|
| Illinois No. 6 - Air Blown: | | | |
| Single Quench | 0.73 | 5.02 | 1.72 |
| Double Quench | 0.73 | 3.55 | 1.33 |
| Illinois No. 6 - Oxygen Blown: | | | |
| Single Quench | 0.65 | 3.95 | 1.92 |
| Double Quench | 0.65 | 3.03 | 1.50 |

Optimum costs of disposal of biotreated condensates from conventional single quench gas cleaning by store and barging were obtained with inclusion of intermediate two-stage reverse osmosis coupled with multiple effect evaporation to yield a concentrate with 9.0 percent strong electrolyte content for delivery to the barges. Optimum costs of disposal of biotreated condensates from two stage quenching by store and barging were obtained by concentration of the secondary condensate by reverse osmosis followed by multiple effect evaporation of the combined primary condensate and reverse osmosis concentrate to a strong electrolyte content of 9.0 percent.

Optimum costs for disposal by thermal evaporation/incineration were obtained with feeds of biotreated condensates that were preconcentrated to 9.0 percent strong electrolyte content as per disposal by store and barge to sea.

### 6.6.2 Disposal at Carson City Site

The disposal of inorganic wastewater residues at the Carson City site would presumably be accomplished by discharge to natural evaporation ponds inasmuch as suitable surface watercourses are unavailable. The ponds would be sealed against infiltration and the groundwater underlying the ponds would be monitored by wells to verify the integrity of the sealing procedure. There is an outside chance that deep well disposal would be applicable at Carson City, but the geology of the area has been identified as being generally unfavorable (26).

The cost of disposal by evaporation ponds is largely dependent upon the magnitude of the discharge flow. Therefore an economic trade-off analysis can be made between the cost of reduction of flow of wastewater blowdown by reverse osmosis and the cost of evaporation pond capacity.

The results of the analysis indicate that the present value unit costs of reverse osmosis were considerably lower than those of pond evaporation for the condensate flow rates considered in this study. As a result of this observation, the proposed process scheme for each operating case was based on maximum flow of condensate to the reverse osmosis unit.

For the case of single stage quench, both air and oxygen blown, the entire condensate flow was sent to a reverse osmosis unit which included partial recycle of the permeate. The maximum total strong electrolyte concentration for the reverse osmosis concentrate was set at 9000 mg/liter. The concentrate stream comprised the total flow to the evaporation pond.

For the case of two stage quench, both air and oxygen blown, the second stage condensate was sent to the reverse osmosis unit. The performance criteria of the reverse osmosis unit was based on the maximum allowable sulfate concentration of the concentrate and was set at 6000 mg/l for this study. The total flow to the evaporation pond consisted of the first stage quench condensate and the concentrate from the reverse osmosis unit.

Present value unit costs are presented in Table 6-19 for the disposal scheme consisting of reverse osmosis and pond evaporation. As can be observed, the costs are quite low and are essentially the same for single quench and double quench systems.

An alternate disposal scheme consisting of reverse osmosis, thermal evaporation, and incineration was also considered for all cases. The present value unit costs of this disposal technique are also shown in Table 6-19. The alternate scheme for the single stage quench case specifies that the condensate flow be sent to a reverse osmosis unit and that the concentrate undergo further processing via thermal evaporation. The design criteria for the kettle type evaporation units was based on the evaporator bottoms having a dissolved solids concentration of 60 percent. The bottoms from the evaporator were then sent to an incineration unit.

For the case of the two stage quench, both air and oxygen blown, the processing is similar to that previously described with the exception of an additional flow stream. The condensate from the first stage quench is combined with the concentrate from the reverse osmosis unit and then sent to the evaporator.

The present value unit costs of the alternate disposal scheme are significantly higher than those specified for the pond evaporation method. These higher unit costs can be attributed to the energy demands of thermal evaporation and incineration.

Table 6-19: Optimized Present Value Unit Costs for Disposal of Wastewater Blowdowns at Carson City, Nevada Site

| | Pond Evaporation ($/Ton) | Evaporation and Incineration ($/Ton) |
|---|---|---|
| Montana Rosebud - Air Blown: | | |
| Single Quench | 0.16 | 0.94 |
| Double Quench | 0.164 | 1.06 |
| Montana Rosebud - Oxygen Blown: | | |
| Single Quench | 0.15 | 0.90 |
| Double Quench | 0.158 | 0.98 |

GENERAL DISCUSSION

# 7.0 DISCUSSION

## 7.1 General

The study analyzed alternative gas cleaning systems employing single stage and two stage gas quenching for a particular gasifier installation. The results of the study indicated that the alternative of preliminary absorption of halides by application of a two stage quench was a cost effective design procedure for operation of the selected gasifier installation on high halide coals. The results also indicated the alternative of the conventional single stage quench design to be cost effective for operation of the selected gasifier installation on low halide coals. Subject to the assumptions and constraints employed in the study, the alternative of the two stage quench possessed a cost advantage of 0.99 \$/ton and 1.07 \$/ton for gasifier operation air blown and oxygen blown respectively on the selected Illinois No. 6 coal. Parenthetically, the analysis of the alternative of the single stage quench yielded a cost advantage of 0.95 \$/ton and 1.26 \$/ton for gasifier operation air blown and oxygen blown respectively on the selected Montana Rosebud coal. The dominant variable affecting the cost effectiveness of the alternatives was indicated to be the halide content of the coal feed. Under the conditions of the study, the trade-off cost between single stage and two stage gas quenching systems was indicated to occur at coal total halide concentrations of 0.15 percent and 0.16 percent (expressed as chlorides) respectively for air blown and oxygen blown gasification.

Some discussion is warranted of the validity and applicability of the aforestated findings. Firstly, it should be recognized that the findings were derived using a rational water management analysis using representative water quality criteria in cooling tower circuits. The application of the procedure enabled the comparison of alternatives on the basis of equal reuse in recycle circuits and equal blowdown to discharge. It may be noted that the single stage quench alternative would be favored in situations with ample, inexpensive water supply and relaxed effluent standards. The two stage quench alternative would be favored in situations where water supply is restricted or where effluent regulations are stringent.

Reverse osmosis characteristically passes about 17 percent of feed chloride and 1-5 percent of feed sulfate - a performance ratio of roughly five-fold. It follows that the low halide, a sulfate bearing secondary condensate from the two stage quench alternative has exceptional potential for tight closeup using reverse osmosis recycle loops. For example, if the biotreated secondary condensate for operation on Illinois No. 6 - air blown were upgraded by reverse osmosis treatment, the permeate might be projected to contain 1 mg/l of halide, 4 mg/l of sulfate and 39 mg/l of ammonia. A permeate from a biotreated single quench condensate might be projected to contain 380 mg/l of halide, 3 mg/l of sulfate and 41 mg/l of ammonia. Organic matter would be substantially absent from both permeates. Thus the permeate from the secondary condensate is of far superior quality for reuse in that it approaches the expected quality of evaporator condensate. The water quality advantage of the two stage quench warrants consideration relative to alternative selection but it was not fully evaluated in the study in terms of economics.

The findings of the cost analysis were based on the presence of an absorber with 9 sieve trays in the two stage quench system. The absorption capability was selected to achieve near equilibrium in respect to ammonia absorption - which fixed pH and thereby facilitated the theoretical analysis of the chemical characteristics of the condensate. In actuality, the analysis indicated that over 90 percent of the chlorides and fluorides were absorbed after one theoretical tray. It follows that a venturi scrubber followed by an efficient moisture separator and mist eliminator would likely provide sufficient absorption capability for capture of chlorides and fluorides. Such a system would minimize the cost impact of the absorber on the two stage quench alternative.

The aforementioned findings of cost advantage for alternative systems did not allow for costs associated with increased energy loss through the two stage quench system. It was reasoned that the degree of gas cleaning for feed to the electrostatic precipitator should be the same for the single stage and two stage quench alternatives. It followed that the total energy requirement for cleaning the gases should be approximately the same whether accomplished by a venturi scrubber alone or a venturi scrubber in conjunction with a sieve tray absorber. That is, in a two stage system it should be possible to adjust the energy loss in the venturi scrubber to essentially compensate for the energy loss in the absorber.

Opportunities exist for improvement of the cost effectiveness of the two stage quench alternative through optimization of process engineering. The excellent absorption characteristic of the halides suggests that the unit flow incorporated in the first stage quench could be sized for the minimum flow to transport tar and particulates (possibly 10 to 20 gallons/ton). The gain would be more secondary condensate available for recycle which would improve the cost advantage indicated for the two stage quench alternative with Illinois No. 6 coal. Another potential design refinement might consist of the conversion of the tar decanter in the second stage quench to a side stream device inasmuch as over 99 percent of the total tar collected is removed in the first stage. Such revision could provide a decanter size reduction of possibly 75 percent at a significant cost reduction.

Some reflection is appropriate concerning possible limits to the applicability of the two stage quench concept. The concept is simple and is considered to have extensive application to coal conversion process gas cleaning, but it does entail the inclusion of additional decanter units, and possibly absorption units, in the gas train. Since the cost of tanks and vessels increases with increased pressure rating, it may be presumed that the feasibility of the two stage quench is adversely affected by operation at high pressure conditions. The cost impact of the pressure limitation could be particularly severe if the supplementary units (decanter, etc.) had to be fabricated from exotic materials (e.g. titanium) to withstand the aggressive condensate characteristics due to the presence of halides in combination with low pH. However, a situation could conceivably prevail whereby corrosion problems were confined to the first stage system enabling the use of less expensive materials in subsequent units.

Problems of tar fouling would be expected to be diminished somewhat by application of two stage gas quenching because the venturi scrubber and decanters are operated nearer the temperature of tar condensation. Hard tars are expected to be collected by a tar scrubber that precedes the gas stream heat recovery unit immediately upstream of the venturi scrubber. Soft tars would be collected in the first stage quench and would be held at about 225°F in the first stage decanter. The second stage quench will condense moisture and light oil - it can be presumed that captured tar will be dissolved or entrained in the emulsion. Thus the staged temperature situation should assist in the limiting of tar deposition.

It is noted that tar collection and fouling was evaluated in METC pilot plant performance tests (1). In the tests for gas cleanliness after venturi and valve tray scrubbing, twenty liters per minute of gas were passed through an 0.8 micron filter for periods of 20 hours per test. "The filter when observed under scanning electron microscope showed: (1) no particles of any size, and (2) no stain or discoloration incident to tar or light oil deposition." The tests on a closed circuit quench system indicated that tars and particles that remained entrained in the decanter overflow passed the heat exchangers "easily." No increase in pressure drop across the heat exchangers was observed with time of service and the heat transfer coefficients remained at 100 to 105 Btu/hr-ft$^2$-°F. It was concluded that the system represented "a series of highly functional unit operations." Steam tracing of the tar lock hopper and discharge valve was the only external aid employed for tar removal.

The cost analysis employed in the study presumed the applicability of carbon steel vessels and tanks in the gas train. The presumption is considered sound pending experimental evaluation of carbon steel in the service and the possibility of corrosion control by use of neutralization and/or chemical inhibitors. In view of the present lack of data, the possibility exists that removal of corrosives in a first stage quench could register as a cost advantage by localizing the need for exotic materials of construction in the gas train. A supplemental review of materials of construction is presented in Section 11, Appendix I.

## 7.2 Accuracy of the Modified WAVES Program

The results obtained with the original WAVES program has shown good agreement with experimental data for a system containing volatile weak electrolytes (e.g., $NH_3$ - $H_2S$ - $H_2O$ and $NH_3$ - $CO_2$ - $H_2O$, Ref. (5)). The program used for purposes of this study is based on the original WAVES program with modifications made to include chemical species pertinent to the proposed system. Data was unavailable for a similar system containing all of the weak electrolytes in the modified WAVES program. However, the results from the program were compared to data from a study of sour water stripping (24) and reasonable correlation was obtained. Although the predictions of the program were not verified completely, the trends were in-line with expectations.

The solubility of ammonia, for instance, is increased in the quench streams due to the presence of strong acids. The acid gases, $CO_2$, $H_2S$ and HCN are made more volatile in this environment. Another expected trend is the nearly complete absorption of HCl and HF with minimal quenching. The large ionization potentials and thus, low vapor pressures for these chemical species indicated that such a phenomenon could be expected. These are qualitative interpretations of the results. However, experimental data for a similar system would be required for verification of study projections.

The quench unit operates at concentrations and temperatures which are within the range for which the theoretical basis is valid (5) (7). Additional work is necessary to account for the molecule-molecule and molecule-ion interaction parameters for the HCl and HF species. At the time of the study these parameters were not available.

This study was an attempt to use a basically sound theoretical study to model a more complex system. As the theory undergoes further development to simulate more complex systems there will be considerable refinement in the designs based on this theory. Recent developments have already been made by Edwards, et. al. (7) to include the activity of the water together with the vapor phase fugacity coefficient. The improvements also incorporated the molecular and interaction parameters as functions of temperature.

The program used for this study did not take into account the reaction of cyanide with sulfur to form thiocyanate. The reaction mechanisms possible for the formation of the thiocyanate are numerous and complex in a multicomponent mixture such as the quench water from a gasification plant. Current environmental studies of coal gasification plants in conjunction with bench scale studies should generate considerable information about the reaction kinetics of sulfur compounds with cyanide.

## 7.3 Thiocyanate Formation

Most of the sulfur contained in the producer gas is in the form of $H_2S$ (27). The quantity of hydrogen sulfide and hydrogen cyanide that is present in the first and second stage quench water was predicted using the modified WAVES program. There are several reaction mechanisms that have been proposed to describe the formation of thiocyanate in the aqueous phase (28). Most of the mechanisms proposed require the presence of an oxidizing agent (28). One possible mechanism is the oxidation of sulfide to form sulfur and ultimately polysulfide. The polysulfide then reacts with cyanide to form thiocyanate. Another possible mechanism proposes that the sulfide is oxidized to form sulfite which can then react with polysulfide to form thiosulfate. The thiosulfate then reacts with cyanide to form thiocyanate. There are additional reaction mechanisms possible for the formation of thiocyanate but at the operating conditions of the proposed system the previously mentioned reactions seem most credible. Results of bench scale experiments performed by Luthy, et al. (28) indicate that the oxidation of sulfide to yield polysulfide is the rate controlling step for the formation of thiocyanate. The sulfide oxidation reaction is a complex mechanism and is not well understood. Chen and Morris (29) and O'Brian and Birkner (30) have recently reported kinetic data for the oxidation of sulfide in aqueous solutions. The overall reaction rate is a function of the solution pH, the sulfide/oxygen ratio, and the catalytic effects of metal ions and organic species present in the system. There also exists an interdependent relationship between two of these variables for the system. As the pH of the condensate changes, the extent of $H_2S$ dissociation changes, thus affecting the sulfide ion concentration in solution. This in turn affects the rate of polysulfide formation.

It has been reported that carbon disulfide is reacted with ammonia to form ammonium thiocyanate in a commercial process (31). Given the proper conditions and the presence of carbon disulfide or substitute compound, it is conceivable that similar reactions could take place in the system.

In addition to the reactions mentioned previously the cyanide ion can also react with many metal cations to form stable complex anions (32). Iron cyanide complexes are the most stable of the metal anions and the most likely to be present in the quench water. The rate of formation of iron cyanide complexes is slow compared to the formation of complexes containing nickel, copper and zinc (25). The source of iron for the formation of these complexes could be the carbon-steel process vessels or the ash from the gasified coal.

The chemistry of the thiocyanate formation is complex and is not a principal objective of this report. A conservative approach was taken in predicting the amount of thiocyanate that would be present in the second stage condensate. It was assumed that 70 percent of the cyanide contained in the condensate from the second stage quench would undergo further reaction to form thiocyanate. Of the total cyanide ion remaining in the condensate, it was assumed that forty percent would be removed by stripping. This value was based on the results of an API study on sour water stripping (24).

Extensive fundamental study is needed to improve the basic understanding of the chemistry of coal conversion processes (27). It is conceivable that with better resolution of the mechanisms and points of formation of thiocyanate that the gasifier could be operated in a manner to minimize the production of products that contribute to thiocyanate formation in the secondary condensate. In this regard, some success has been achieved with the operation of blast furnaces to minimize the formation of cyanide. The resolution of mechanisms would also enable the analysis of the feasibility of the feed of ammonium polysulfide to the first stage quench to react with cyandide. The use of ammonium polysulfide has been advocated for the conversion of cyanide to thiocyanate in some sour water stripper and desulfurization systems. The secondary condensate would possess a lower dissolved solids concentration if the concentration of cyanide in the gas could be lowered or if the thiocyanate formation could be promoted in the first stage quench.

