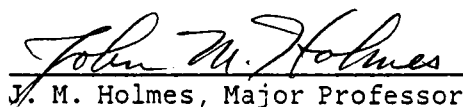
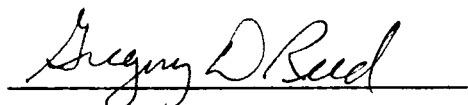


To the Graduate Council:

I am submitting herewith a thesis written by Dale R. Heron entitled "Computer-Aided Design of a Combined Phenosolvan--Wet Air Oxidation Treatment Process to Treat Contaminated Wastewater from a Coal Conversion Facility Using ASPEN." I have examined the final copy of this thesis for form and content and recommend that it be accepted in partial fulfillment of the requirements for the degree of Master of Science, with a major in Chemical Engineering.


J. M. Holmes, Major Professor

We have read this thesis
and recommend its acceptance:





Accepted for the Council:


The Graduate School

COMPUTER-AIDED DESIGN OF A COMBINED PHENOSOLVAN--WET AIR
OXIDATION TREATMENT PROCESS TO TREAT CONTAMINATED
WASTEWATER FROM A COAL CONVERSION FACILITY
USING ASPEN

A Thesis
Presented for the
Master of Science
Degree
The University of Tennessee, Knoxville

Dale R. Heron
March 1985

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I hereby acknowledge with thanks the permission granted by Dr. Tipton L. Randall to reproduce Tables 1 and 2 from his paper entitled "Wet Oxidation of Toxic and Hazardous Compounds" and to reproduce Tables I and IX from Patrick J. Canney's paper entitled "Detoxification of Hazardous Industrial Wastewaters by Wet Air Oxidation". I also express

my thanks to Dr. D. Bhattacharyya for his permission to use Tables 1 and 2 from his article in Environmental Progress entitled "Coal Gasification Process Wastewater Reusability: Separation of Organics by Membranes".

ABSTRACT

The treatment of contaminated wastewater from a synthetic-fuel plant using coal as a feedstock represents a significant problem in pollution control. This is especially true for the removal of phenol due to the solubility of phenol in water.

Many different treatment processes are available for wastewater treatment such as biological oxidation, carbon adsorption, and Wet Air Oxidation, (WAO), but, since very few synthetic-fuel facilities have been constructed, it is not clear how these processes will perform with wastewaters from these facilities. The primary objective of this study was to evaluate the viability of one possible treatment system involving the combined processes of a proprietary solvent extraction system called Phenosolvan and Wet Air Oxidation to treat phenolic wastewaters from a synthetic-fuel plant processing 20,000 ton/day of coal. This combination offers the advantages of recovering phenol as a by-product which can be sold or burned and effective treatment for residual concentrations of many different organic contaminants. The secondary objective of this study was to evaluate the effectiveness of the Advance System for Process ENgineering, (ASPEN), program in performing a mass and energy balance simulation and preparing a capital cost estimate for the combined treatment process.

The results from this study indicate that the combined process reduced concentrations of phenol in the wastewater stream from 7000 ppm in the aqueous feed stream down to 0.25 ppm in the discharge stream from the system. The removal of phenol provided a Chemical Oxygen Demand, (COD), reduction from 16,300 ppm to 0.72 ppm. The discharge value for

phenol, however, is still above the expected pollution standard of 0.13 ppm indicating that additional post-treatment is needed. From the economic evaluation of the process, it was found that a capital investment of \$126 MM, (1984), is needed to construct the process. In addition, it was found that using this process to treat wastewater from a 20,000 ton/day of coal synthetic natural gas facility resulted in a \$0.28/MMBTU increase in the gas price or a cost of \$20.52/MGAL of treated wastewater.

From the study, it was concluded that the combined process could treat phenolic wastewater to a significant degree at a reasonable cost increase to the product of only 3.4%. It was also concluded that ASPEN is a very valuable tool in the evaluation of chemical processes.

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I. INTRODUCTION

Since the oil crisis in 1974, the United States has been trying to achieve greater energy independence. Coal, the most abundant fossil fuel source in the United States, has played a major role in this movement.¹⁰ Most of the emphasis for using coal has been geared towards the production of gaseous and liquid synthetic-fuel products. Several major synthetic-fuels projects are currently planned including the following: Tennessee Eastman in Kingsport, Tennessee; TVA in Muscle Shoals, Alabama; and American Natural Resources, Tenneco SNG and others for the Great Plains Coal Gasification Project in Beulah, North Dakota.^{11,47}

Facilities of this nature present many engineering problems which are new and potentially difficult to resolve. One major concern involves the release of contaminated wastes into the environment. Because water is a significant component in most U.S. coals and is also a constituent in many of the reactions which take place in the coal conversion process, its treatment as a plant effluent or as a recycle stream must be an integral part of any synthetic-fuel plant. This water, however, contains a wide range of dissolved organic and inorganic contaminants in low concentrations which are very difficult to remove. Many of these compounds are not only toxic to both plant and animal life but also carcinogenic to humans as well.²⁶

Because of the high toxicity of these wastewaters, conventional biological treatment is not acceptable unless substantial dilution is employed. Even with this dilution, however, biological processing can be economically unfeasible because of the slow bioflocculation times of the microorganisms involved and thus major capital investments will be

required for land and equipment.⁴² For this reason, some form of pretreatment must exist in order to lower the toxic compounds to a level at which conventional biological oxidation can take place economically.

Many processes to perform pretreatment of wastewaters similar to those from coal conversion facilities have been proposed based upon experience gained from coking and petroleum refining operations. However, it is unclear how these schemes will perform with actual wastewaters from a synthetic-fuels plant. Evaluation of these process options can be performed in many different ways. They range from in-depth plant studies to quick estimation calculations. One method which has gained much popularity recently is computer aided simulation. This method offers the advantages of reliability, consistency, ease in operating parameter manipulation and low costs. The most recent and one of the most versatile of the many simulation programs on the market is the Advanced System for Process ENgineering, (ASPEN), program developed by the Massachusetts Institute of Technology, (MIT), for the U.S. Department of Energy specifically for the modeling of synthetic-fuel plants.¹⁷

II. BACKGROUND PRESENTATION

Wastewater Characterization

Before the various processing methods can be investigated, the major constituents of the contaminated wastewater must be identified. The wastewater from a synthetic-fuels facility consists primarily of sour liquor condensed from the coal gasifier effluent. During the gasification process, a large number of products are produced by the complex reactions which take place. All of these products are present to some extent in the wastewater produced from the process, however, only those components which are highly water soluble are present in significant concentrations. Of these chemicals, ammonia, cyanide, thiocyanate, sulphur and phenols* represent the major contaminants. Table 1 presents data pertaining to some of the major contaminants present in gasifier effluents. Table 2 presents additional data on the overall wastewater characteristics.⁷

The concentrations of the major contaminants in the wastewater can be effected by several factors. These factors include coal type, gasifier type and operating conditions. The data presented in Tables 1 and 2 were taken from low temperature fixed and fluidized bed gasifiers such as Lurgi, Synthane and Hygas. As is indicated, phenols make up a large percentage of the organic contaminants present in the wastewater. This is not the case for high temperature entrained-bed gasifiers such as

*The word phenols in this text may include aromatic organic compounds with 6,7 and 8 carbon atoms with one or more hydroxyl groups attached.