## 7.4 Differential Energy Consumption Between the Single and Two-Stage Quench Processes

It was of interest to consider any differences in energy requirements for the single stage quench and two stage quench alternatives.

Thermal energy balances for each of the processing schemes considered are near identical. The major energy discrepancy between the two systems is a result of the higher pressure drop the product gas undergoes through the two stage quench units. Calculations predict a .6 psi pressure drop for the single stage spray coolers and a 2.6 psi pressure drop for the combined absorber and spray cooler units in the two stage quench system. For purposes of gas distribution or for meeting elevated pressure requirements of process units downstream of the gas treatment facilities, the available product gas pressure has an impact on compression costs.

To compute the cost of the additional energy requirements for the two stage quench process, it was assumed that the product gas would be at the same delivery pressure for each process. To meet this requirement, it was necessary to increase the pressure delivered by the air/oxygen compressor to the inlet of the gasifier for the two stage quench process. The additional energy costs attributed to the increase in compressor load for the two stage process were computed for each of the gasification cases considered. It was assumed that the compressor is a two stage unit having an overall efficiency of 75 percent. Power costs were based on current industrial rates of 4.0 cents per kilowatt hour.

The results of these calculations are shown in Table 7-1. These additional operating costs were incorporated into the overall economics to establish their effect on the total incremental annualized costs. As can be seen in Table 7-1, these additional operating costs for the two stage quench process are significant, but do not affect the most economical choice for each case.

Table 7-1: Estimated Increased Energy Consumption for the Two Stage Quench System and the Effect on the Economic Analysis in Table 6-17.

| | Illinois No. 6 Air Blown | Illinois No. 6 $O_2$ Blown | Montana Rosebud Air Blown | Montana Rosebud $O_2$ Blown |
|---|---|---|---|---|
| Most Economical Process as per Table 6-17 | Two Stage | Two Stage | Single Stage | Single Stage |
| Increased Energy Consumption for Two Stage System, MM Kwhr | 3.6 | 2.42 | 2.03 | 1.17 |
| Additional Energy Costs, MM $/yr | .144 | .097 | .081 | .047 |
| Adjusted Table 6-17 Incremental Operating Cost Advantage MM $/yr | .176 | .243 | .221 | .197 |
| Adjusted Table 6-17 Total Incremental Annualized Cost Advantage Over Alternative System, MM $/yr | .376 | .463 | .581 | .607 |

The two stage quench system offers a potential opportunity for energy conservation in that the heat extracted in the first stage quench is of sufficient availability for the heating of biological wastewater treating units, such as activated sludge aeration tanks, during cool weather. The energy recovery concept could feature the circulation of decanter overflow through coils in the aeration tank enroute to the venturi scrubber. The heat abstracted would displace steam allocated to the heating of the aeration tank plus fan energy for the cooling tower associated with the displaced heat exchanger.

### 7.5 Economics of a Modified Two-Stage Quench Process

The results of the absorber performance predictions indicate that essentially all the chlorides and fluorides may be removed on one theoretical stage. This is due to the relatively low volatility of hydrogen chloride and hydrogen fluoride and their high ionization potential values. The results suggest that a scaled down absorber unit or a combination venturi scrubber and multiple mist eliminator arrangement could achieve the desired strong electrolyte removal capability. As stated before, the minimum quench rate is based on that amount of water which is necessary to effectively remove tars and oils from the system. This level of quench can be maintained for the alternate processing schemes by reducing the quench water temperature or increasing the quench water feed flow rate depending on economics and technical constraints.

The economic projections for the single and two stage quench processes were modified to incorporate the capital cost savings realized as a result of using a smaller absorber unit. The absorber unit was specified with two trays based on the theoretical tray requirement of one tray and a tray efficiency of 50 percent. The diameter of the absorber units remained unchanged for each case and the vessel height was reduced according to the number of trays removed. The economic results for each case are shown in Table 7-2. As can be observed, a size reduction of the absorber units has a significant effect on the total capital cost requirement for each case. The outcome does not alter the original predictions for the most economical process scheme for each case, but does suggest that optimization of the two stage quench process design could substantially improve its cost effectiveness. Logic suggests that the energy requirement to perform the gas cleaning function should be essentially the same for single stage and two stage quench systems.

Table 7-2: Comparative Economic Analysis of Single Stage and Modified Two Stage Gas Cleaning System

| | Illinois No. 6 Air Blown | Illinois No. 6 $O_2$ Blown | Montana Rosebud Air Blown | Montana Rosebud $O_2$ Blown |
|---|---|---|---|---|
| Most Economical Process as per Table 6-17 | Two Stage | Two Stage | Single Stage | Single Stage |
| Incremental Capital Cost Advantage, MM $ | 2.035 | 2.05 | 1.617 | 1.963 |
| Incremental Operating Cost Advantage, MM $/yr | .32 | .34 | .14 | .15 |
| Annualized Incremental Capital Cost Advantage, MM $/yr | .33 | .33 | .258 | .314 |
| Total Incremental Annualized Cost Advantage Over Alternative System, MM $ | .65 | .67 | .398 | .464 |
| Percentage Change in Total Incremental Annualized Cost in Table 6-17 | + 25% | + 19.6% | - 20.4% | - 29.7% |

## 7.6 Disposal Methods

Methodology for the disposal of wastewater residues or solid waste residues is often impossible to define prior to maturation of regulatory negotiations. In the case of a proposed discharge of a treated wastewater blowdown from a site in the vicinity of Wheeling, West Virginia, store and discharge to the Ohio River was indicated to be relatively inexpensive and on par with the quality of prevailing wastewater discharge practices. However, new installations are generally required to provide exemplary wastewater treatment and disposal facilities. It is quite conceivable that the discharge of salts and trace levels of target substances would be unacceptable in that the water is reused downstream many times prior to reaching the Gulf of Mexico. The downriver uses include public water supply, agricultural irrigation, recreation, fishing and industrial service - all of which would be theoretically impacted by increased salinity (even though it may be infinitesimal).

In the present state of development, evaporation to dryness is not considered as an alternative that features general applicability. On first analysis, it possesses inherent limitations of excessive consumption of energy and economic resources. Additional limitations surface upon consideration of equipment life and of disposal of the highly leachable dry residue. Regulatory perspectives concerning the disposal of such materials can be very restrictive in that the necessity of disposal of residues may not be recognized. However, as energy conservation and environmental perspectives mature, it may become apparent that a potential opportunity exists for the production of a road salt product using heat of low availability from coal conversion facilities. The concept seems more rational than the present practice of discharging salt bearing treated wastes into watercourses in addition to mining salt for application to road surfaces which subsequently enters watercourses as runoff.

The disposal of inorganic wastewater blowdowns to evaporation ponds at a site in the vicinity of Carson City, Nevada appears to be a practical and presumably acceptable alternative. Regulations, aesthetics and the availability of suitable land area are possible limitations.

The water management concepts developed in the study are consistent with the objective of convergence on "zero discharge." The low volume concentrates obtained from application of two stage gas quenching, or by desalination, are amenable to processing by disposal methods that produce dry residues.

CONCLUSIONS
RECOMMENDATIONS

## 8.0 CONCLUSIONS AND RECOMMENDATIONS

The study developed an assortment of findings relative to the selected gasifier systems. The following conclusions and recommendations are made on the basis of the findings and interpretations of principal significance:

8.1 The results indicated that the alternative of preliminary absorption of halides by application of a two stage quench system was a cost effective design procedure for operation of the selected gasifier installation on Illinois No. 6 coal containing 0.25 percent chloride and 0.012 percent fluoride. Subject to the assumptions employed in the study, the alternative of the two stage quench possessed a cost advantage over a single quench system of 0.99 $/ton and 1.07 $/ton respectively for gasifier operation air blown and oxygen blown.

8.2 The results indicated that a conventional single quench system was the cost effective design procedure for operation of the selected gasifier installation on Montana Rosebud coal containing 0.03 percent chloride and 0.003 percent fluoride. Subject to the assumptions employed in the study, the single quench alternative possessed a cost advantage over a two stage quench system of 0.95 $/ton and 1.26 $/ton respectively for gasifier operation air blown and oxygen blown.

8.3 The dominant variable affecting the cost effectiveness of the alternatives was indicated to be the halide content of the coal feed for situations involving substantial reuse of gas condensate in cooling tower circuits. In terms of cost effectiveness, the trade-off coal halide concentrations between single quench and two stage quench designs was indicated as 0.15 percent and 0.16 percent, expressed as chloride, for gasifier operation air blown and oxygen blown respectively.

It was concluded that the two stage quench would be widely applicable to coal conversion processes employing wet cleaning of halide bearing gases such as are characteristic of many Eastern coals.

8.4 The study demonstrated that a rational comparison of process oriented and treatment oriented water management alternatives is attainable by application of water quality criteria to reuse circuits followed by economic analysis.

8.5 A refinement of the engineering and economic analyses of the study is recommended. The broad base thermodynamic analysis employed in the study was contingent upon the assumption of equilibrium conditions in respect to target components at selected points in the gas train. The analysis of the results indicated that the formulation of a gas train design to make the equilibrium assumptions realistic resulted in overdesign relative to the removal of chlorides and fluorides. It is believed that the gas train units required for a first stage quench to collect tar, particulates and halides could consist of a venturi scrubber, knockout drum and an efficient mist eliminator. In addition, it is believed that the second stage tar decanter could be designed as a sidestream device of reduced size inasmuch as over 99 percent of the total tar collected is indicated to be removed in the first stage. Although the suggested design revisions would improve the economics of the two stage quench system, the system possesses the inherent economic burdens of a more complex controls system, an extra tar decanter, and dual train condensate processing systems.

8.6 The two stage quench system employed in the study featured a larger energy loss in the gas train than the conventional single quench system. However, logic suggests that the energy losses from single stage and two stage gas trains would be similar for comparable performance in terms of particulate removal and gas absorption. The design revisions suggested under Item 8.5 would do much to provide for equalization of energy consumption between the alternative systems.

8.7 Alternative control concepts were suggested to achieve the collection of a selected fraction of the total gas condensate in a first stage quench. Both control concepts are considered to be viable from the engineering standpoint.

8.8 The complexity of the chemistry of thiocyanate formation induced resort to empirical projections of thiocyanate concentrations in the gas condensates. Experimental verification of expected thiocyanate residuals in secondary condensates is recommended inasmuch as biological treatment for degradation of organic matter will convert thiocyanate to ammonia and residual sulfate, a strong electrolyte. The presence of strong electrolyte compromises the potential for reuse of the condensate.

8.9 The selection of materials of fabrication for gas train units was identified as an enigma. Study estimates were based on carbon steel decanters and vessels although it was recognized that the presence of halides and other aggressive substances, possibly in conjunction with low pH, could conceivably impair the serviceability of the system. Coke plant experience suggests that tar coatings render carbon steel as an acceptable material for many corrosive gas train environments. The situations involved in the study differ somewhat from coke plant experience, therefore it is recommended that the materials enigma be resolved by experimental investigation including evaluation of pH control and corrosion inhibitors. It is noted that there could be significant materials differences between single stage and two stage gas quench systems, with no advantage evident for either system.

8.10 The treatment of gas condensates by biological processes is considered to be technically feasible on the basis of an extended although somewhat inconsistent performance record. Developmental studies are suggested to verify process applicability to specific situations.

8.11 The desalination of effluents from biotreatment processes by reverse osmosis is considered to be technically feasible on the basis of a limited record of inconsistent performance. It is speculated that the consistency of performance could be improved to acceptable levels by application of developmental studies incorporating upgraded pretreatment applicable to operation on the inorganic wastewater blowdown.

8.12 Reverse osmosis was indicated to be substantially less expensive than thermal evaporation for desalination of inorganic wastewater blowdowns. Thermal evaporation was practical for the desalination of reverse osmosis concentrates and condensates from first stage quenching where the unit costs of disposal exceeded evaporation costs. The existence of a substantial source of heat of low availability at coal gasification plants could improve the economic attractiveness of thermal evaporation.

8.13 Incineration with disposal of salt cake to sea, using a two stage gas quench system in conjunction with desalination by reverse osmosis and thermal evaporation, was indicated as the preferred disposal concept for inorganic wastewater blowdown from a gasification plant operating on high halide coal at a location in the vicinity of Wheeling, West Virginia.

8.14 Pond evaporation, using a single gas quench system in conjunction with desalination by reverse osmosis, was considered as the preferred disposal concept for inorganic wastewater blowdown from a gasification plant operating on low halide coal near Carson City, Nevada.

8.15 The water management concepts developed in the study are consistent with the objective of convergence on "zero discharge." The low volume concentrates obtained from application of two stage gas quenching, or by desalination, are amenable to processing by disposal methods that produce dry residues.

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Dr. K. C. Vyas - Lead Alternate Fuels Engineer
Dr. W. A. Parsons - Principal Investigator

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APPENDIX

## 11.0 APPENDIX

The evaluation of the concept of preliminary absorption of halides involved the assessment of water management effects produced by installation of alternative single stage and two stage gas quenching systems on comparable gasifiers. During the course of the investigation some detailed calculations were produced and some supporting materials were generated relative to auxiliary water management systems. These materials are included in this section as a series of Appendices to the principal theme of the study. They are as follows:

Appendix A - Gas Train Material Balances
Appendix B - Activated Sludge Treatment
Appendix C - Disposal by Store and Discharge
Appendix D - Disposal by Store and Barge
Appendix E - Disposal by Pond Evaporation
Appendix F - Reverse Osmosis Treatment
Appendix G - Economic Analysis
Appendix H - Thermal Evaporation
Appendix I - Materials of Construction

## 11.0 APPENDIX A - GAS TRAIN MATERIAL BALANCES

Detailed material balances were prepared for the single stage quench and two stage gas trains illustrated in Figures 4-2 and 4-5 in Section 4. The results of the analysis are given in Tables 11-A1 to 11-A20, with stream numbers referenced to Figures 4-2 and 4-5. The results pertain to air blown and oxygen blown gasifier operation on Illinois No. 6 and Montana Rosebud Coals with various moisture fractions condensed in the first stage quench.

## TABLE 11-A1: GAS TRAIN MATERIAL BALANCES, ILLINOIS NO. 6 COAL - GASIFIED WITH AIR, SINGLE STAGE CONDENSATION AT 110°F

| STREAM NO.* | 1 | 2 | 3 | 4 |
|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER SINGLE STAGE QUENCH | CONDENSATE FROM SINGLE STAGE QUENCH | TAR FROM SINGLE STAGE QUENCH |
| Component <u>Mass Flow (lb/hr)</u> | | | | |
| $H_2O$ | 157000. | 4734. | 152200. | |
| $NH_3$ | 820.7 | 12.49 | 808.2 | |
| $CO_2$ | 119300. | 117600. | 1634. | |
| $H_2S$ | 4540. | 4463. | 76.81 | |
| HCN | 132.5 | 106.0 | 26.48 | |
| HCl | 321.4 | $9.342 \times 10^{-10}$ | 321.4 | |
| HF | 15.87 | $4.677 \times 10^{-6}$ | 15.87 | |
| CO | 78600. | 78600. | | |
| $H_2$ | 8577. | 8577. | | |
| $CH_4$ | 7230. | 7230. | | |
| $C_2H_6$ | 1193. | 1193. | | |
| $N_2$ | 257100. | 257100. | | |
| Phenol | 563.0 | 20.90 | 506.6 | 35.48 |
| Tar/Oil | 1750. | 10.5 | | 1740. |
| Total Flow (lb/hr) | 637100. | 479600. | 155600. | 1775. |
| Temperature (°F) | 400 | 110 | 130 | 130 |
| Pressure (psia) | 95.6 | 95.0 | 95.1 | 95.1 |
| Condensate pH | | | 7.11 | |