Table 1

Typical Organic Compounds in Coal Conversion Wastewaters

Organic Compounds	Synthane PDU Condensates mg/liter	Lurgi Wastewaters mg/liter
Phenol	420-4480	1200-5650
Cresols	530-3580	483-1965
C ₂ -Phenols	140-2185	100-453
C ₃ -Phenols	20-365	40
Dihydric phenols	20-540	546-1751
Naphthols	30-290	NR ^a
Indenols	20-990	NR
Indanols	40-384	NR
Biphenols and biphenols	0-360	19
Pyridines	15-580	40-222
Anilines	41-49	12
Quinolines	0-100	NR
Indoles	0-100	NR
Carbozoles	NR	9
Acetic acid	620	171
Benzoic acid	10-110	NR

^aNR - not reported

Source: Bhattacharyya, D. R., R. I. Kermode, and R. L. Dickinson, "Coal Gasification Process Wastewater Resuability: Separation of Organics by Membranes," Environmental Progress, Vol. 2, No. 1, Feb., 1983, pp. 38-46.

Table 2

Typical Wastewater Characteristics of a
Comercial Scale Gasification Facility

Component	Concentration mg/liter (except pH)	Effluent Production Rate ^a kg/10 ³ kg Coal MAF
pH	7.8 - 9.3	
COD ^b	4050 - 43,000	56 - 184
TOC ^c	4190 - 9090	9.2 - 109
Phenols	710 - 6600	2.6 - 40.2
Tars	1150-1180 (Lurgi only)	Negligible-234
Oils	840-910 (Lurgi only)	Moderate - 130
Ammonia	4480 - 11,200	2.5 - 45
Sulfide	138 - 400	0.2 - 15.2
Thiocyanate	22 - 200	0 - 14.2
Cyanide	0.07 - 0.6	0 - 0.12

^aCoal Feed rate 260 kg coal/s.

^bCOD is the chemical oxygen demand.

^cTOC is the total organic carbon.

Source: Bhattacharyya, D. R., R. I. Kermode, and R. L. Dickinson,
"Coal Gasification Process Wastewater Resuability: Separation of
Organics by Membranes," Environmental Progress, Vol. 2, No. 1, Feb.,
1983, pp. 38-46.

Koppers-Totzek, Bigas and Texaco. Phenols, tars and oils appear in negligible concentrations in the wastewater from these gasifiers. In all of these gasifier types, the operating conditions and coal characteristics significantly effect both the type of organic contaminants and concentrations of organics in the wastewater. It has been found that by proper control of the major operating parameters of the gasifier, reductions in the pollutant levels can be achieved.⁷

Of the contaminants in the wastewater, phenols are particularly important from a design standpoint. Phenols exist in both the monohydric, (i.e. phenol, cresol, xlenol and ethyl phenols), and the polyhydric, (i.e. catechol and resorcinol), forms. Typically, the monohydric forms comprise 75 to 85 % of the phenols in the waste stream.⁴⁸ They are in concentrations sufficient enough that recovery could be economically feasible, however, removal down to acceptable levels for discharge to the environment is very difficult. This is due in part to the chemical nature of phenols in aqueous solutions.

When phenols are dissolved in water, the hydrogen on the OH^- group tends to dissociate into H_3O^+ and the phenolic cation, (phenoxide). The phenoxide ion, $\text{C}_6\text{H}_5\text{O}^-$, is electrically more stable than phenol due to delocalization of the negative charge throughout the benzene ring. The resulting ionic molecules are readily solvated by water which is highly polar. The degree of delocalization, and therefore the solubility, is a function of the groups attached to the phenol. As in most acid-base equilibrium reactions, pH can effect the extent of dissociation.⁴⁹

Processing Methods

A wide variety of methods have been proposed to treat wastewaters of the kind that would come from a synthetic-fuels plant. Figure 1 is a summary of some of the major processing schemes proposed to date.⁴⁸ Included among these are the Phenosolvan and Wet Air Oxidation, (WAO), processes.

Phenosolvan

The Phenosolvan process is a proprietary process developed by Lurgi Company in Germany around 1940 to remove phenols from coke-oven liquors. The process is designed to recover phenols through a combination of solvent extraction and solvent recovery systems shown in the simplified flow diagram in Figure 2.⁴⁸

In the basic process, the wastewater is passed through a gravel-bed filter to remove any suspended solids. The filtered water then enters a series of mixer-settler extractors where it is repeatedly contacted countercurrently with an appropriate solvent. The original solvent used in this process was isopropyl ether.⁴⁸ This was later replaced by a proprietary solvent called Phenisol which is a mixture of butyl, propyl, amyl and hexyl acetates.³² The solvent dissolved in the raffinate phase is stripped from the water using direct steam, leaving a dephenolized waste which requires final polishing to remove residual organics and other contaminants down to acceptable levels. The stripped solvent is further processed to separate the solvent from the remaining water and finally returned as fresh solvent.⁴⁸

F-1	V-4	V-1	V-5	V-6	V-2	V-3
GRAVEL BED FILTER	GAS SCRUBBER (phenol recovery)	MIXER-SETTLER VESSEL (solvent extractor)	GAS SCRUBBER (solvent recovery)	RAFFINATE STRIPPER (solvent recovery)	SOLVENT DISTILLATION COLUMN	CRUDE PHENOL STRIPPER