*Reference Figure 4-2

### TABLE 11-A2: GAS TRAIN MATERIAL BALANCES, ILLINOIS NO. 6 COAL - GASIFIED WITH AIR, TWO STAGE QUENCH WITH 5% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| **Component Mass Flow (lb/hr)** | | | | | | | |
| $H_2O$ | 157000. | 149100. | 7848. | | 4743. | 144400. | |
| $NH_3$ | 820.7 | 654.6 | 166.1 | | 11.42 | 643.2 | |
| $CO_2$ | 119300. | 119300. | 2.086 | | 117600. | 1632. | |
| $H_2S$ | 4540 | 4539 | 0.2707 | | 4463. | 76.04 | |
| HCN | 132.5 | 132.5 | 0.01346 | | 107.1 | 25.42 | |
| HCl | 321.4 | $3.524 \times 10^{-4}$ | 321.4 | | $1.028 \times 10^{-15}$ | $3.524 \times 10^{-4}$ | |
| HF | 15.87 | 0.7143 | 15.16 | | $2.128 \times 10^{-7}$ | 0.7143 | |
| CO | 78600. | 78600. | | | 78600. | | |
| $H_2$ | 8577. | 8577. | | | 8577. | | |
| $CH_4$ | 7230. | 7230. | | | 7230. | | |
| $C_2H_6$ | 1193. | 1193. | | | 1193. | | |
| $N_2$ | 257100. | 257100. | | | 257100. | | |
| Phenol | 563.0 | 538.8 | 16.59 | 7.623 | 22.46 | 516.3 | 0.065 |
| Tar/Oil | 1750. | 12.22 | | 1738. | 10.5 | | 1.715 |
| Total Flow (lb/hr) | 637100. | 627000. | 8370. | 1746. | 479700. | 147300. | 1.78 |
| Temperature (°F) | 400 | 249.2 | 256 | 256 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 5.53 | | | 7.14 | |

*Reference Figure 4-5

[redacted] [redacted]

### TABLE 11-A3: GAS TRAIN MATERIAL BALANCES, ILLINOIS NO. 6 COAL - GASIFIED WITH AIR, TWO STAGE QUENCH WITH 10% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| Component Mass Flow (lb/hr) | | | | | | | |
| $H_2O$ | 157000. | 141300. | 15696. | | 4741. | 136500. | |
| $NH_3$ | 820.7 | 650.0 | 170.7 | | 12.41 | 637.6 | |
| $CO_2$ | 119300. | 119300. | 5.644 | | 117642. | 1610. | |
| $H_2S$ | 4540. | 4539. | 0.7263 | | 4464. | 74.62 | |
| HCN | 132.5 | 132.5 | 0.03319 | | 108.2 | 24.32 | |
| HCl | 321.4 | $7.683 \times 10^{-5}$ | 321.4 | | $2.264 \times 10^{-16}$ | $7.683 \times 10^{-5}$ | |
| HF | 15.87 | 0.1776 | 15.70 | | $5.334 \times 10^{-8}$ | 0.1776 | |
| CO | 78600. | 78600. | | | 78600. | | |
| $H_2$ | 8577. | 8577. | | | 8577. | | |
| $CH_4$ | 7230. | 7230. | | | 7230. | | |
| $C_2H_6$ | 1193. | 1193. | | | 1193. | | |
| $N_2$ | 257100. | 257100. | | | 257100. | | |
| Phenol | 563.0 | 520.3 | 34.68 | 8.046 | 22.87 | 497.3 | 0.064 |
| Tar/Oil | 1750. | 12.16 | | 1738. | 10.5 | | 1.663 |
| Total Flow (lb/hr) | 637100. | 619200. | 16240. | 1746. | 479700. | 139300. | 1.73 |
| Temperature (°F) | 400 | 245.9 | 253 | 253 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 5.77 | | | 7.15 | |

*Reference Figure 4-5

## TABLE 11-A4: GAS TRAIN MATERIAL BALANCES, ILLINOIS NO. 6 COAL - GASIFIED WITH AIR, TWO STAGE QUENCH WITH 20% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| Component <br> Mass Flow (lb/hr) | | | | | | | |
| $H_2O$ | 157000. | 125600. | 31390. | | 4736. | 120800. | |
| $NH_3$ | 820.7 | 639.4 | 181.3 | | 14.82 | 624.6 | |
| $CO_2$ | 119300. | 119200. | 16.63 | | 117700. | 1560. | |
| $H_2S$ | 4540. | 4538. | 2.126 | | 4466. | 71.48 | |
| HCN | 132.5 | 132.4 | 0.09381 | | 110.4 | 22.05 | |
| HCl | 321.4 | $1.496 \times 10^{-5}$ | 321.4 | | $4.521 \times 10^{-17}$ | $1.496 \times 10^{-5}$ | |
| HF | 15.87 | 0.03691 | 15.84 | | $1.132 \times 10^{-8}$ | 0.03691 | |
| CO | 78600. | 78600. | | | 78600. | | |
| $H_2$ | 8577. | 8577. | | | 8577. | | |
| $CH_4$ | 7230. | 7230. | | | 7230. | | |
| $C_2H_6$ | 1193. | 1193. | | | 1193. | | |
| $N_2$ | 257100. | 257100. | | | 257100. | | |
| Phenol | 563. | 480.7 | 73.65 | 8.692 | 23.74 | 456.9 | 0.061 |
| Tar/Oil | 1750. | 12.02 | | 1738. | 10.5 | | 1.523 |
| Total Flow (lb/hr) | 637100. | 603300. | 32000. | 1747. | 479800. | 123500. | 1.58 |
| Temperature (°F) | 400 | 240.5 | 247 | 247 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 6.10 | | | 7.19 | |

*Reference Figure 4-5

## TABLE 11-A5: GAS TRAIN MATERIAL BALANCES, ILLINOIS NO. 6 COAL - GASIFIED WITH AIR, TWO STAGE QUENCH WITH 30% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| Component Mass Flow (lb/hr) | | | | | | | |
| $H_2O$ | 157000. | 109900. | 47090. | | 4731. | 105100. | |
| $NH_3$ | 820.7 | 624.2 | 196.5 | | 17.93 | 606.3 | |
| $CO_2$ | 119300. | 119200. | 35.35 | | 117700. | 1496. | |
| $H_2S$ | 4540. | 4535. | 4.444 | | 4467. | 67.76 | |
| HCN | 132.5 | 132.3 | 0.2005 | | 112.7 | 19.67 | |
| HCl | 321.4 | $4.800 \times 10^{-6}$ | 321.4 | | $1.503 \times 10^{-17}$ | $4.800 \times 10^{-6}$ | |
| HF | 15.87 | 0.01219 | 15.86 | | $3.858 \times 10^{-9}$ | 0.01219 | |
| CO | 78600. | 78600. | | | 78600. | | |
| $H_2$ | 8577. | 8577. | | | 8577. | | |
| $CH_4$ | 7230. | 7230. | | | 7230. | | |
| $C_2H_6$ | 1193. | 1193. | | | 1193. | | |
| $N_2$ | 257100. | 257100. | | | 257100. | | |
| Phenol | 563. | 437.0 | 116.6 | 9.357 | 24.61 | 412.3 | 0.059 |
| Tar/Oil | 1750. | 11.92 | | 1738. | 10.5 | | 1.418 |
| Total Flow (lb/hr) | 637100. | 587500. | 47780. | 1747. | 479800. | 107700. | 1.48 |
| Temperature (°F) | 400 | 234.8 | 242 | 242 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 6.33 | | | 7.24 | |

*Reference Figure 4-5

## TABLE 11-A6: GAS TRAIN MATERIAL BALANCES, ILLINOIS NO. 6 COAL - GASIFIED WITH OXYGEN, SINGLE STAGE CONDENSATION AT 110°F

| STREAM NO.* | 1 | 2 | 3 | 4 |
|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER SINGLE STAGE QUENCH | CONDENSATE FROM SINGLE STAGE QUENCH | TAR FROM SINGLE STAGE QUENCH |
| Component Mass Flow (lb/hr) | | | | |
| $H_2O$ | 223000. | 2556. | 220400. | |
| $NH_3$ | 822.5 | 1.795 | 820.7 | |
| $CO_2$ | 134400. | 132400. | 1967. | |
| $H_2S$ | 4545. | 4446. | 98.70 | |
| HCN | 133.0 | 80.31 | 52.69 | |
| HCl | 321.1 | $9.954 \times 10^{-10}$ | 321.1 | |
| HF | 15.64 | $4.972 \times 10^{-6}$ | 15.64 | |
| CO | 63960. | 63960. | | |
| $H_2$ | 8790. | 8790. | | |
| $CH_4$ | 10110. | 10110. | | |
| $C_2H_6$ | 1176. | 1176. | | |
| $N_2$ | 1739. | 1739. | | |
| Phenol | 563.0 | 8.137 | 529.5 | 25.32 |
| Tar/Oil | 1750. | 10.5 | | 1740. |
| Total Flow (lb/hr) | 451300. | 225300. | 224200. | 1765. |
| Temperature (°F) | 400 | 110 | 130 | 130 |
| Pressure (psia) | 95.6 | 95.0 | 95.1 | 95.1 |
| Condensate pH | | | 6.66 | |

*Reference Figure 4-2

## TABLE 11-A7: GAS TRAIN MATERIAL BALANCES, ILLINOIS NO. 6 COAL - GASIFIED WITH OXYGEN, TWO STAGE QUENCH WITH 5% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| Component<br>Mass Flow (lb/hr) | | | | | | | |
| $H_2O$ | 223000. | 211800. | 11150. | | 2559. | 209200. | |
| $NH_3$ | 822.5 | 657.4 | 165.1 | | 1.638 | 655.8 | |
| $CO_2$ | 134400. | 134400. | 4.130 | | 132500. | 1946. | |
| $H_2S$ | 4545. | 4545. | 0.4379 | | 4448. | 96.53 | |
| HCN | 133. | 133. | $2.639 \times 10^{-3}$ | | 81.93 | 51.07 | |
| HCl | 321.1 | $1.062 \times 10^{-3}$ | 321.1 | | $3.318 \times 10^{-15}$ | $1.062 \times 10^{-3}$ | |
| HF | 15.64 | 1.934 | 13.70 | | $6.228 \times 10^{-7}$ | 1.934 | |
| CO | 63960. | 63960. | | | 63960. | | |
| $H_2$ | 8790. | 8790. | | | 8790. | | |
| $CH_4$ | 10110. | 10110. | | | 10110. | | |
| $C_2H_6$ | 1176. | 1176. | | | 1176. | | |
| $N_2$ | 1739. | 1739. | | | 1739. | | |
| Phenol | 563. | 542.4 | 15.86 | 4.723 | 8.639 | 533.7 | 0.071 |
| Tar/Oil | 1750. | 13.16 | | 1737. | 10.5 | | 2.66 |
| Total Flow (lb/hr) | 451300. | 437900. | 11670. | 1742. | 225400. | 212500. | 2.73 |
| Temperature (°F) | 400 | 282.9 | 292 | 292 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 5.23 | | | 6.69 | |

*Reference Figure 4-5

### TABLE 11-A8: GAS TRAIN MATERIAL BALANCES, ILLINOIS NO. 6 COAL - GASIFIED WITH OXYGEN, TWO STAGE QUENCH WITH 10% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| Component Mass Flow (lb/hr) | | | | | | | |
| $H_2O$ | 223000. | 200700. | 22300. | | 2558. | 198100. | |
| $NH_3$ | 822.5 | 651.5 | 171.1 | | 1.781 | 649.7 | |
| $CO_2$ | 134400. | 134400. | 9.865 | | 132500. | 1913. | |
| $H_2S$ | 4545. | 4544. | 1.062 | | 4450. | 94.01 | |
| HCN | 133.0 | 133.0 | $6.581 \times 10^{-3}$ | | 83.62 | 49.37 | |
| HCl | 321.1 | $2.542 \times 10^{-4}$ | 321.1 | | $8.002 \times 10^{-16}$ | $2.542 \times 10^{-4}$ | |
| HF | 15.64 | 0.5507 | 15.09 | | $1.785 \times 10^{-7}$ | 0.5507 | |
| CO | 63960. | 63960. | | | 63960. | | |
| $H_2$ | 8790. | 8790. | | | 8790. | | |
| $CH_4$ | 10110. | 10110. | | | 10110. | | |
| $C_2H_6$ | 1176. | 1176. | | | 1176. | | |
| $N_2$ | 1739. | 1739. | | | 1739. | | |
| Phenol | 563. | 525.0 | 33.03 | 4.942 | 8.824 | 516.1 | 0.072 |
| Tar/Oil | 1750. | 13.13 | | 1737. | 10.5 | | 2.625 |
| Total Flow (lb/hr) | 451300. | 426700 | 22850. | 1742. | 225400. | 201300. | 2.70 |
| Temperature (°F) | 400 | 280.3 | 289 | 289 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 5.49 | | | 6.71 | |

*Reference Figure 4-5

## TABLE 11-A9: GAS TRAIN MATERIAL BALANCES, ILLINOIS NO. 6 COAL - GASIFIED WITH OXYGEN, TWO STAGE QUENCH WITH 20% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| Component Mass Flow (lb/hr) | | | | | | | |
| $H_2O$ | 223000. | 178400. | 44590. | | 2557. | 175800. | |
| $NH_3$ | 822.5 | 639.5 | 183.0 | | 2.136 | 637.4 | |
| $CO_2$ | 134400. | 134400. | 24.86 | | 132500. | 1843. | |
| $H_2S$ | 4545. | 4542 | 2.778 | | 4454. | 88.75 | |
| HCN | 133.0 | 133.0 | 0.01905 | | 87.22 | 45.76 | |
| HCl | 321.1 | $5.384 \times 10^{-5}$ | 321.1 | | $1.727 \times 10^{-16}$ | $5.384 \times 10^{-5}$ | |
| HF | 15.64 | 0.1251 | 15.51 | | $4.120 \times 10^{-8}$ | 0.1251 | |
| CO | 63960. | 63960. | | | 63960. | | |
| $H_2$ | 8790. | 8790. | | | 8790. | | |
| $CH_4$ | 10110. | 10110. | | | 10110. | | |
| $C_2H_6$ | 1176. | 1176. | | | 1176. | | |
| $N_2$ | 1739. | 1739. | | | 1739. | | |
| Phenol | 563. | 487.6 | 70.09 | 5.288 | 9.216 | 478.3 | 0.069 |
| Tar/Oil | 1750. | 12.92 | | 1737 | 10.5 | | 2.415 |
| Total Flow (lb/hr) | 451300. | 404400. | 45210. | 1742. | 225400. | 178900. | 2.48 |
| Temperature (°F) | 400 | 276.1 | 285 | 285 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 5.81 | | | 6.75 | |

*Reference Figure 4-5

## TABLE 11-A10: GAS TRAIN MATERIAL BALANCES, ILLINOIS NO. 6 COAL - GASIFIED WITH OXYGEN, TWO STAGE QUENCH WITH 30% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| Component Mass Flow (lb/hr) | | | | | | | |
| $H_2O$ | 223000. | 156100. | 66890. | | 2555. | 153500. | |
| $NH_3$ | 822.5 | 623.8 | 198.7 | | 2.604 | 621.2 | |
| $CO_2$ | 134400 | 134400 | 46.55 | | 132600. | 1762. | |
| $H_2S$ | 4545. | 4540. | 5.365 | | 4457. | 83.10 | |
| HCN | 133.0 | 133.0 | 0.04221 | | 91.14 | 41.82 | |
| HCl | 321.1 | $1.842 \times 10^{-5}$ | 321.1 | | $6.064 \times 10^{-17}$ | $1.842 \times 10^{-5}$ | |
| HF | 15.64 | 0.04375 | 15.59 | | $1.474 \times 10^{-8}$ | 0.04375 | |
| CO | 63960. | 63960. | | | 63960. | | |
| $H_2$ | 8790. | 8790. | | | 8790. | | |
| $CH_4$ | 10110. | 10110. | | | 10110. | | |
| $C_2H_6$ | 1176. | 1176. | | | 1176. | | |
| $N_2$ | 1739. | 1739. | | | 1739. | | |
| Phenol | 563. | 446.2 | 111.1 | 5.644 | 9.63 | 436.5 | 0.069 |
| Tar/Oil | 1750. | 12.81 | | 1737. | 10.5 | | 2.310 |
| Total Flow (lb/hr) | 451300. | 382000. | 67590. | 1743. | 225500. | 156400. | 2.38 |
| Temperature (°F) | 400 | 271.6 | 281 | 281 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 6.05 | | | 6.79 | |

*Reference Figure 4-5

### TABLE 11-A11: GAS TRAIN MATERIAL BALANCES, MONTANA ROSEBUD COAL - GASIFIED WITH AIR, SINGLE STAGE CONDENSATION AT 110°F

| STREAM NO.* | 1 | 2 | 3 | 4 |
|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER SINGLE STAGE QUENCH | CONDENSATE FROM SINGLE STAGE QUENCH | TAR FROM SINGLE STAGE QUENCH |
| Component Mass Flow (lb/hr) | | | | |
| $H_2O$ | 67960. | 3326. | 64630. | |
| $NH_3$ | 816.9 | 41.95 | 775.0 | |
| $CO_2$ | 91630. | 89870. | 1752. | |
| $H_2S$ | 1332. | 1303. | 28.46 | |
| HCN | 130.6 | 113.0 | 17.60 | |
| HCl | 38.86 | $7.756 \times 10^{-11}$ | 38.86 | |
| HF | 3.908 | $7.541 \times 10^{-7}$ | 3.908 | |
| CO | 59110. | 59110. | | |
| $H_2$ | 5845. | 5845. | | |
| $CH_4$ | 4816. | 4816. | | |
| $C_2H_6$ | 908.1 | 908.1 | | |
| $N_2$ | 181500. | 181500. | | |
| Phenol | 500.0 | 28.04 | 407.1 | 64.85 |
| Tar/Oil | 1625. | 9.75 | | 1615. |
| Total Flow (lb/hr) | 416200. | 364900. | 67650. | 1680. |
| Temperature (°F) | 400 | 110 | 130 | 130 |
| Pressure (psia) | 95.6 | 95.0 | 95.1 | 95.1 |
| Condensate pH | | | 7.52 | |