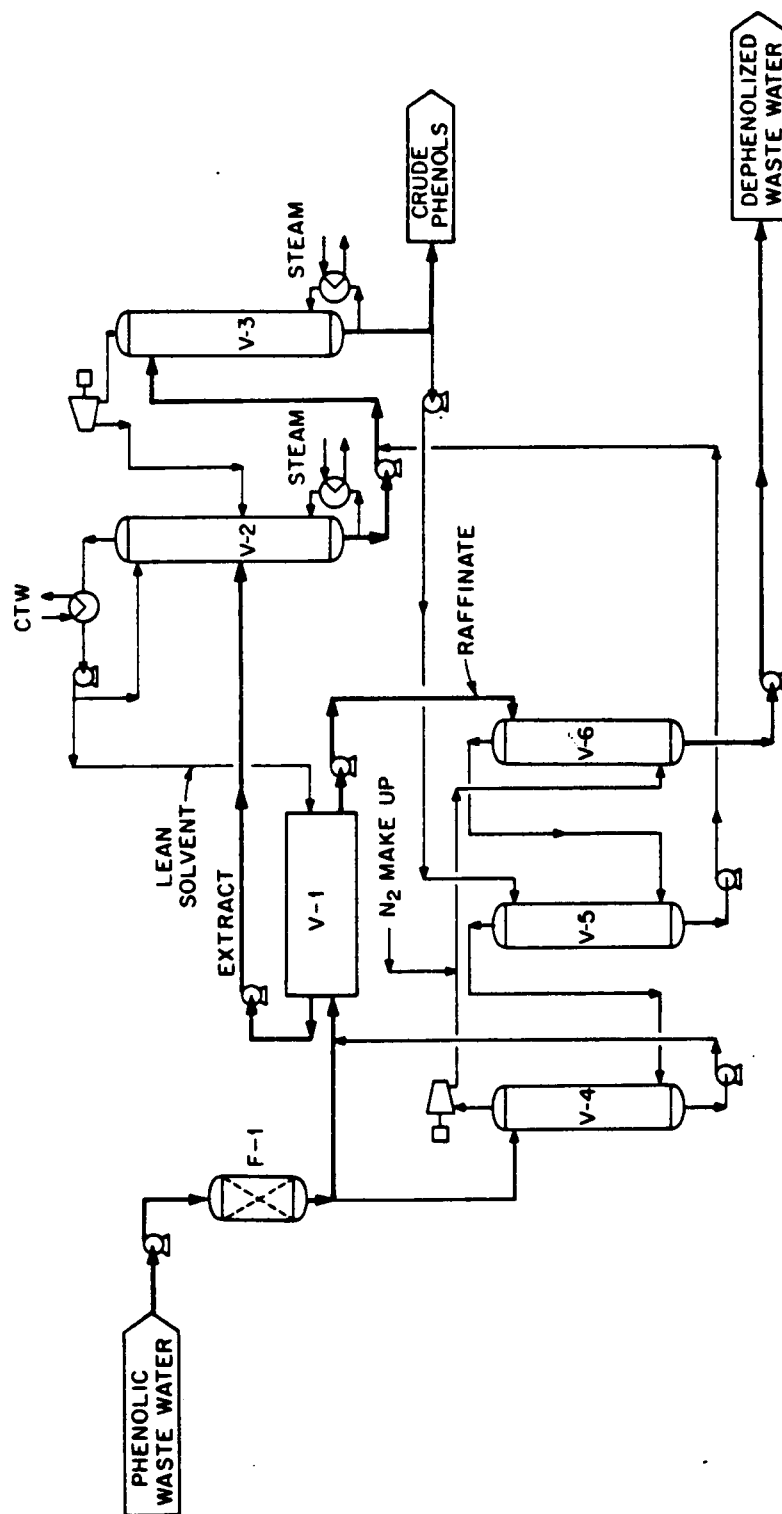


Figure 2. Simplified Flow Diagram of the Lurgi Phenosolvan Process

Source: Singh, S. P. N., et al., "Costs and Technical Characteristics of Environmental Control Processes for Low-BTU Coal Gasification Plants," ORNL Report, No. 5425, June 1980.

The phenol-laden extract is processed through a recovery column where the phenols are fractionally separated from the solvent. The more volatile solvent is condensed, cooled and returned to the mixer-settler vessels as fresh solvent. The phenol rich stream from this column is further stripped to remove any residual solvent. This residual solvent is used as reflux for the first stripper while the crude phenols which are recovered in the tower bottoms are sold or burned.⁴⁸

The Phenosolvan process and other extraction processes similar in nature have been used for phenol removal from contaminated waters in coking and petroleum refining operations for many years. In 1952, Carbide and Carbon Chemicals Company in Institute, West Virginia began operation of a phenol recovery system for their coal-hydrogenation plant.¹⁵ The process incorporates a packed extraction tower using isopropyl ether as the solvent to remove phenols from the wastewater. The solvent is recovered in the same fashion described for the Phenosolvan process and the extracted phenols are mixed with other plant oil streams and sold. The water rate to the Carbide extraction system was 120 GAL/HR containing 1.25 % wt phenols. The water raffinate of 122 GAL/HR contained only 0.095 % wt of phenols indicating a 92 % removal efficiency.¹⁵

In 1961, the Pittsburgh works constructed a solvent recovery system based on the Jones and Laughlin Steel Corporation design. The process is essentially the same as that for the Phenosolvan process except that the extractor is a countercurrent tower and a different solvent is used. The ammonia liquor feed rate to this system was 200 GPM and contained 2000

ppm of phenols. The dephenolized waste stream contained 1-4 ppm resulting in a removal efficiency in excess of 99%.²⁵

For the Phenosolvan process, removal efficiencies have been estimated to be; 99.5% for monohydric phenols; 60% for polyhydric phenols and 15% for other organic contaminants. The effluent water is claimed to contain 5 to 20 ppm of phenols while the crude phenol by-product is said to contain 85 to 95 % phenols.⁴⁸

Wet Air Oxidation

Wet air oxidation is a process developed by the Zimpro Corporation for the removal of dilute concentrations of organic compounds from aqueous solutions.⁴⁸ A simplified flowsheet of the principal components in the process is shown in Figure 3.

In this process, the contaminated water is first delivered to a storage tank for temporary feed holdup. The water is then pumped from this tank to increase the stream pressure, (300-3000 psi), and then it is mixed with air or oxygen. The stream temperature is increased to meet reactor conditions, (347-608 °F), by contacting it countercurrently with the oxidized reactor effluent through a series of heat exchangers.⁹ This conditioned stream is feed into the oxidation reactor where it is held for residence times ranging from 0.5 to 2 hours depending on which organics are being oxidized.³⁶

The organic materials in the wastewater are oxidized primarily to CO₂, CO, NH₃, H₂O and their respective inorganic oxides. Along with these gaseous products, low molecular weight acids such as acetic, formic and succinic acid are formed with lesser amounts of low molecular weight aldehydes as is shown in Figure 4.⁴ The high temperature and pressure of

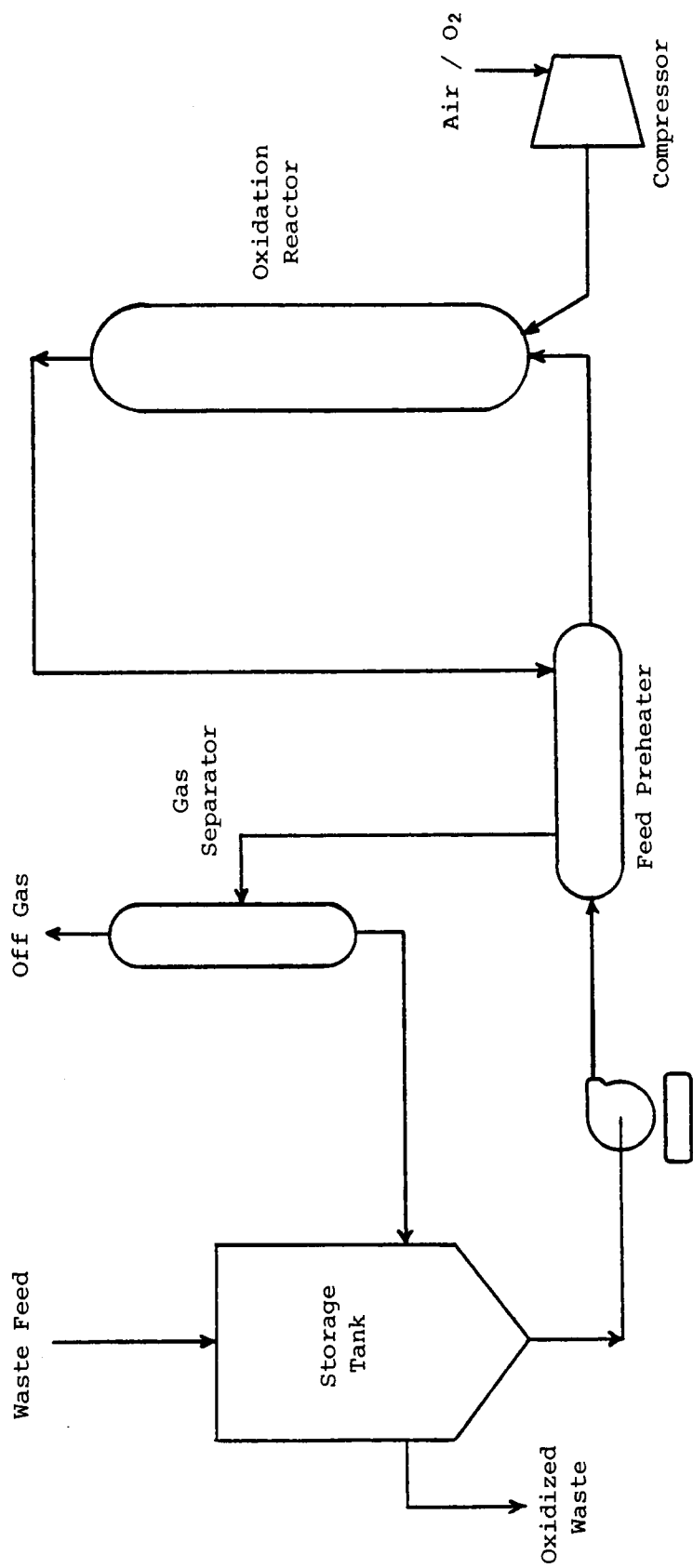


Figure 3. Simplified Wet Air Oxidation Flow Diagram

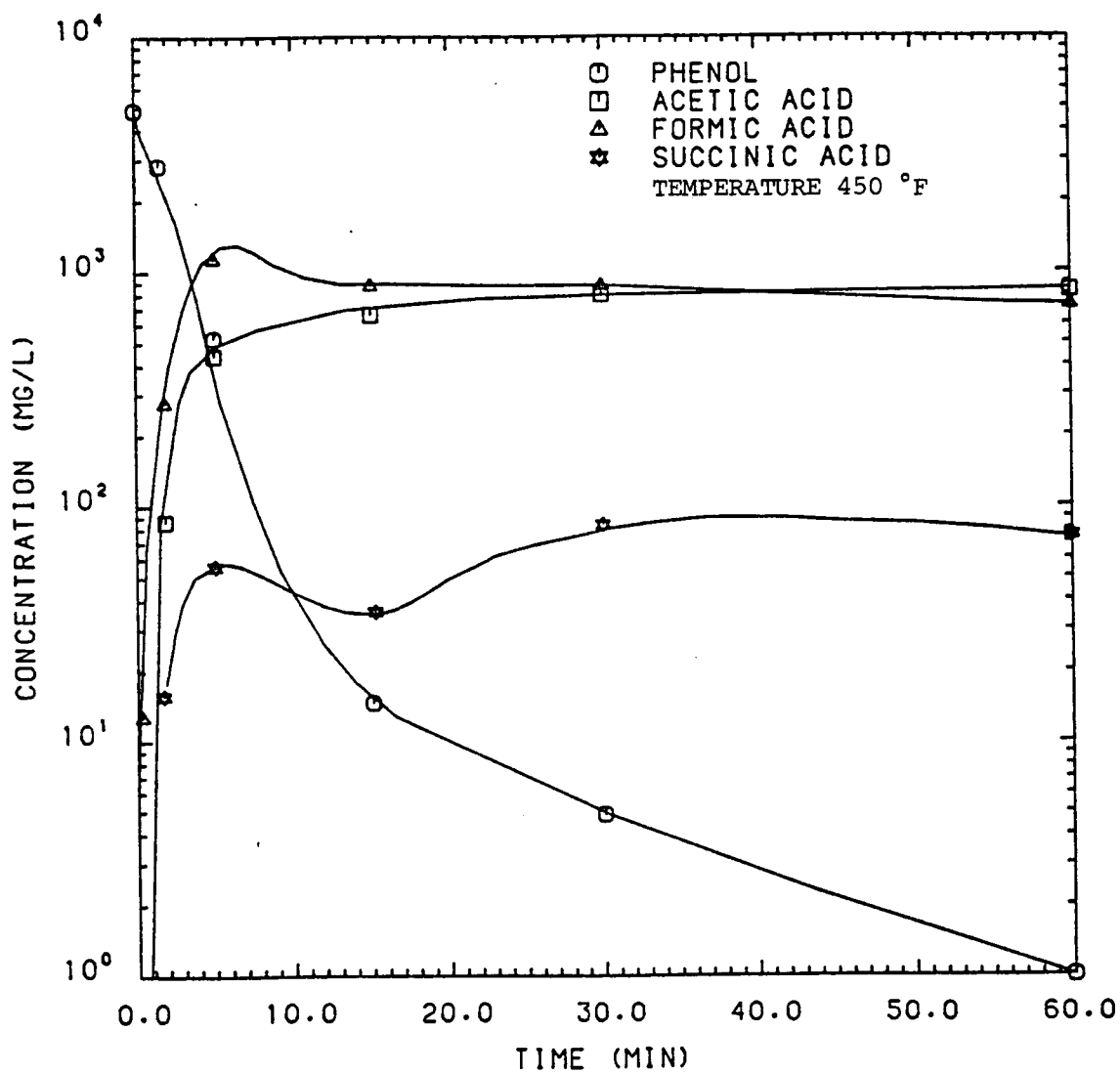


Figure 4. Reaction Curves for Phenol
During Wet Air Oxidation

the process increase the solubility and diffusivity of the oxygen in the reaction mixture providing a strong chemical driving force for oxidation.⁵⁴ This results in relatively high removal efficiencies on the order of 80 to 99% for many compounds. The efficiency, however, is strongly dependent upon the reaction conditions and the material being oxidized. Tables 3 and 4 present results of WAO tests performed on several pollutants. For higher conversions of some pollutants at less severe operating conditions, catalysts such as Cu^{++} have been employed.³⁷

Depending upon the concentrations and type of organic material in the wastewater, the heat of reaction supplies the energy required to maintain the reactor temperature and may even produce excess energy to permit steam generation.⁵⁴ However, with low concentrations of contaminants, steam may be required for this purpose. To minimize steam requirements or maximize steam generation (if feasible), the reactor effluent is used to preheat the incoming feed to the reactor. The effluent then flows into a separator tank to allow gas and liquid phases to separate and reduce the pressure.³⁵ The oxidized liquid phase may require final polishing to bring the residual organic concentrations down to acceptable limits for environmental discharge. This is most readily achieved through biological oxidation at a municipal waste treatment plant or an on-site biological pond.

The removal of phenol and phenolic compounds has been studied by several different workers.^{4,5,14,36,41,42} The focus of much of this work was in developing the reaction kinetics for phenol to estimate the

Table 3
One Hour Wet Oxidation Tests

Compound	Starting Concentration	% Starting Material Destroyed		
		320 °C	275 °C	275 °C Cu ++ Catalyst
Acenaphthene	7.0 g/l	99.96%	99.99%	-----