*Reference Figure 4-2

## TABLE 11-A12: GAS TRAIN MATERIAL BALANCES, MONTANA ROSEBUD COAL - GASIFIED WITH AIR, TWO STAGE QUENCH WITH 5% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| Component Mass Flow (lb/hr) | | | | | | | |
| $H_2O$ | 67960. | 64560. | 3398. | | 3325. | 61240. | |
| $NH_3$ | 816.9 | 790.8 | 26.14 | | 44.17 | 746.6 | |
| $CO_2$ | 91630. | 91620. | 4.049 | | 89890. | 1729. | |
| $H_2S$ | 1332. | 1332. | 0.1885 | | 1304. | 27.93 | |
| HCN | 130.6 | 130.5 | 0.03120 | | 113.7 | 16.84 | |
| HCl | 38.86 | $2.619 \times 10^{-6}$ | 38.86 | | $5.309 \times 10^{-18}$ | $2.619 \times 10^{-6}$ | |
| HF | 3.908 | 0.01344 | 3.895 | | $2.631 \times 10^{-9}$ | 0.01344 | |
| CO | 59110. | 59110. | | | 59110. | | |
| $H_2$ | 5845. | 5845. | | | 5845. | | |
| $CH_4$ | 4816. | 4816. | | | 4816. | | |
| $C_2H_6$ | 908.1 | 908.1 | | | 908.1 | | |
| $N_2$ | 181500. | 181500. | | | 181500. | | |
| Phenol | 500. | 468.0 | 15.45 | 16.58 | 31.70 | 436.1 | 0.097 |
| Tar/Oil | 1625. | 10.97 | | 1614. | 9.75 | | 1.219 |
| Total Flow (lb/hr) | 416200. | 411100. | 3487. | 1631. | 346900. | 64200 | 1.32 |
| Temperature (°F) | 400 | 227.0 | 230 | 230 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 6.50 | | | 7.54 | |

*Reference Figure 4-5

### TABLE 11-A13: GAS TRAIN MATERIAL BALANCES, MONTANA ROSEBUD COAL - GASIFIED WITH AIR, TWO STAGE QUENCH WITH 10% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| Component Mass Flow (lb/hr) | | | | | | | |
| $H_2O$ | 67960. | 61160. | 6796. | | 3322. | 57840. | |
| $NH_3$ | 816.9 | 783.8 | 33.07 | | 47.43 | 736.4 | |
| $CO_2$ | 91630. | 91610. | 12.83 | | 89920. | 1693. | |
| $H_2S$ | 1332. | 1331. | 0.5768 | | 1304. | 27.23 | |
| HCN | 130.6 | 130.5 | 0.07516 | | 114.5 | 16.05 | |
| HCl | 38.86 | $7.045 \times 10^{-7}$ | 38.86 | | $1.459 \times 10^{-18}$ | $7.045 \times 10^{-7}$ | |
| HF | 3.908 | $3.737 \times 10^{-3}$ | 3.904 | | $7.448 \times 10^{-10}$ | $3.737 \times 10^{-3}$ | |
| CO | 59110. | 59110. | | | 59110. | | |
| $H_2$ | 5845. | 5845. | | | 5845. | | |
| $CH_4$ | 4816. | 4816. | | | 4816. | | |
| $C_2H_6$ | 908.1 | 908.1 | | | 908.1 | | |
| $N_2$ | 181500. | 181500. | | | 181500. | | |
| Phenol | 500. | 451.0 | 31.77 | 17.20 | 32.21 | 418.7 | 0.092 |
| Tar/Oil | 1625. | 10.89 | | 1614. | 9.75 | | 1.138 |
| Total Flow (lb/hr) | 416200. | 407700. | 6917. | 1631. | 346900. | 60730. | 1.23 |
| Temperature (°F) | 400 | 224.5 | 227 | 227 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 6.76 | | | 7.55 | |

*Reference Figure 4-5

## TABLE 11-A14: GAS TRAIN MATERIAL BALANCES, MONTANA ROSEBUD COAL - GASIFIED WITH AIR, TWO STAGE QUENCH WITH 20% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| Component Mass Flow (lb/hr) | | | | | | | |
| $H_2O$ | 67960. | 54370. | 13590. | | 3315. | 51050. | |
| $NH_3$ | 816.9 | 764.7 | 52.19 | | 54.80 | 709.9 | |
| $CO_2$ | 91630. | 91590. | 40.93 | | 89980. | 1609. | |
| $H_2S$ | 1332. | 1330. | 1.756 | | 1304. | 25.61 | |
| HCN | 130.6 | 130.4 | 0.2074 | | 116.0 | 14.42 | |
| HCl | 38.86 | $1.797 \times 10^{-7}$ | 38.86 | | $3.927 \times 10^{-19}$ | $1.797 \times 10^{-7}$ | |
| HF | 3.908 | $9.774 \times 10^{-4}$ | 3.907 | | $2.040 \times 10^{-10}$ | $9.774 \times 10^{-4}$ | |
| CO | 59110. | 59110. | | | 59110. | | |
| $H_2$ | 5845. | 5845. | | | 5845. | | |
| $CH_4$ | 4816. | 4816. | | | 4816. | | |
| $C_2H_6$ | 908.1 | 908.1 | | | 908.1 | | |
| $N_2$ | 181500. | 181500. | | | 181500. | | |
| Phenol | 500. | 414.7 | 66.79 | 18.46 | 33.23 | 381.4 | 0.091 |
| Tar/Oil | 1625. | 10.84 | | 1614. | 9.75 | | 1.089 |
| Total Flow (lb/hr) | 416200. | 400800. | 13790. | 1632. | 347000. | 53790. | 1.18 |
| Temperature (°F) | 400 | 219.3 | 222 | 222 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 6.99 | | | 7.59 | |

*Reference Figure 4-5

## TABLE 11-A15: GAS TRAIN MATERIAL BALANCES, MONTANA ROSEBUD COAL - GASIFIED WITH AIR, TWO STAGE QUENCH WITH 30% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| Component <br> Mass Flow (lb/hr) | | | | | | | |
| $H_2O$ | 67960. | 47570. | 20390. | | 3308. | 44260. | |
| $NH_3$ | 816.9 | 738.9 | 77.95 | | 63.68 | 675.3 | |
| $CO_2$ | 91630. | 91540. | 81.82 | | 90040. | 1505. | |
| $H_2S$ | 1332. | 1328. | 3.355 | | 1305. | 23.69 | |
| HCN | 130.6 | 130.2 | 0.4172 | | 117.4 | 12.73 | |
| HCl | 38.86 | $7.477 \times 10^{-8}$ | 38.86 | | $1.753 \times 10^{-19}$ | $7.477 \times 10^{-8}$ | |
| HF | 3.908 | $4.142 \times 10^{-4}$ | 3.908 | | $9.193 \times 10^{-11}$ | $4.142 \times 10^{-4}$ | |
| CO | 59110. | 59110. | | | 59110. | | |
| $H_2$ | 5845. | 5845. | | | 5845. | | |
| $CH_4$ | 4816. | 4816. | | | 4816. | | |
| $C_2H_6$ | 908.1 | 908.1 | | | 908.1 | | |
| $N_2$ | 181500. | 181500. | | | 181500. | | |
| Phenol | 500. | 375.6 | 104.7 | 19.77 | 34.26 | 341.2 | 0.084 |
| Tar/Oil | 1625. | 10.73 | | 1614. | 9.75 | | 0.975 |
| Total Flow (lb/hr) | 416200. | 393900. | 20700. | 1634. | 347100. | 46820. | 1.06 |
| Temperature (°F) | 400 | 213.6 | 217 | 217 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 7.10 | | | 7.63 | |

*Reference Figure 4-5

## TABLE 11-A16: GAS TRAIN MATERIAL BALANCES, MONTANA ROSEBUD COAL - GASIFIED WITH OXYGEN, SINGLE STAGE CONDENSATION AT 110°F

| STREAM NO.* | 1 | 2 | 3 | 4 |
|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER SINGLE STAGE QUENCH | CONDENSATE FROM SINGLE STAGE QUENCH | TAR FROM SINGLE STAGE QUENCH |
| Component Mass Flow (lb/hr) | | | | |
| $H_2O$ | 56930. | 2523. | 54410. | |
| $NH_3$ | 826.0 | 32.30 | 793.7 | |
| $CO_2$ | 105900. | 104000. | 1827. | |
| $H_2S$ | 1322. | 1296. | 26.18 | |
| HCN | 130.3 | 111.5 | 18.87 | |
| HCl | 38.90 | $8.492 \times 10^{-11}$ | 38.90 | |
| HF | 3.880 | $8.024 \times 10^{-7}$ | 3.880 | |
| CO | 46540. | 46540. | | |
| $H_2$ | 12170. | 12170. | | |
| $CH_4$ | 6825. | 6825. | | |
| $C_2H_6$ | 916.7 | 916.7 | | |
| $N_2$ | 1281. | 1281. | | |
| Phenol | 500.0 | 25.00 | 398.0 | 77.02 |
| Tar/Oil | 1625. | 9.75 | | 1615. |
| Total Flow (lb/hr) | 235000. | 175700. | 57520. | 1692. |
| Temperature (°F) | 400 | 110 | 130 | 130 |
| Pressure (psia) | 95.6 | 95.0 | 95.1 | 95.1 |
| Condensate pH | | | 7.46 | |

*Reference Figure 4-2

### TABLE 11-A17: GAS TRAIN MATERIAL BALANCES, MONTANA ROSEBUD COAL - GASIFIED WITH OXYGEN, TWO STAGE QUENCH WITH 5% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| Component<br>Mass Flow (lb/hr) | | | | | | | |
| $H_2O$ | 56930. | 54090. | 2847. | | 2522. | 51560. | |
| $NH_3$ | 826.0 | 799.6 | 26.44 | | 34.09 | 765.5 | |
| $CO_2$ | 105900. | 105900. | 4.764 | | 104100. | 1802. | |
| $H_2S$ | 1322. | 1322. | 0.1942 | | 1296. | 25.69 | |
| HCN | 130.3 | 130.3 | 0.02768 | | 112.2 | 18.05 | |
| HCl | 38.90 | $2.903 \times 10^{-6}$ | 38.90 | | $6.433 \times 10^{-18}$ | $2.903 \times 10^{-6}$ | |
| HF | 3.880 | 0.01446 | 3.866 | | $3.031 \times 10^{-9}$ | 0.01446 | |
| CO | 46540. | 46540. | | | 46540. | | |
| $H_2$ | 12170. | 12170. | | | 12170. | | |
| $CH_4$ | 6825. | 6825. | | | 6825. | | |
| $C_2H_6$ | 916.7 | 916.7 | | | 916.7 | | |
| $N_2$ | 1281. | 1281. | | | 1281. | | |
| Phenol | 500. | 465.5 | 15.23 | 19.30 | 28.77 | 436.6 | 0.120 |
| Tar/Oil | 1625. | 11.00 | | 16.14 | 9.75 | | 1.251 |
| Total Flow (lb/hr) | 235000. | 230500. | 2936. | 1633. | 175800 | 54610. | 1.37 |
| Temperature (°F) | 400 | 231.1 | 234 | 234 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 6.49 | | | 7.48 | |

*Reference Figure 4-5

## TABLE 11-A18: GAS TRAIN MATERIAL BALANCES, MONTANA ROSEBUD COAL - GASIFIED WITH OXYGEN, TWO STAGE QUENCH WITH 10% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| Component <br> Mass Flow (lb/hr) | | | | | | | |
| $H_2O$ | 56930. | 51240. | 5693. | | 2520. | 48720. | |
| $NH_3$ | 826.0 | 792.1 | 33.96 | | 36.71 | 755.3 | |
| $CO_2$ | 105900. | 105900. | 14.97 | | 104100. | 1765. | |
| $H_2S$ | 1322. | 1322. | 0.5893 | | 1297. | 25.05 | |
| HCN | 130.3 | 130.3 | 0.06698 | | 113.0 | 17.21 | |
| HCl | 38.90 | $7.904 \times 10^{-7}$ | 38.90 | | $1.788 \times 10^{-18}$ | $7.904 \times 10^{-7}$ | |
| HF | 3.880 | $4.087 \times 10^{-3}$ | 3.876 | | $8.706 \times 10^{-10}$ | $4.087 \times 10^{-3}$ | |
| CO | 46540 | 46540. | | | 46540. | | |
| $H_2$ | 12170. | 12170. | | | 12170 | | |
| $CH_4$ | 6825. | 6825. | | | 6825. | | |
| $C_2H_6$ | 916.7 | 916.7 | | | 916.7 | | |
| $N_2$ | 1281. | 1281. | | | 1281. | | |
| Phenol | 500. | 448.7 | 31.32 | 20.02 | 29.24 | 419.3 | 0.121 |
| Tar/Oil | 1625. | 10.99 | | 1614. | 9.75 | | 1.235 |
| Total Flow (lb/hr) | 235000. | 227600. | 5817. | 1634. | 175800. | 51700. | 1.36 |
| Temperature (°F) | 400 | 228.4 | 231 | 231 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 6.73 | | | 7.50 | |

*Reference Figure 4-5

## TABLE 11-A19: GAS TRAIN MATERIAL BALANCES, MONTANA ROSEBUD COAL - GASIFIED WITH OXYGEN, TWO STAGE QUENCH WITH 20% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| Component Mass Flow (lb/hr) | | | | | | | |
| $H_2O$ | 56930. | 45550. | 11390. | | 2513. | 43030. | |
| $NH_3$ | 826.0 | 771.4 | 54.66 | | 42.64 | 728.7 | |
| $CO_2$ | 105900. | 105800. | 46.74 | | 104100. | 1677. | |
| $H_2S$ | 1322. | 1320. | 1.758 | | 1297. | 23.55 | |
| HCN | 130.3 | 130.1 | 0.1862 | | 114.7 | 15.48 | |
| HCl | 38.90 | $2.067 \times 10^{-7}$ | 38.90 | | $4.923 \times 10^{-19}$ | $2.067 \times 10^{-7}$ | |
| HF | 3.880 | $1.096 \times 10^{-3}$ | 3.879 | | $2.436 \times 10^{-10}$ | $1.096 \times 10^{-3}$ | |
| CO | 46540. | 46540. | | | 46540. | | |
| $H_2$ | 12170. | 12170. | | | 12170. | | |
| $CH_4$ | 6825. | 6825. | | | 6825. | | |
| $C_2H_6$ | 916.7 | 916.7 | | | 916.7 | | 0 |
| $N_2$ | 1281. | 1281. | | | 1281. | | |
| Phenol | 500. | 412.7 | 65.86 | 21.49 | 30.17 | 382.4 | 0.114 |
| Tar/Oil | 1625. | 10.87 | | 1614. | 9.75 | | 1.121 |
| Total Flow (lb/hr) | 235000. | 221700. | 11600. | 1635. | 175800. | 45860. | 1.24 |
| Temperature (°F) | 400 | 223.3 | 226 | 226 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 5.94 | | | 7.54 | |