Acrolein	8.41 g/l	>99.96%	99.05%	-----
Acrylonitrile	8.06 g/l	99.91%	99.00%	99.50%
2-Chlorophenol	12.41 g/l	99.86%	94.96%	99.88%
2,4-Dimethylphenol	8.22 g/l	99.99%	99.99%	-----
2,4-Dinitrotoluene	10.0 g/l	99.88%	99.74%	-----
4-Nitrophenol	10.0 g/l	99.96%	99.60%	-----
Pentachlorophenol	5.0 g/l	99.88%	81.96%	97.30%
Phenol	10.0 g/l	99.97%	99.77%	-----

Source: Randal, T. R., "Wet Oxidation of Toxic and Hazardous Compounds," Presented at the 13th Mid-Atlantic Industrial Waste Conference, Newark, Delaware, June 29-30, 1981.

Table 4
Wet Oxidation Conversion Studies

Compound	Temp.	Catalyst	% Destroyed
Sodium Cyanide (NaCN)	275 °C	0	99.96%
Potassium Thiocyanate (KSCN)	275 °C	0	>99.98%
Formic Acid (CO ₂ H ₂)	275 °C	0	99.3%
N-Nitrosodimethylamine (C ₂ H ₅ N ₂ O)	275 °C	0 ^a	99.56%
	275 °C	+ ^b	99.36%
Toluene (C ₇ H ₈)	275 °C	0	99.73%
	275 °C	+	>99.96%
Pyrene (C ₁₆ H ₁₀)	275 °C	0	99.995%
	275 °C	+	99.997%
Malathion (C ₁₀ H ₁₉ O ₆ PS ₂)	200 °C	0	99.87%
	250 °C	0	99.85%
	300 °C	0	99.97%
1,2-Dichloroethane (C ₂ H ₄ CL ₂)	275 °C	0	99.8%
	275 °C	+	99.9%
Chloroform (CHCL ₃)	275 °C	0	99.92%
	275 °C	+	99.94%
Carbon Tetrachloride (CCL ₄)	275 °C	0	99.99%
	275 °C	+	99.99%
Hexachlorocyclopentadiene (C ₅ CL ₆)	250 °C	0	90.0%
	275 °C	0	98.2%
	300 °C	0	>99.85%
Kepone (C ₁₀ CL ₁₀ O)	280 °C	+	31%
Aroclor 1254 (C ₁₂ H ₄ CL ₂)	320 °C	0	63.0%
	320 °C	+	2.0%
1,2-Dichlorobenzene (C ₆ H ₄ CL ₂)	275 °C	0	2.98%
	275 °C	+	32.2%
	320 °C	+	69.11%

^a No Catalyst Present

^b Catalyst Present

Source: Randal, T. R., "Wet oxidation of Toxic and Hazardous Compounds," Presented at the 13th Mid-Atlantic Industrial Waste Conference, Newark, Delaware, June 29-30, 1981.

removal efficiencies and investigate into the end products of the reaction.

The liquid phase reaction of phenol during WAO involves a series of reaction mechanisms. The reaction is heterogeneous due to the fact that the reactants, oxygen and phenol, are initially in two different phases, gas and liquid respectively. In order for the reaction to take place, the oxygen must diffuse through the gas phase to the gas-liquid interface and then through the interface into the bulk liquid. It has been found that the gas phase resistance for diffusion of this nature is small when compared with the liquid phase resistance and can be neglected in the overall rate of reaction. The rate equation for mass transfer of oxygen through a bubble-liquid interface can be expressed in terms of phenol disappearance as:³⁶

$$r'_{ph} = k_L a V_R S (C^* - C) \quad (1)$$

where

r'_{ph} = rate of diffusion of oxygen through the gas-liquid interface in terms of phenol disappearance, [moles/hr]

k_L = gas-liquid mass transfer coefficient, [cm/hr]

a = gas-liquid interfacial area per unit volume of reactor, [cm /cm]

V_R = reactor volume, [l]

S = stoichiometric ratio of phenol to oxygen

C^* = concentration of oxygen at saturation, [moles/l]

C = reactor concentration of oxygen, [moles/l].