*Reference Figure 4-5

## TABLE 11-A20: GAS TRAIN MATERIAL BALANCES, MONTANA ROSEBUD COAL - GASIFIED WITH OXYGEN, TWO STAGE QUENCH WITH 30% OF THE WATER VAPOR CONDENSING IN THE FIRST STAGE

| STREAM NO.* | 1 | 2 | 3 | 4 | 5 | 6 | 7 |
|---|---|---|---|---|---|---|---|
| DESCRIPTION | PRODUCER GAS | GAS AFTER FIRST STAGE QUENCH | CONDENSATE FROM FIRST STAGE QUENCH | TAR FROM FIRST STAGE QUENCH | GAS AFTER SECOND STAGE QUENCH | CONDENSATE FROM SECOND STAGE QUENCH | OIL FROM SECOND STAGE QUENCH |
| Component Mass Flow (lb/hr) | | | | | | | |
| $H_2O$ | 56930. | 39850. | 17080. | | 2506. | 37350. | |
| $NH_3$ | 826.0 | 743.7 | 82.39 | | 49.82 | 693.8 | |
| $CO_2$ | 105900. | 105800. | 92.18 | | 104200. | 1570. | |
| $H_2S$ | 1322. | 1319. | 3.319 | | 1297. | 21.79 | |
| HCN | 130.3 | 129.9 | 0.3776 | | 116.3 | 13.68 | |
| HCl | 38.90 | $8.731 \times 10^{-8}$ | 38.90 | | $2.226 \times 10^{-19}$ | $8.731 \times 10^{-8}$ | |
| HF | 3.880 | $4.711 \times 10^{-4}$ | 3.880 | | $1.109 \times 10^{-10}$ | $4.711 \times 10^{-4}$ | |
| CO | 46540. | 46540. | | | 46540. | | |
| $H_2$ | 12170. | 12170. | | | 12170. | | |
| $CH_4$ | 6825. | 6825. | | | 6825. | | |
| $C_2H_6$ | 916.7 | 916.7 | | | 916.7 | | |
| $N_2$ | 1281. | 1281. | | | 1281. | | |
| Phenol | 500.0 | 373.7 | 103.3 | 23.0 | 31.12 | 342.5 | 0.106 |
| Tar/Oil | 1625. | 10.75 | | 1614. | 9.75 | | 1.008 |
| Total Flow (lb/hr) | 235000. | 216000. | 17400. | 1637. | 175900. | 39990. | 1.11 |
| Temperature (°F) | 400 | 217.6 | 221 | 221 | 110 | 130 | 130 |
| Pressure (psia) | 97.5 | 95.0 | 97.0 | 97.0 | 94.9 | 95.0 | 95.0 |
| Condensate pH | | | 7.05 | | | 7.59 | |

*Reference Figure 4-5

## 11.0 APPENDIX B - ACTIVATED SLUDGE TREATMENT

A wastewater treatment concept was developed to biologically treat the bottoms from the ammonia stripper by the activated sludge process. The process flowsheet illustrated in Figure 11-B1 included storage of inflow, cooling, aeration, settling and dual media filtration. The food to microorganism ratio was 0.05 pounds of phenol per day per pound of mixed liquor volatile suspended solids. The aeration tank temperature was maintained between 75 and 90°F. Excess sludge from the bio-reactors and solids from filter backwash were concentrated by dissolved air flotation prior to dewatering on belt presses. The pressed sludge cake was spread on the coal pile for feed to the gasifier. The performance of the activated sludge plant was estimated from correlations with waste treatment results at coke plants and gasification plants.

Estimates for wastewater treatment by the activated sludge process were prepared from engineering flow sketches and equipment lists for three flow and organic loading situations. For estimation purposes the components of the activated sludge system were segregated into cost modules correlated to input flow or input phenol loading. Estimates were prepared for flow/phenol situations pertaining to 15 gpm/370 lb per day, 62 gpm/1600 lbs per day and 600 gpm/11000 lb per day. Components allocated to flow included input wastewater storage, clarifiers, deep bed filters, wastewater pumping facilities and control buildings. Components allocated to phenol loading included aeration tanks and mechanisms, as well as sludge dewatering and disposal facilities. The segregated cost modules facilitated the factoring of estimates of activated sludge systems to intermediate flow and loading situations.

The estimates of capital cost components of activated sludge treatment are summarized in Figure 11-B2 for flow and in Figure 11-B3 for phenol. Phenol was selected as the wastewater strength parameter inasmuch as there exists a broader data base for phenol than for BOD relative to coal conversion wastes. The total estimated capital cost is obtained as the sum of the flow and phenol components. Operation and maintenance costs were estimated as differences between single stage and two stage quench systems.

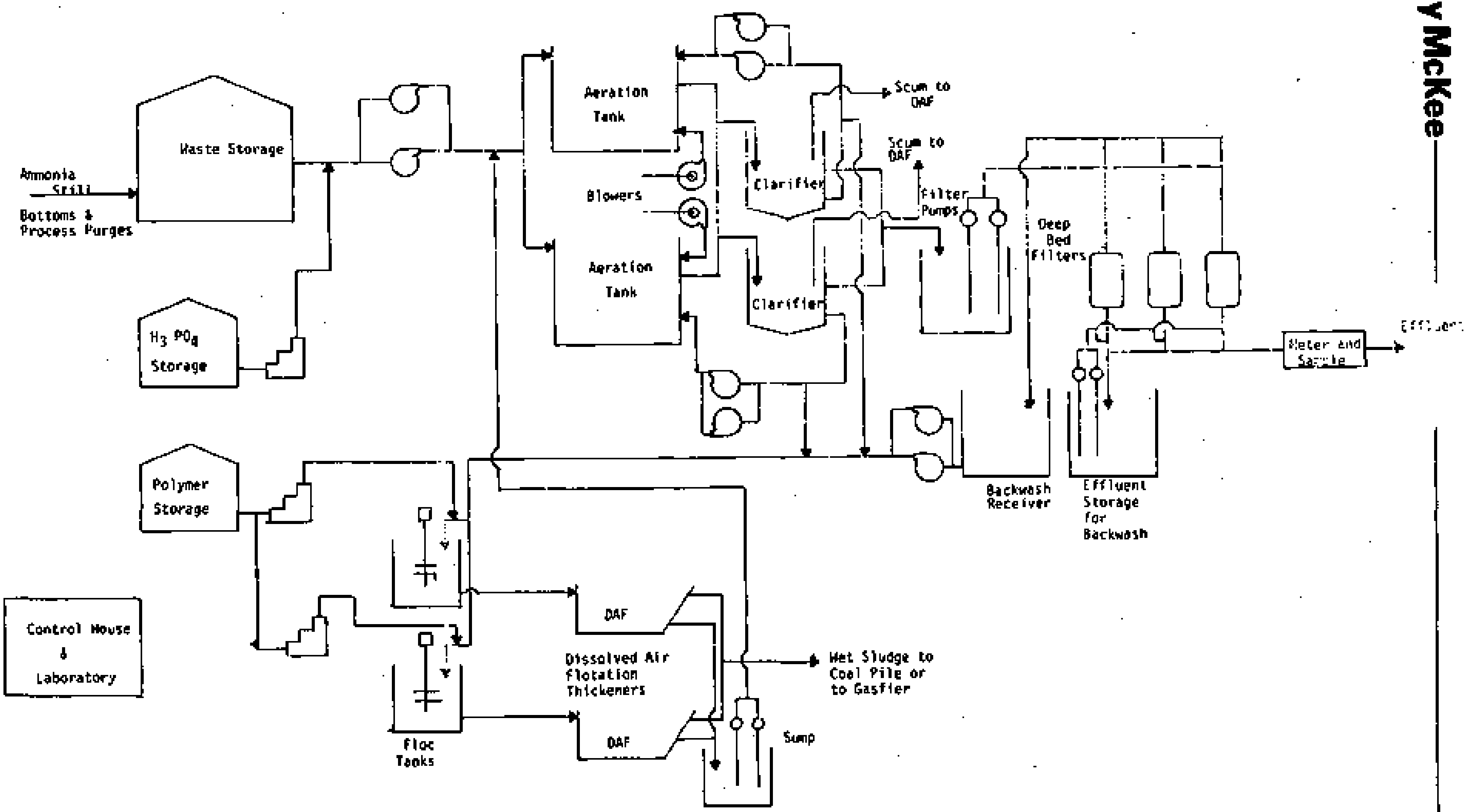

FIGURE 11-01: ACTIVATED SLUDGE MODULE

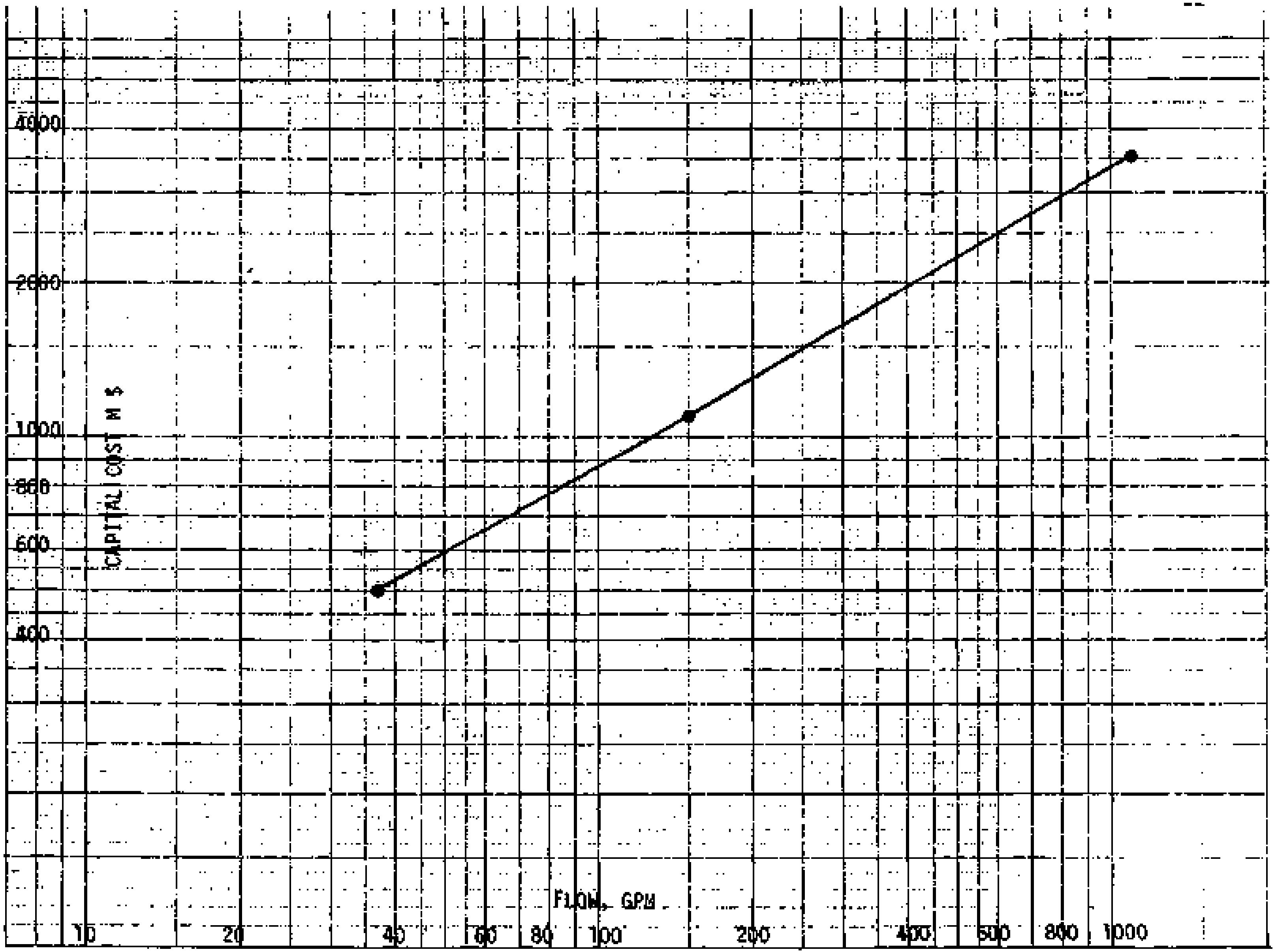

FIGURE 11-52: CAPITAL COST OF ACTIVATED SLUDGE SYSTEM COMPONENTS ASSIGNED TO FLOW FUNCTION

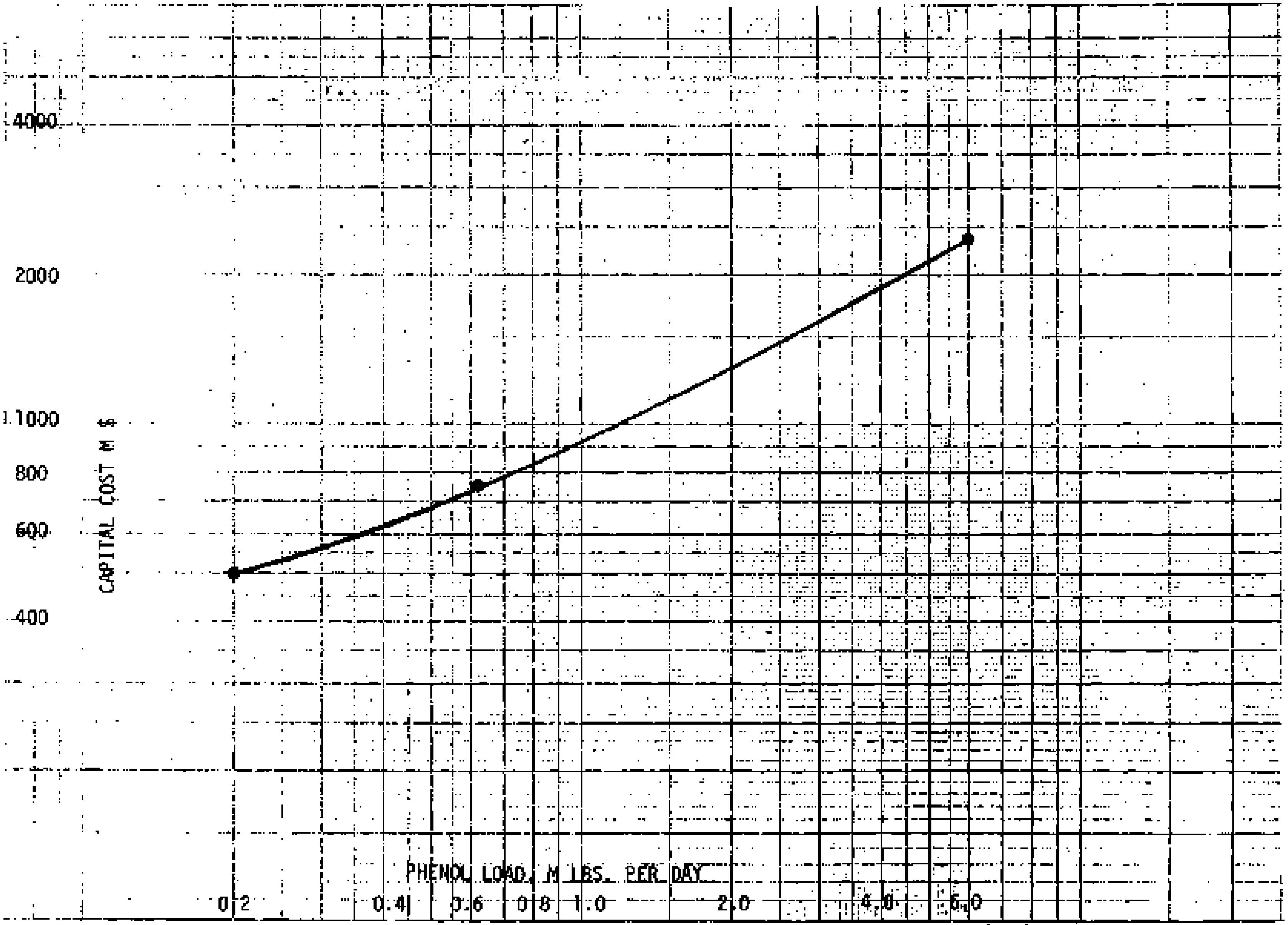

FIGURE 11-53: CAPITAL COST OF ACTIVATED SLUDGE SYSTEM COMPONENTS ASSIGNED TO PHENOL LOAD FUNCTION

## 11.0 APPENDIX C - DISPOSAL BY STORE AND DISCHARGE

A concept was prepared to receive treated effluent from the activated sludge process for disposal by store and discharge. The store and discharge concept illustrated in Figure 11-C1 provided for storage of the waste lined lagoons for a period of up to 4-months so as to enable avoidance of discharge during periods of low flow or when the chloride concentration in the receiving stream was above 250 mg/l. The concept would be inapplicable in the area of the Carson City site and would be of uncertain applicability at the Wheeling site - depending upon regulatory negotiations.

The store and discharge approach is not considered as an ideal means of disposal of inorganic wastewater residues from Wheeling, West Virginia because the inorganic salts would slightly impact a large number of lower riparian users along the Ohio and Mississippi Rivers. The concept is best suited for discharge to tidal waters. However, the concept is superior to prevailing discharges from nearby coke plants that provide no storage of treated effluent. In addition, the concept is a comparatively economical means of disposal without contravention of water quality standards (Section 4.9). Therefore, it is conceivable that regulatory or judicial authorities would consider the value of the energy production operation to offset the slightly negative aspects of the disposal concept and thereby grant approval.

For cost estimation purposes, engineering flow sketches of store and discharge concepts were prepared for flows of 20,100 and 600 gpm. The inorganic wastewater residue was held in lined earthen basins with a storage capacity of 4 months. The basins had pumped discharge to the river as per Figure 11-C1. The capital cost of the systems were estimated and the results are summarized in Figure 11-C2.

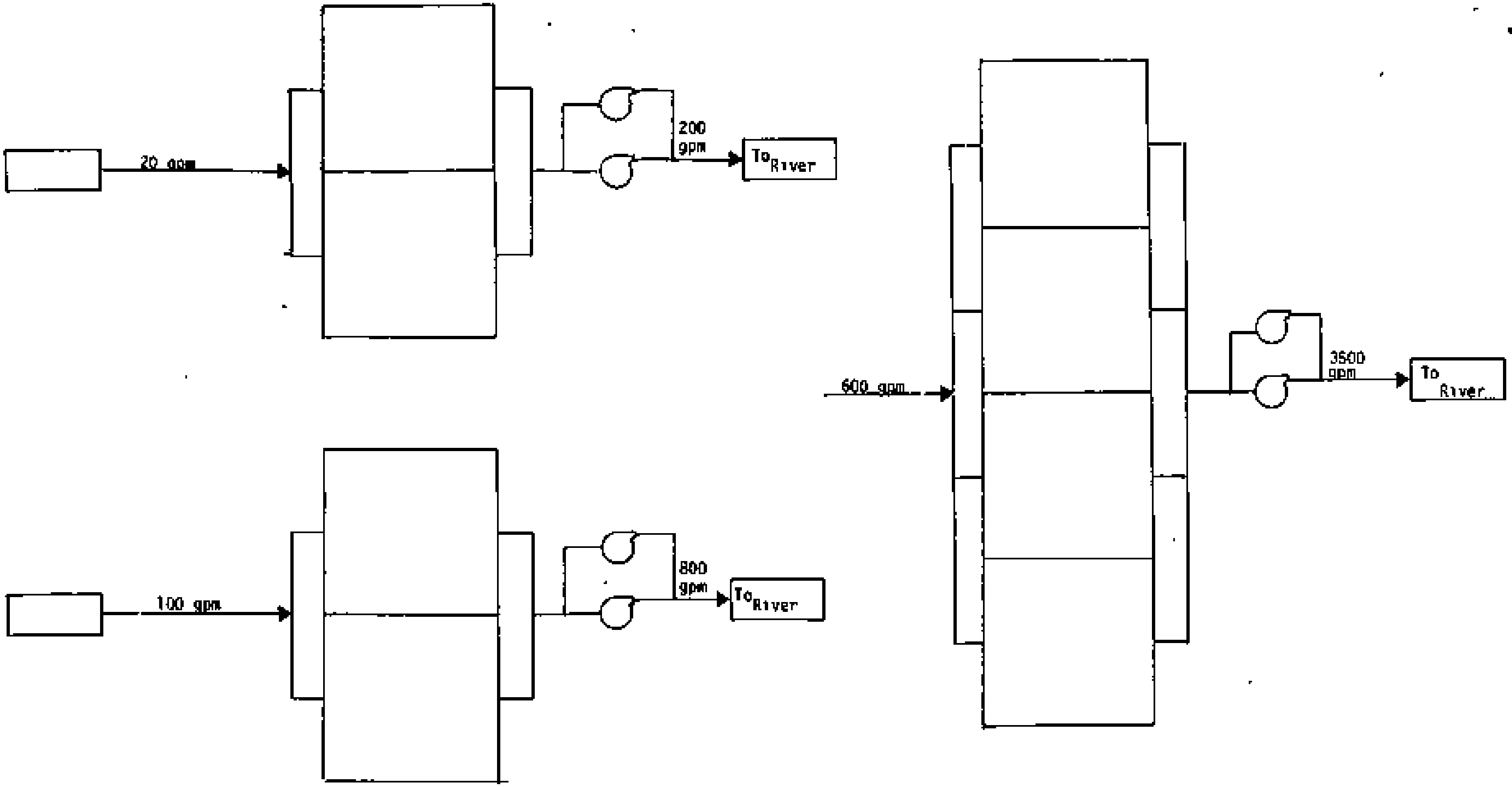

FIGURE 11-C1: DISPOSAL BY STORE AND DISCHARGE

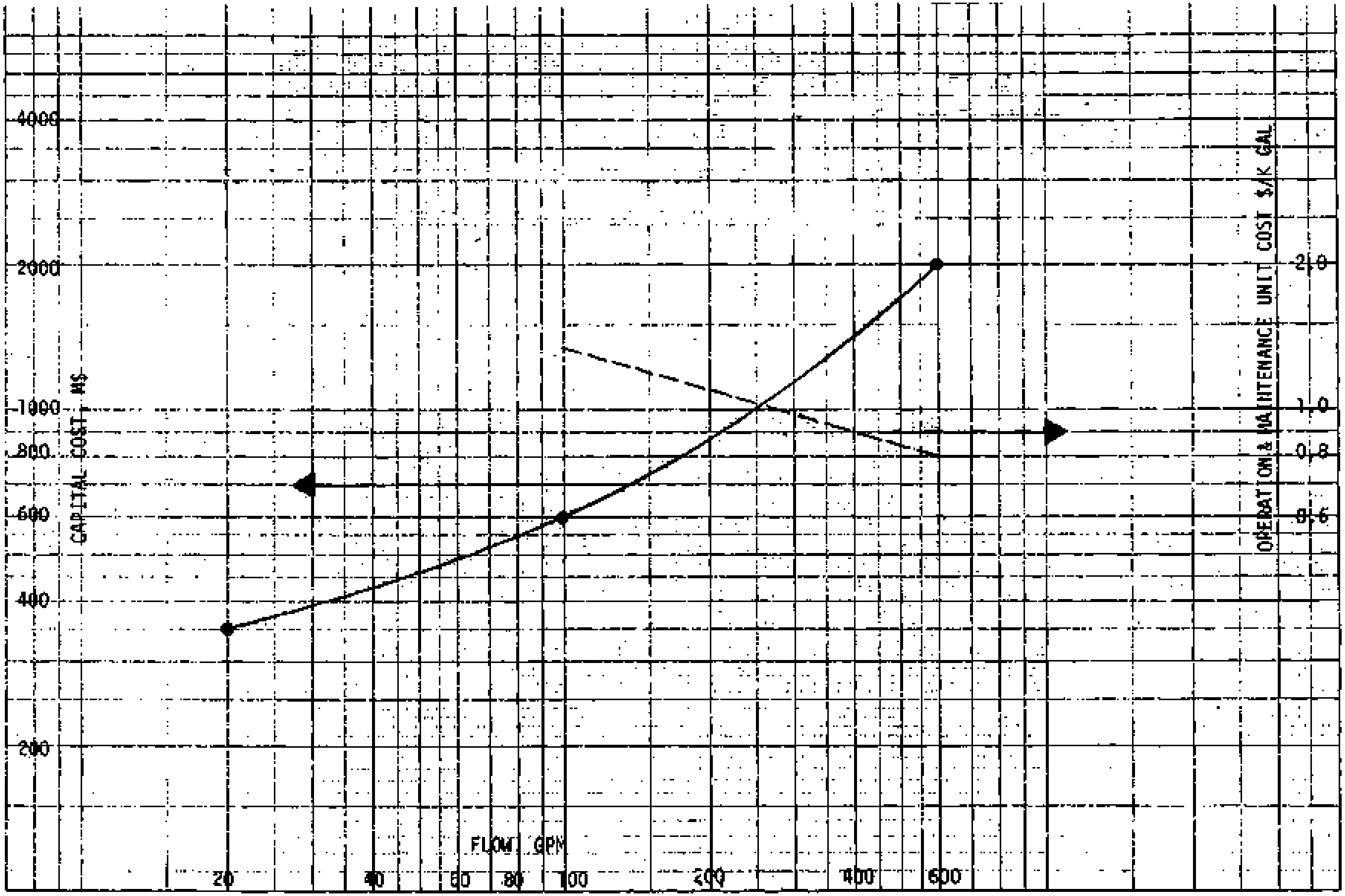

FIGURE 11-C2: CAPITAL COST AND COST OF OPERATION AND MAINTENANCE FOR DISPOSAL BY STORE AND DISCHARGE

## 11.0 APPENDIX D - DISPOSAL BY STORE AND BARGE

A concept was prepared to receive treated effluent from the activated sludge process for disposal by store and barging. The concept illustrated in Figure 11-D1 provided for storage of the waste in lined steel tankage for a period of up to 28 days to accommodate barge scheduling during worst conditions. The barge would transport the waste from the Wheeling site downriver to the open sea for disposal. Although the disposal method is favored by logic, the applicability of the concept would be conditional to regulatory negotiations. The concept would be inapplicable to Carson City.

The disposal of inorganic wastewater residues from coal conversion processes to the sea is a superior concept from the standpoint of logic because the salts are primarily derivatives of the sea. This is not to imply that there would be zero impact from a disposal operation. The contention is that the potential benefits of the gasification operation are substantial compared to the very minimal impact of returning essentially sea derived salts to their place of origin. The concept is substantially superior to prevailing discharges from nearby coke plants that discharge treated effluent to natural watercourses. Thus the features of superiority to prevailing discharges and minimal impact on receiving waters make the store and barge concept a strong candidate for permit negotiation.

For cost estimation purposes, engineering flow sketches were prepared of store and barge concepts for flows of 20,100 and 600 gpm. The concepts included corrosion resistant tankage with high flow pumped discharge and barge docking facilities. The capital cost of the system was estimated and the results are presented as Figure 11-D2. An estimated barging cost of $0.042 per gallon was obtained from an Ohio River barge operator for a 1400 ton load hauled from Wheeling, West Virginia to New Orleans.

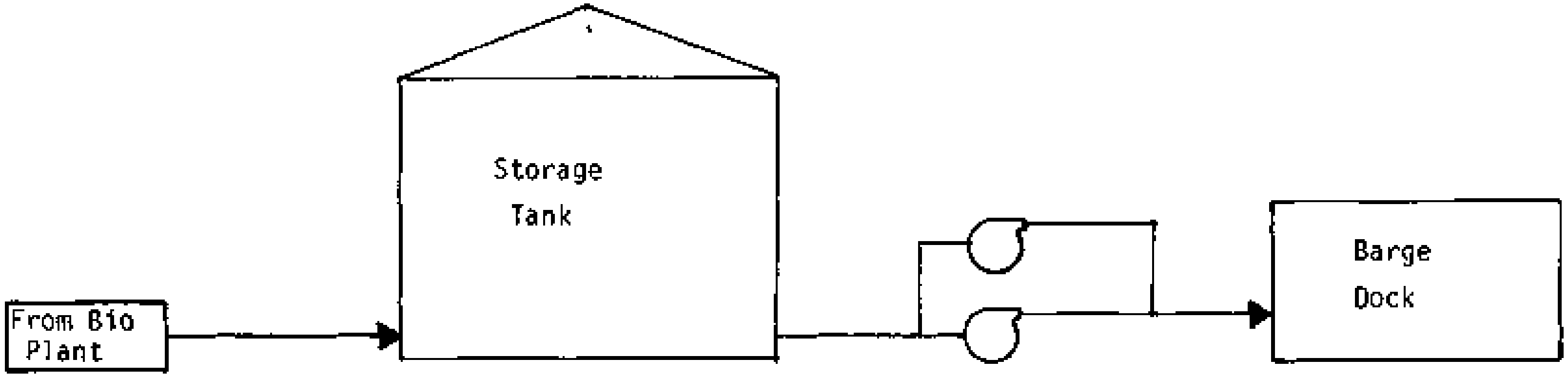

FIGURE 11-D1: DISPOSAL BY STORE AND BARGE

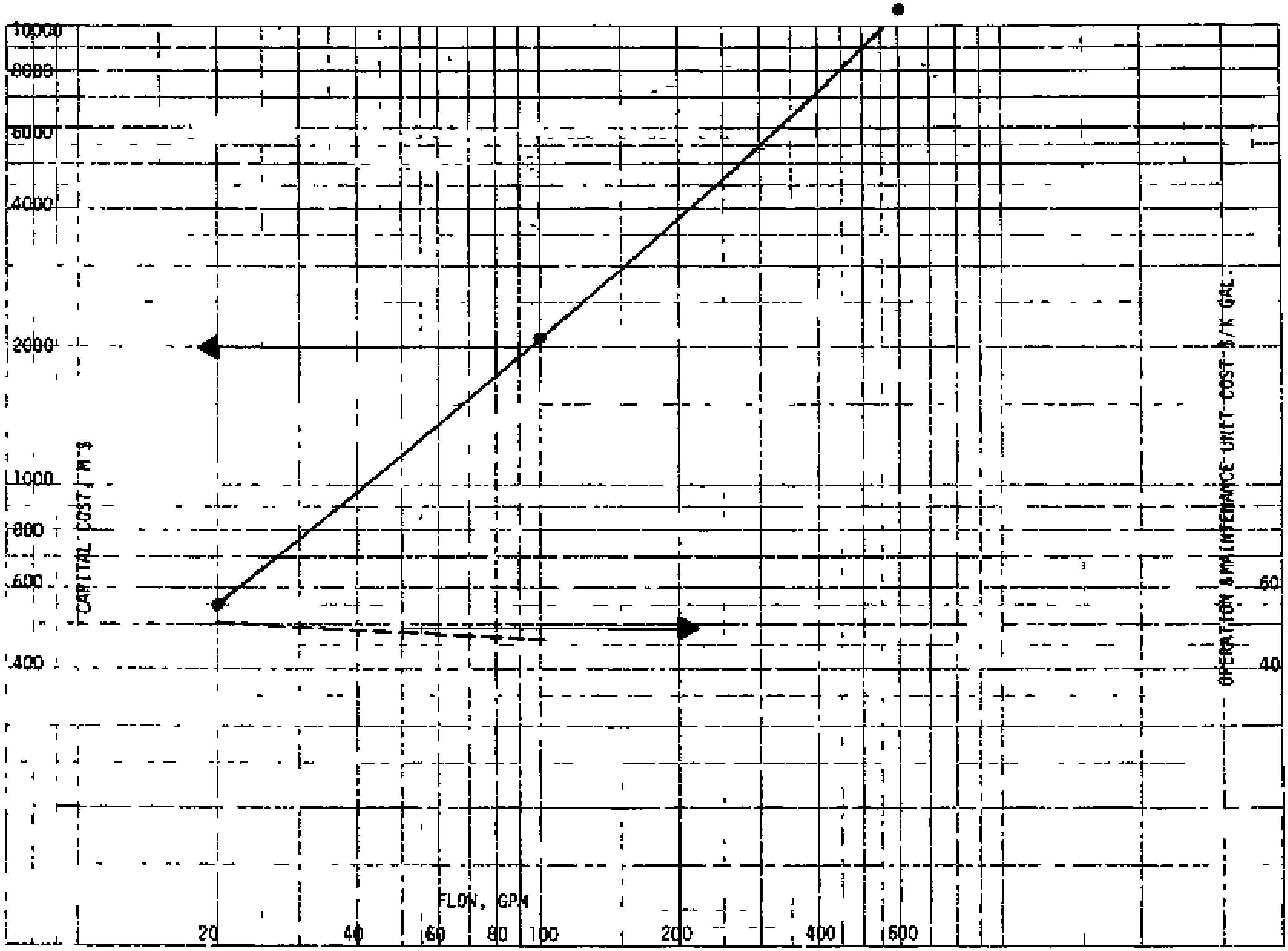

FIGURE 11-D2 CAPITAL COST AND COST OF OPERATION AND MAINTENANCE FOR DISPOSAL BY STORE AND BARGE

## 11.0 APPENDIX E - DISPOSAL BY POND EVAPORATION

A concept was prepared for disposal of treated activated sludge effluent in the Carson City area by pond evaporation. Disposal by pond evaporation is infeasible in the Wheeling area. The Climate Atlas of the U.S. reports for the Carson City area an annual lake evaporation of 40 inches and an annual precipitation of 8.43 inches for a net evaporation of 40 - 8.43 = 31.6 inches per year. The evaporation ponds were lined to restrict percolation and were designed to provide a net evaporation of 1.25 times the inflow plus storage for 5 months. The concept employed for pond evaporation is illustrated in Figure 11-E1.

Pond evaporation is considered to be the preferred method of disposal of inorganic wastewater residue at the Carson City, Nevada site. For cost estimation purposes, engineering flow sketches were prepared for flows of 16, 80 and 480 gpm. The capital cost estimates are presented in Figure 11-E2.