Once the oxygen is in the liquid phase, the reaction between phenol and oxygen can take place. This reaction has been found to be second

order in nature with a rate equation of the following form:³⁶

$$r_{ph} = k_R V_R C_{ph} \quad (2)$$

where

r_{ph} = rate of disappearance of phenol, [moles/l]

k_R = reaction rate constant, [mole/hr-l]

C_{ph} = reactor phenol concentration, [moles/l].

The chemical reaction rate can also be expressed as

$$r_{ph} = V \frac{dC_{ph}}{dt} = F \Delta C_{ph} \quad (3)$$

where

F = liquid feed rate, [l/hr]

ΔC_{ph} = change in concentration of phenol in the reactor, [moles/l]

Rewriting this in terms of the residence time of the reactor, $\theta = V/F$, the rate becomes:³⁶

$$r_{ph} = (V/\theta) \Delta C_{ph} \quad (4)$$

Since the overall reaction requires both diffusion and chemical reaction to take place in succession, at steady state the rates for each step will be equal and therefore:

$$r'_{ph} = r_{ph} \quad (5)$$

Given this and the three rate expressions, C can be eliminated to yield³⁶

$$\frac{C^*}{C_{ph}} = \frac{1}{k_R C_{ph}} + \frac{1}{k_L a S} \quad (6)$$

To express this in terms of conversion of phenol, further manipulation to eliminate C^* must be done. This is achieved by

expressing C^* as:³⁶

$$C^* = P_{O_2}/H \quad (7)$$

where

H = Henry's law constant for oxygen, [atm-l/mole].

P_{O_2} = partial pressure of oxygen in the gas phase, [atm]

This is further modified to account for the fact that the gas bubbles become saturated with water vapor during formation and rise by expressing

P_{O_2} as:

$$P_{O_2} = (P - P^{\circ}_{H_2O}) \quad (8A)$$

where

P = total system pressure, [atm]

$P^{\circ}_{H_2O}$ = vapor pressure of water, [atm]

If air is used rather than pure oxygen, this expression becomes:

$$P_{O_2} = .21(P - P^{\circ}_{H_2O}) \quad (8B)$$

since air is 21% oxygen. The conversion of phenol can be found by combining equations 6,7 and 8A and dividing both sides by the initial concentration of phenol in the feed, $C_{ph,o}$, to yield:³⁶

$$\text{Conv.} = \frac{\Delta C_{ph}}{C_{ph,o}} = \frac{\frac{\theta}{HC_{ph,o}} (P - P^{\circ}_{H_2O})}{\frac{1}{k_R C_{ph}} + \frac{1}{k_L aS}} \quad (9)$$

From this equation, the effects of residence time, pressure and temperature on the removal efficiency can be calculated given the necessary constants for the equation.

Experiments have been performed for the WAO of phenol at both 392 and 482 °F to determine values for k_R and $k_L aS$. The results of these experiments gave values for k_R of 3146 and 9417 l/hr-mole and values for $k_L aS$ of 4.71 and 10.5 hr⁻¹ at 392 and 482 °F respectively. Assuming that these constants are function of temperature only, the Arrhenius equation may be used to estimate the values of the constants at different temperatures.³⁶

Since WAO's introduction in the mid 1950's, its acceptance as a viable alternative in the treatment of contaminated wastewater has grown rapidly. WAO has been used in industry to effectively remove many different organic contaminants. Table 5 is a brief survey of operating WAO processes in several different industries.⁹ Table 6 indicates the results from a pilot plant study for the WAO of coke plant wastewaters.⁹ As is shown in the table, the removal of phenols was 99.8% using a copper catalyst.

ASPEN Simulator

In the operation of a chemical process plant, many decisions must be made concerning new and modification to existing processes. Factors affecting these decisions include return on investment, yields, product specifications, safety, environmental protection, etc.. The difficulty in evaluating these parameters can vary significantly depending on the degree of accuracy desired in the estimate. In the past, obtaining a fairly detailed estimate required many man-hours. However, with the recent developments in computer-aided design, evaluation of process details can be obtained rapidly.

Table 5
Zimpro Wet Air Oxidation Installations

No. Plants	Type of Waste
109	Municipal Sludge
29	Night Soil
12	Carbon Regeneration
7	Acrylonitrile
6	Metallurgical Coking
6	Petrochemical
3	Paper Filler
2	Industrial Activated Sludge
2	Pulping Liquor
2	Hazardous Waste
1	Paper Mill Sludge
1	Explosives, 345-T, Malathion
1	Monosodium Glutamate
1	Polysulfide Rubber
1	Textile Sludge
1	Chrome Tannery Waste
1	Petroleum Refining
1	Misc. Industrial Sludges
186	Total Plants

Source: Canney, P. J. and P. T. Schaefer, "Detoxification of Hazardous Industrial Wastewaters by Wet Air Oxidation," Presented at the National AIChE meeting, March 27-31, 1983.

Table 6
Pilot Scale Wet Air Oxidation of
Coke Plant Wastewater

Wet Air Oxidation Conditions:			
Temperature, °C			279
Flow, GPH			6.3
Residence Time, min.			69
Pressure, PSIG			1558
Catalyst			Cu
Contaminant	Concentration		%Removal
	Feed, [g/l]	Effluent, [g/l]	
COD	5.52	0.48	91.3
Cyanide	309	< 3	99
Phenols	743	1.05	99.8
Cresol	188	< 0.2	99.9
Quinoline	38	13	65.8

Source: Canney, P. J. and P. T. Schaefer, "Detoxification of Hazardous Industrial Wastewaters by Wet Air Oxidation," Presented at the National AIChE meeting, March 27-31, 1983.

One of the most recent design simulation programs is ASPEN (Advanced System for Process ENgineering). The program was written in 1979 by a group of faculty, graduate and undergraduate students at MIT using the routines in the FLOWTRAN ⁴⁶ process simulator developed by Monsanto as an initial starting point. The program was then tested in a two year program by fifty-five industrial companies and government groups. During this phase, many errors in the program were found and corrected to produce a final version of ASPEN in 1981 containing 353,482 lines of FORTRAN code in 1511 subroutines.¹⁸

The program itself was developed in a modular concept which allows the user to select and execute only the subroutines needed to perform the desired simulation. The complex process of generating a calling program to link the subroutines together is greatly simplified through the use of a new language and interpreter. The interpreter, (or input translator), performs several functions which include checking the input language for syntax errors; retrieval of physical-property data from the data base; generation of the calculation sequence if not specified; and creation of a calling program to link together the necessary subroutines to perform the simulation. The ASPEN programming language was also simplified through the use of syntax based on terminology that is commonly used in the design of a chemical process.²⁹

The ASPEN program allows the user a great deal of flexibility and a wide variety of program functions. Some of the major features of ASPEN include a physical-property data base, data regression system, process simulator, cost and sizing estimation, capital investment estimation,

operating costs estimation and profitability analysis. What follows is a brief discussion about some of these features.