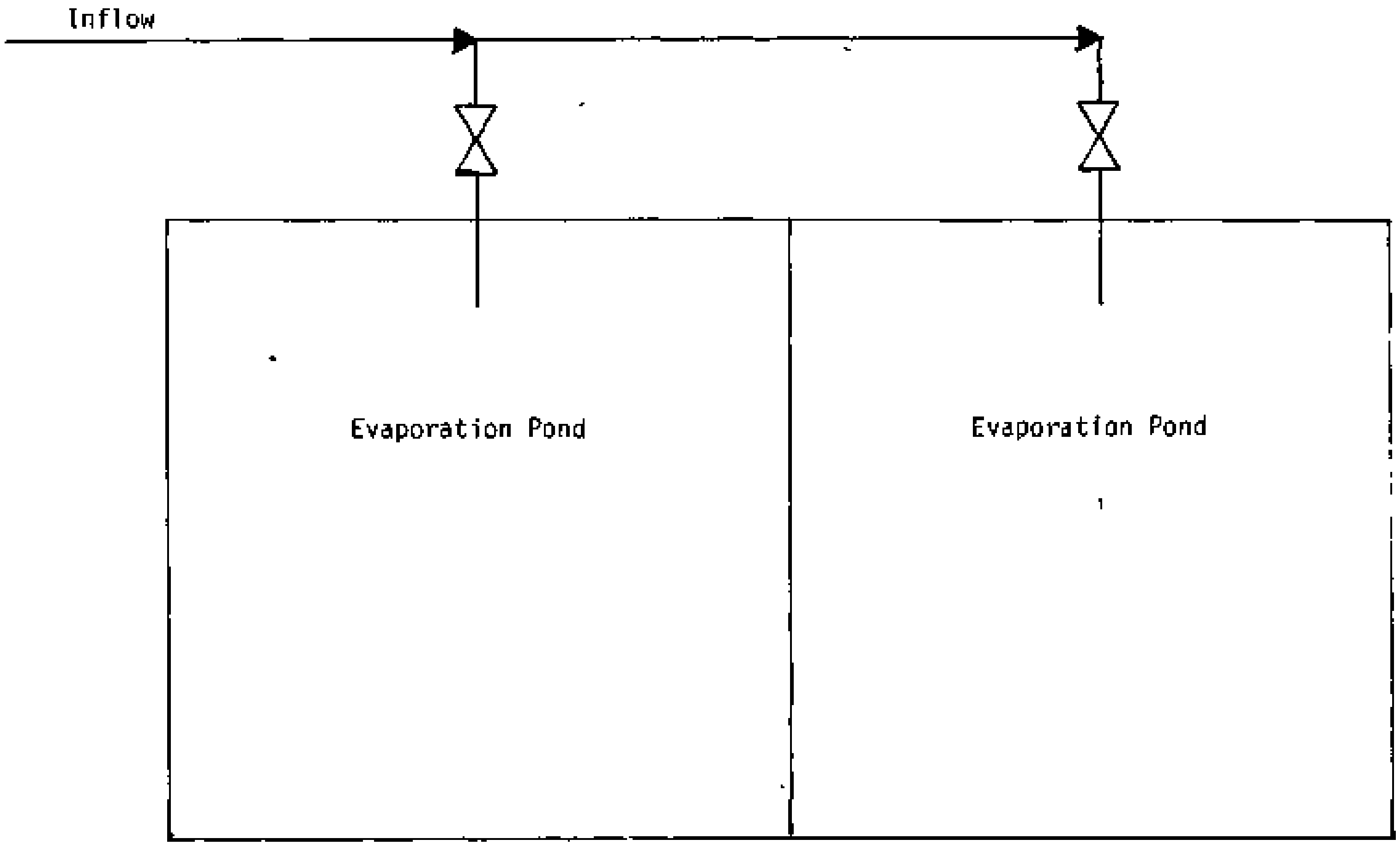

FIGURE 11-E1: DISPOSAL BY POND EVAPORATION (CARSON CITY, NEVADA ONLY)

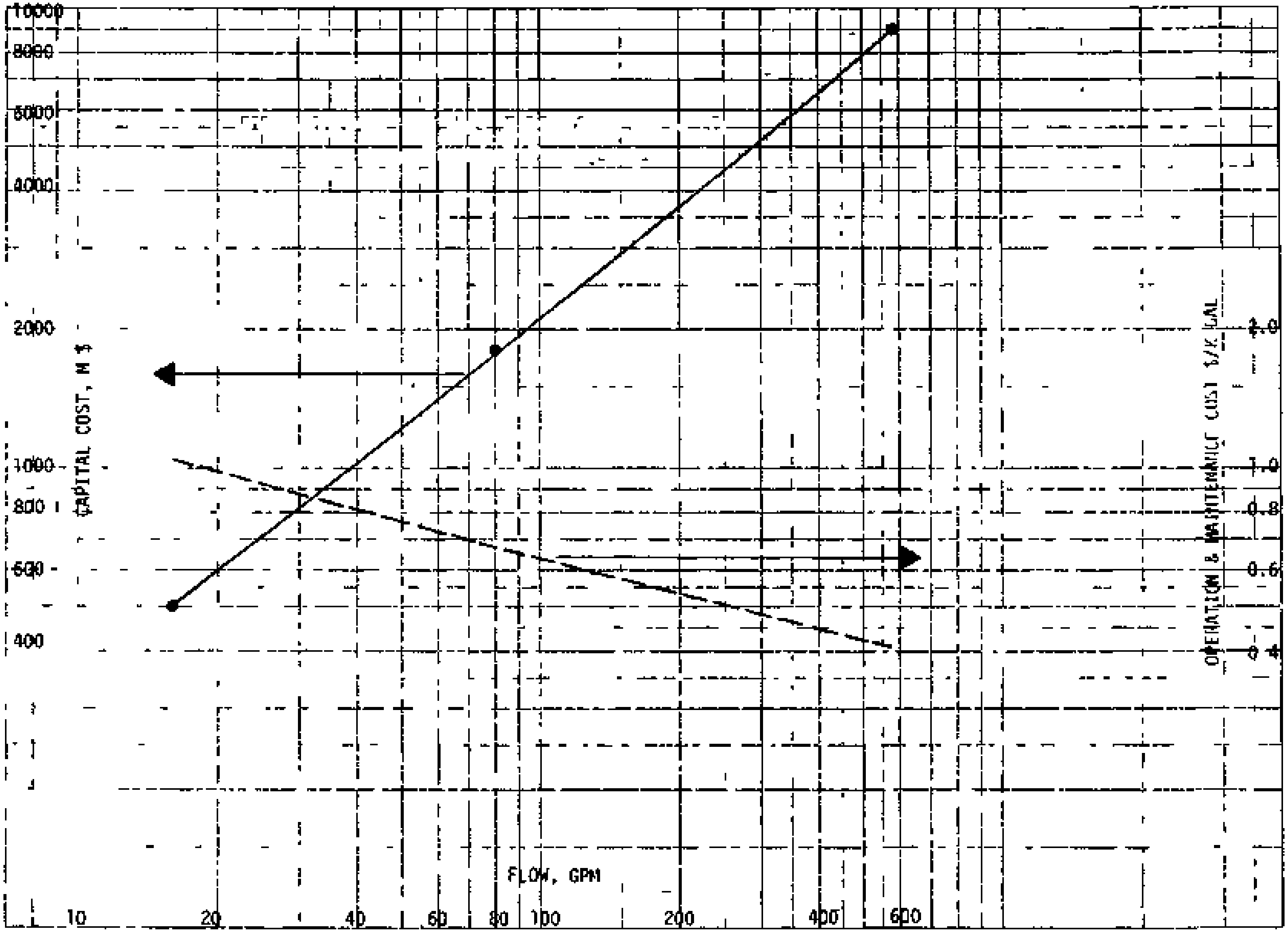

FIGURE 11-E2 CAPITAL COST AND COST OF OPERATION AND MAINTENANCE FOR DISPOSAL BY POND EVAPORATION

## 11.0 APPENDIX F - REVERSE OSMOSIS TREATMENT

A concept was prepared to desalinate treated activated sludge effluent by reverse osmosis as illustrated in Figure 11-F1. The concept included inflow storage, chlorination, sand filtration, cartridge filtration, dechlorination, acidification and membrane filtration. The performance of the system was estimated from correlation with results reported for reverse osmosis plants operating on cooling tower blowdown (1) and on brackish river water (2). The permeate from reverse osmosis was revised for cooling tower makeup and the concentrate was sent to disposal.

Reverse osmosis offers a practical means of recovery of many wastewater flows. The process provides for removal of particulate matter and colloids, as well as for partial removal of ionic or non-ionic dissolved substances. One of the limitations of the process are fouling of the membrane by some organic materials and by precipitates formed as the concentrations increase 3 to 10 fold in the concentrate. Membrane fouling can be alleviated to some extent by adjustment of process conditions such as pH or degree of concentration, or by feed of precipitation suppression agents.

A second limitation of reverse osmosis is that the separation performance differs between substances and between ionic species. Some leeway relative to performance is available by selection of membrane materials for specific applications but leakage of solute components into the permeate is an inherent feature of reverse osmosis processes. The following values were employed in the present study as representative of reverse osmosis leakage: $Na^+$ - 15%, $CL^-$ - 17%, $SO_4^=$ - 1.2% and $NH_4^+$ - 94%.

For purposes of capital cost estimation, engineering flow sketches were prepared for flows of 20,100 and 600 gpm. The capital cost estimates are presented as Figure 11-F2. Estimates of cost of operation and maintenance were developed from values reported in the literature (1)(2).

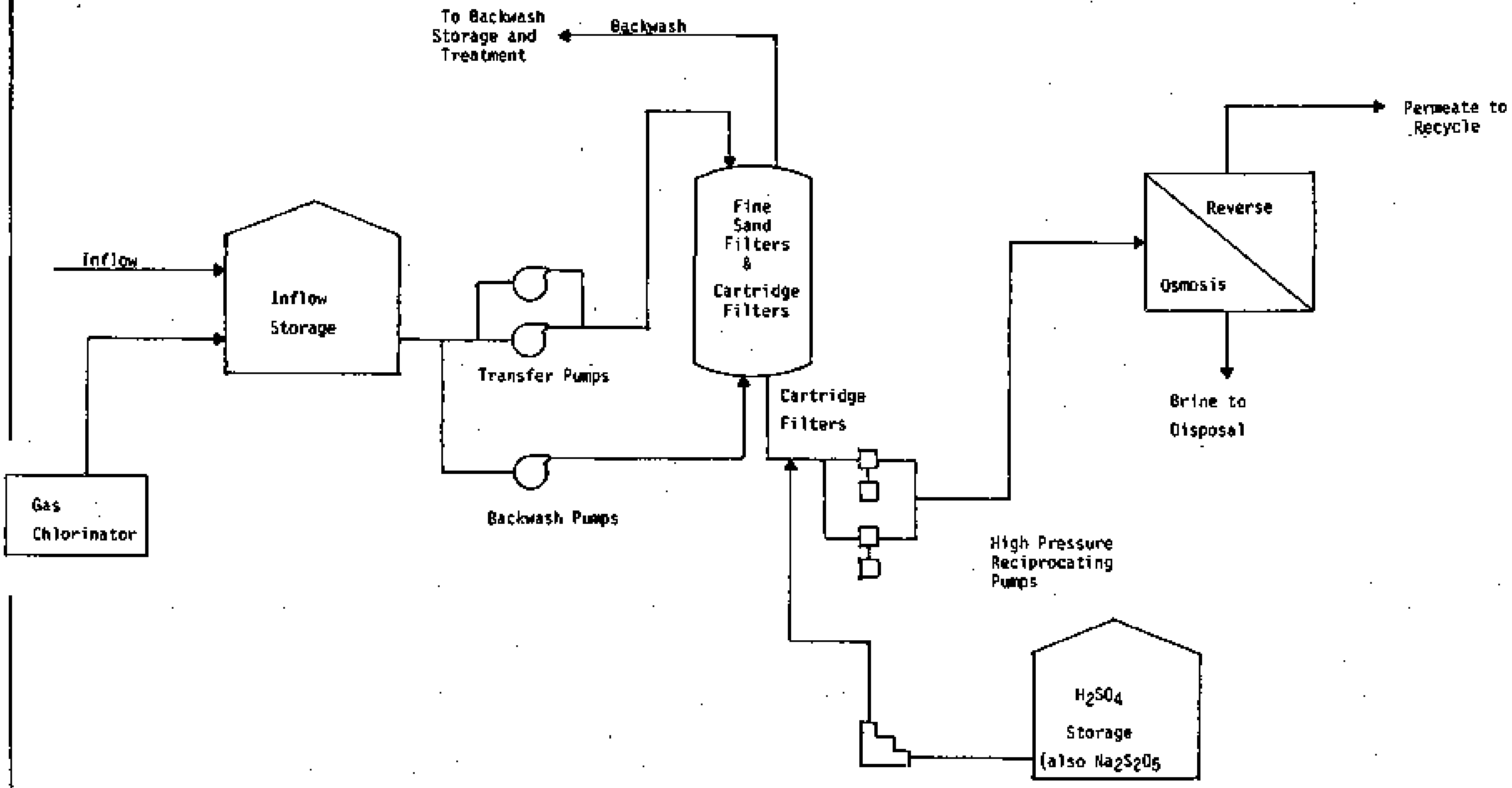

FIGURE 11-F1: REVERSE OSMOSIS TREATMENT MODULE

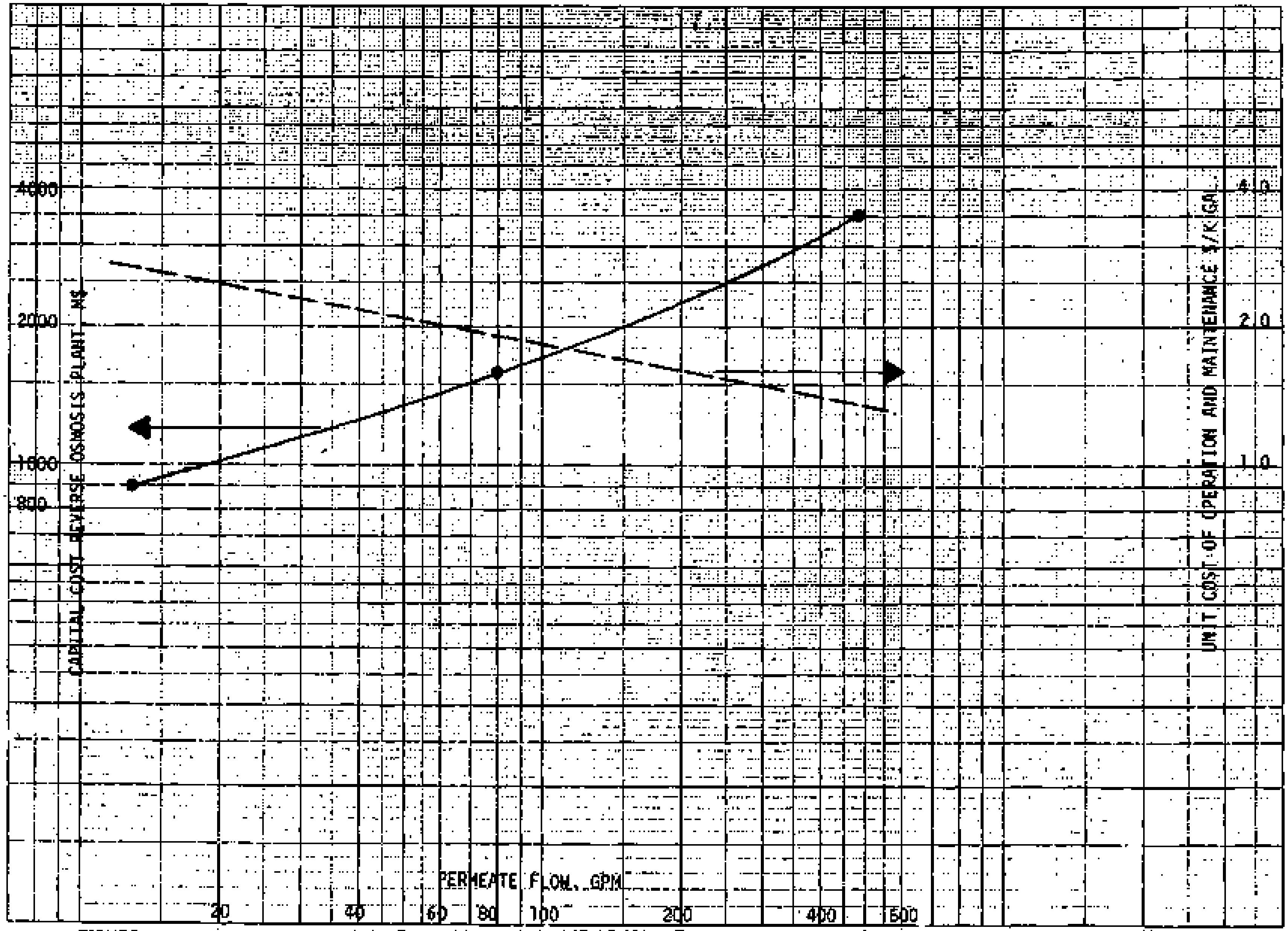

FIGURE 11-F2: ESTIMATED CAPITAL COST AND ESTIMATED COST OF OPERATION AND MAINTENANCE FOR REVERSE OSMOSIS TREATMENT

## Appendix F - References

1. Dascher, R. E. and Lepper, R., Power Vol. 21, No. 8 pg. 23-28 (Aug. 1977).

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## 11.0 APPENDIX G - ECONOMIC ANALYSIS

The following economic analysis compares a single stage and two stage quench system required for pollution control on various coal conversion processes. The quench systems are evaluated as separate entities totally unrelated (financially) to the rest of the facility. However it should be mentioned that certain economic advantages can be obtained as the quench system relates to the entire coal conversion process. Two of these advantages are, a 20% investment tax credit or a 5 year amortization on pollution control equipment. These savings can be applied to revenues derived from the entire facility. Examining the quench systems by themselves prevents the evaluation of these advantages in any economic study since the systems do not provide a positive cash flow but rather a drain on the cash flow of the project. The economic evaluation of these systems is therefore approached in the following manner.