The physical-property system in ASPEN is a comprehensive system which includes a data base containing pure component physical-property constants for over 480 conventional compounds and property models to calculate enthalpy, entropy, free energy, molar volume, equilibrium ratios, fugacity coefficients, viscosity, thermal conductivity and diffusion coefficients. In addition to this, the user can simulate properties for non-conventional compounds such as coal, oil, wood, or any compounds with complicated or indefinite chemical structures. The user can also add chemicals not in the data base for the simulation by providing the necessary constants for the property models from the input language. To assist the user in doing this, ASPEN has a data regression system which can regress property parameters from data for any chemical supplied by the user. With these features, the user can retrieve or develop the physical properties necessary to perform a simulation of a process.²⁹

The process simulator in ASPEN is a powerful system capable of performing mass and energy balances around any of the equipment models shown in Table 7.¹⁷ These models can be combined together in any sequence desired by the user to perform a flowsheet simulation in terms of a steady state mass and energy balance. In addition to the models shown in Table 7, the user can also supply models to use in conjunction with the ASPEN models to simulate a special or proprietary system.

Once the mass and energy balance has been calculated using the simulator, the user has the option of performing sizing and costing of

Table 7

Unit Operations Available in ASPEN

<u>Splitters and Separators</u>	<u>Flash Models</u>
Flow Splitter	Two-Phase Flash
Substream Splitter	Three-Phase Flash
Component Splitters	General Flash
Two-Product Separator	
Substream Attribute Separator	
	<u>Distillation</u>
<u>Pressure Changers</u>	Shortcut, Design (Winn, Fenske Underwood), DSTWU
Pump/Slurry Pump	Shortcut, Rating (Edmister)
Compressor/Turbine	Rigorous Distillation, (KB Method), FRAKB
	Rigorous Absorption, Rating
<u>Heaters and Heat Exchangers</u>	Rigorous Multistage Separator, RADFRC
General Purpose Heater	Liquid-Liquid Extraction, EXTRC
Cocurrent/Countercurrent Heat Exchanger	
	<u>Reactors</u>
<u>Solids Handling</u>	Yield-Based Reactor
Cyclone	Fractional Conversion Reactor
Electrostatic Precipitator	CSTR Reactor Model
Fabric Filter	Plug Flow Reactor Model
Venturi Scrubber	Combined-Phase Chemical Equilibrium
Crusher	General Phase and Chemical Equilibrium
Screen/Classifier	
Hydroclone	
Rotary Vacuum Filter	
Filtering Centrifuge	
	<u>Stream Manipulators</u>
<u>Mixer</u>	Multiplier
General Mixer	Class Changer
	Duplicator

the equipment in the process flowsheet using ASPEN's costing routines. The costing data base includes cost correlations for 17 different equipment types developed from 1979 cost data. With these routines, the user can develop a preliminary-grade estimate for purchased-equipment costs with accuracy in the range of $\pm 30\%$.¹⁸ The system allows the user to override the calculated cost of any equipment either by supplying the base and purchased cost of the equipment or by supplying a user correlation to estimate the cost.

The purchase equipment costs can be further utilized to generate a capital cost estimate for the overall process using ASPEN's capital investment estimation routines. This estimate is based on a factor method to calculate both direct and indirect capital costs associated with the process. Direct equipment and labor costs can be calculated based on the total purchased-equipment cost or on individual purchased-equipment item costs and can be divided into eight components or combined as one overall component. The eight components for which cost can be estimated are piping, concrete, steel, instrumentation, electrical installation, insulation, paint and a miscellaneous cost. This allows the user to account for increased installation cost due to equipment type and complexity. Capital costs for utility construction, buildings, service systems, site development, project engineering, project management, engineering home office, insurance, contingencies, and land are also calculated via factors as well.²⁹

In addition to capital cost, ASPEN can also estimate operating costs. The information concerning consumption rates for raw materials and utilities along with the production rates for products and

by-products can be completely specified by the user or retrieved from calculations made during a simulation. Prices for raw materials, utilities and by-products must be specified by the user. Additional overhead costs for labor, engineering, maintenance and supplies are calculated using factors or values supplied by the user. This information can be used with the capital cost estimate to perform a profitability analysis to estimate the return on investment or the product selling price.²⁹

To resolve the problem where in the costing information is based on 1979 dollars, cost indexing to any desired year can be performed using the Marshall and Swift, Chemical Engineering or Nelson Refinery indices.

When all of these features are combined, ASPEN becomes a very powerful and flexible tool which can be used to analyze and compare process alternatives. ASPEN's power as a process simulator has been demonstrated through its use in the simulation of a large variety of processes. Table 8 is a summary of a few of the types of simulations which have been performed by ASPEN.¹⁸

Additional Programs of Interest

In addition to ASPEN, two other programs were utilized for the work development of this thesis. These programs are PRPIEA and PATTERN. What follows is a brief summary of the history and major features of each program.

Table 8

Summary of Simulations Performed by ASPEN

<u>Coal Gasification</u>	<u>Chemicals</u>
HY-GAS Gasifier	Reactor Models
CO ₂ -Acceptor Process	Distillation Systems
Conoco Gasification Process	Ethanol-Water Separations via
Air Separation	Azeotropic Distillation
Lurgi Gasifier- (Non-Kinetic)	Mixed Xylenes Separation via
Lurgi Gasifier- (Kinetic)	Superfractionation and
Gasifier Quench	Crystallization
Rectisol- H ₂ S Removal	Critical Solvent Separations
Shift and Methanation	Sulfuric Acid Processs from SO ₂
Reaction System	Second Law Analysis for
Hot Carbonate CO ₂ Removal	Energy Efficiency
Compression and Drying	Soda Ash Manufacturing
Combined Cycle Power	
Generation	<u>Papermaking</u>
	Pulp Refiner
<u>Coal Liquefaction</u>	<u>Hydrometallurgical Process</u>
Coal Liquid Properties Estimation	Bayer Alumina Purification
Coal Liquid Pseudo Fraction	Multiple Effect Evaporator
Characterization	Material Sizing Operations
Exxon Donor Solvent Process	
Liquefaction Reactor Area	<u>Environmental Control Processes</u>
Solvent Hydrogenation Area	Flue Gas Desulfurization via
Product Distillation Area	Lime Solution Scrubbing
	Particulate Removal from Gases
<u>Shale Oil Retorting</u>	Sour Water Stripping
Fluidized Bed Retort and	Sour Gas Cleanup
Regenerator System with	
Solid-Gas Equilibrium for	<u>Process Costing</u>
CO ₂	Exxon Donor Solvent Process
<u>Petroleum</u>	Conoco Gasification Process
Distillation Systems	Nuclear Processing Facilities
Hydrodesulfurization System	
Alkylation Unit Fractionator	
With Heat Pump	
Natural Gasoline Separation	
Train	