Each specific case as previously defined contains costs for a single stage and double stage quench system. That system which has both the higher capital costs and the higher operating costs is obviously the more expensive process. To quantify this procedure we will use an incremental difference in capital costs and incremental difference in operating costs between the single and double quench processes. The capital cost difference however has to be annualized to be added to yearly operating costs in order to arrive at a total annualized cost. The method and reasoning by which the incremental capital costs are annualized is as follows:

The incremental difference in the capital cost of the two system (single and double quench) can be thought of as an investment which can be put to use at a return of 15 percent (or whatever minimum return on investment (ROI) the company chooses to measure investment potential). The investment is assumed to have the same life as the coal conversion facility, in this case 20 years.

It is desired to determine what amount of yearly payments can be drawn down on the investment (incremental capital cost) such that a rate of 15% the amount left after 20 years is zero. This can be seen mathematically as an annuity defined as follows:

I = incremental capital cost (difference between single and double stage processes)

C = annualized cost

$$I = C\frac{(1 + i)^n - 1}{i(1 + i)^n}$$

i = .15

n = 20 years

Solving for C then yields the annualized cost equivalent of the difference in capital costs of the two systems, and the result can then be added to the yearly operating cost difference to arrive at a quantified difference called the incremental total annualized cost.

Illustration

| | Single Quench | Double Quench | Difference |
|---|---|---|---|
| Capital Cost (MM$) | 19.1 | 17.8 | 1.3 |
| Operating Costs (MM$/yr) | x + 0.32 | x | 0.32 |

Annualized capital cost difference of $1.3 MM

$$C = I \frac{(1 + i)^n - 1}{i (1 + i)^n}$$

i = .15

n = 20

I = $1.3 MM

$$\frac{(1 + i)^n - 1}{i (1 + i)^n} = \text{Present value factor} = 6.25$$

$$C = \frac{1.3}{6.25} = \$.20 \text{ MM}$$

Total annualized incremental cost + .32 + .20 = $.52 MM

Thus the double quench system in the illustration is obviously the more economical process. Going to a single quench would cost an additional $520,000/yr over the life of the project.

## 11.0 APPENDIX H - THERMAL EVAPORATION

Thermal evaporation of inorganic wastewater residues has particular application where steam is available at low cost or where a high purity desalination product is desired. In the usual situation reverse osmosis is more economical than thermal evaporation for desalination of residues with a total strong electrolyte content of less than one percent - above which the risk of membrane fouling from precipitation increases. It follows that thermal evaporation of reverse osmosis concentrate is a viable concept. Two thermal evaporation concepts were prepared. One concept covered situations where reverse osmosis concentrate was to be further concentrated to reduce costs of disposal by barging to sea or to evaporation ponds. The second thermal evaporation concept pertained to further concentration of evaporator bottoms to obtain a feed for a wastewater incinerator that produced a dry product residue.

The concept for concentration of reverse osmosis concentrate employed a high efficiency 14 effect vertical tube evaporator. The evaporator was fabricated from corrosion resistant materials (e.g. titanium) and had the capability of concentrating a feed with one percent solids to a concentrate of ten percent solids. The system would yield 10 gallons of condensate per pound of steam. The estimated capital cost of the evaporator facility is given in Figure 11-H1.

The second thermal evaporation concept consisted of a pan evaporator coupled to a wastewater incinerator. The pan evaporator accepts the bottoms from the vertical tube evaporator and further concentrated them to 60 percent solids. The estimated capital cost of the evaporator-incineration facility is given in Figure 11-H2.

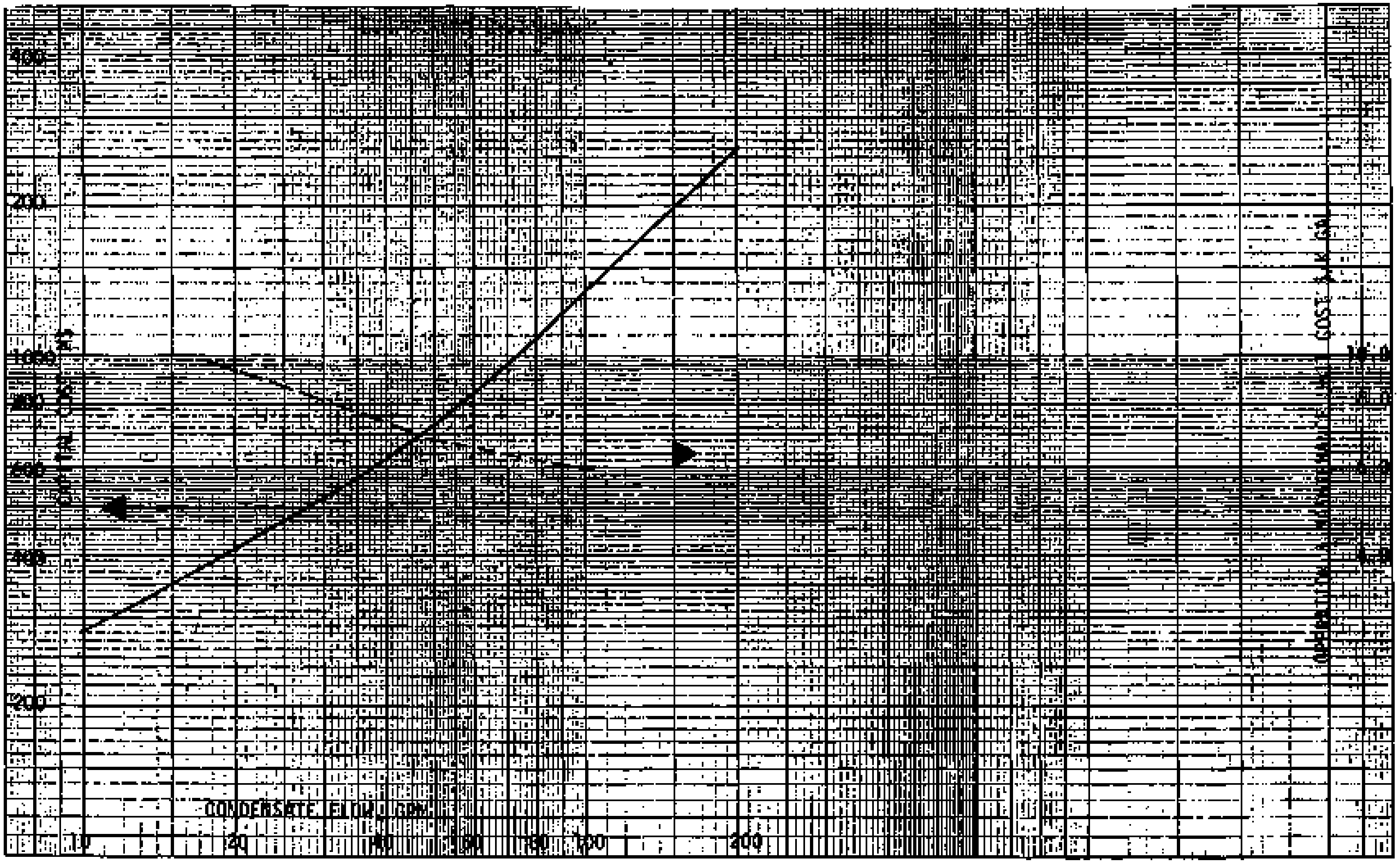

FIGURE 11-H1: CAPITAL COST AND COST OF OPERATION AND MAINTENANCE FOR HIGH EFFICIENCY EVAPORATION

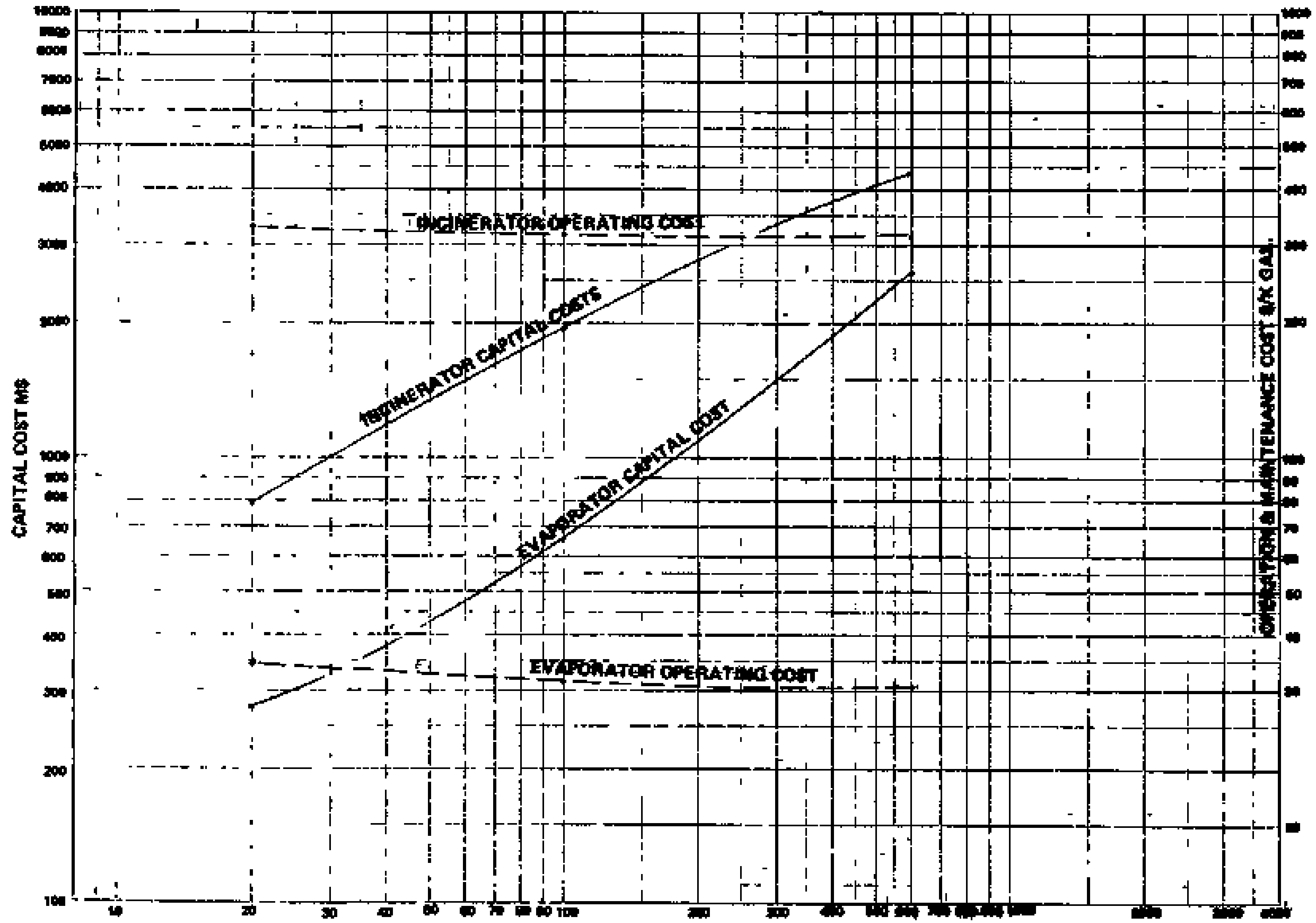

FIGURE 11-H2 CAPITAL COST, MAINTENANCE COST AND OPERATING COST FOR THERMAL EVAPORATION AND INCINERATION

## 11.0 APPENDIX I - MATERIALS OF CONSTRUCTION

No attempt was made to optimize the material selections for the gas cleaning of wastewater stripping equipment. The equipment is large, and obvious economic incentives exist to use the least expensive alloys that will perform in a given environment. The environment in the first stage scrubber features high halides and low pH in a reducing atmosphere. The reducing atmosphere represents a departure from flue gas scrubber installations. The low pH represents a departure from coke plant experience. Thus limited transfer technology is available to cover the situation.

It is difficult to specify the optimum materials without previous operating experience. Many of the corrosion/erosion problems that occur in a scrubber cannot be predicted in the design stage. The use of corrosion test spools in environments closely approximating the gas cleaning system or wastewater strippers would be very beneficial in making an economic choice of materials.

Literature does exist concerning corrosion in aqueous systems by dissolved gases (1, 2, 3, 4, 5, 6). Discussions of materials challenges associated with coal conversion quench systems have been published by Bhattacharya, et.al. (5) and by O'Hara, et.al. (6). Test data on wet scrubbers for the incinerator applications and power plant flue gas scrubbing give some indication towards the proper choice of materials.

Carbon steels, low alloy steels, and type 304 L stainless steel do not generally possess adequate corrosion resistance in wet scrubbing environments. Some scrubber systems use carbon steel with a protective coating or lining.

Coatings or linings will not always perform well. They are particularly vulnerable to temperature excursions in the system, and they may not adhere to the surface if improperly applied.

The material of construction chosen for many scrubber installations has been type 316 L stainless steel.

Type 316 L stainless steel is sometimes subjected to localized corrosive attach in scrubber environments with certain combinations of pH and chloride content. Localized corrosion increases as the chloride level increases and the pH of the scrubbing level decreases. Increasing the chromium and molybdenum contents of the high-nickel alloys decreases this localized attack (2, 3, 4).

The following comments are based on pH and chloride content alone and do not necessarily consider the other corrosive components in the wastewater.

- Severe corrosion may be experienced with 316 L or 317 L stainless steel from the first stage quench wastewater due to high chlorides and low pH. Alloys high in chromium, molybdenum, and nickel such as Inconel 625, Hastelloy C-276, or Titanium may be required for this service.
- Based on pH and chloride content of the single stage quench wastewater, the performance of stainless steel would be questionable.
- Stainless steel or even carbon steel may be acceptable for the second stage quench wastewater because of the nearly neutral pH and absence of chlorides.

Literature is available on corrosion in strippers (6, 7, 8). The top portion of the stripper will be exposed to the most corrosive environment where the use of a stainless steel may be warranted over a carbon steel. This is especially true of the overhead condenser where the use of a high alloy steel or Titanium may be justified.

Carbon steel was chosen as the material of construction for all equipment for the purpose of the cost estimation. A multiplying factor may be used to compare the cost of other materials.

## Appendix I - References

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2. Hoxie, E. C., and Tuffnel, G. W., A summary of INCO Corrosion Tests in Power Plant Flue Gas Scrubbing Processes, "Resolving Corrosion Problems in Air Pollution Control Equipment," sponsored by APCA, IGCI, and NACE, 1976.

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5. Bhattacharya, S., Bock, F., MacNab, A., Cox, T. B., Alloy Selection for Coal Gasification Quench Systems, Paper No. 51, presented at the International Corrosion Forum, March 14-18, 1977.

6. O'Hara, J. B., Lochmann, W. J. and Jentz, N. E., Materials Challenges of Coal Liquefaction, "Chemical Engineering," pp. 147-154 (April 11, 1977).

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8. 1972 Survey of Materials Experience and Corrosion Problems in Sour Water Strippers, API Publication 944.