PRPIEA

The PRPIEA program is a versatile discounted cashflow analysis program which can be used to evaluate the affect of capital investment, operating expenses, interest rates, product and by-product prices and other related information on the main product price or return on investment, (ROI), for a chemical or related process.⁴⁴

The program was originally called PRP and was conceived by Royes Salmon at Oak Ridge National Laboratory to evaluate the profitability of various synthetic-fuels processes in 1977.⁴⁴ This program was modified by Salmon in 1979 to allow the user to specify escalation rates for costs and prices and to allow repayment of debt capital by means of a constant annual principal reduction. The program went through further modifications by IEA Coal Research in London, England to add a section for calculating working capital from the user supplied data. This program was then received at the University of Tennessee, Knoxville in 1982 where it was restructured to run using a WATFIV FORTRAN compiler on an IBM 370 computer. One further addition was made to the program at this point to include the 1981 Tax Law depreciation schedule.

The PRPIEA program calculates the main product price or the ROI based on a discounted cashflow analysis. In this analysis, the main product price or ROI is iterated upon until the sum of the discounted cashflows is zero, i.e. until sum of the discounted cash inflows is equal to the sum of the discounted cash outflows. In this concept the initial capital investment is gradually reduced over the lifetime of the project until the outstanding investment at the end of the project life is either zero or equal to the salvage value of the equipment.⁴⁴

When the after tax ROI is being calculated, the yearly after-tax cash inflows are calculated using the specified product price and the cash outflows are determined from investment data supplied by the user. The discount rate or ROI is then iterated until the sum of the after-tax cash inflows discounted to a specific year using the iterated discount rate equals the sum of the cash outflows discounted to the same year with the same discount rate. On the other hand, if the main product price is the value of interest the program iterates on the main product price until the sum of the calculated after-tax cash inflows discounted to a specific year using the specified ROI equals the sum of the cash outflows discounted to the same year with the given ROI.⁴⁴

The PRPIEA program offers great freedom for the handling of investments, capital structure, operating expenses and taxes. In defining the life of a given project, the user can partition the project into a construction period and operation period. This allows a distinction between expenses and investments which occur during construction and those that occur during operation of the process.⁴⁴

Investments made during the project life can be defined using several different methods. The investments can be separated into depreciable capital investments such as equipment, material and installation labor costs and non-depreciable capital investments such as working capital. Depreciable investments can be entered as an annual sum during each year of the project life at the start of each year or as a fraction of an overall lump sum during each year of construction. The investments can be further broken down into depreciation classes to account for the various types of equipment having different depreciable

lives. The working capital required for plant operation is estimated in the program using information supplied by the user concerning raw material and product storage; cash on hand to pay staff; time required to collect sales and pay bills; land and other miscellaneous material cash requirements. The financing of these investments can also be specified as a constant debt and equity fractions. Interest accrued on the debt is supplied by the user as an annual rate with special features for interest during construction and is calculated based on the outstanding debt at the start of each year.⁴⁴

Annual depreciation of the capital investment can be calculated in the program from one of five different methods. These methods are sum of years digits, double declining balance with conversion to straight line, 1981 Tax revision and no depreciation allowances. If any of the equipment can be salvaged at the end of the project life, the salvage value can also be specified by the user.⁴⁴

Expenses accrued over the project life can be divided into those which occur during construction, start-up and annual operation. The annual operating expenses can be entered as a constant annual expense and as expenses which are proportional to the on-stream plant efficiency. Proportional expenses are entered at 100% efficiency and are adjusted using the plant efficiencies specified for each year of plant operation by the user. Proportional expenses from consumption of raw materials are calculated from consumption rates and prices specified by the user for each raw material. Profits obtained from sale of products and by-products are also scaled using the plant efficiencies for each year given production rates and specified or iterated prices. Overhead

expenses such as maintenance, insurance and labor can be specified either as a percentage of total investment plus contingency or as an add-on sum. Expenses for raw materials, labor and miscellaneous materials accrued during start-up can also be specified by the user.⁴⁴

Taxes on sales and property can be entered by the user allowing some degree of flexibility in how and when taxes are applied. Taxes can be applied at both the federal and state level on income with additional provisions for a state revenue tax. For years in which the net income before taxes is negative, the user can specify whether negative income taxes are allowed and if tax losses are to be carried forward. In addition to income taxes, the program has allowances for investment tax credits and property taxes.⁴⁴

The program can also allow the user to adjust prices, expenses and investments to account for inflation by providing escalation factors.

In summary, the PRPIEA program is a useful tool which allows the user to account for a wide variety of factors which affect the profitability of a given process.

PATERN Multi-parameter Search Program

The PATERN program is basically a simplified numerical search program used to minimize a unimodal function with as many as 1000 independent variables.³⁰

The program was written by C. F. Moore, C. L. Smith and P. W. Murrill in 1968 primarily for use in their studies of variables which affect process control. The program's approach to minimization is more one of trial and error rather than an optimum mathematical approach of evaluating the gradient and performing a one dimensional search along the

gradient. The program starts with initial guesses for the independent variables and perturbs each variable, one at a time, by a specified increment size in both the positive and negative direction to determine the effect of the adjustment on the function to be minimized. The program simply compares the results of these manipulations and adjusts the variable which provided the best improvement. The procedure is repeated using the new values for the variables until successive iterations provide no improvements.³⁰

In order to use PATTERN in a parametric search, the user must create three programs to use in conjunction with PATTERN. These programs include a main program to define all the necessary search parameters required by PATTERN such as the number of independent variables, initial guesses for these variables and the increment size to perturb these variables in the search; a subroutine called PROC containing the function on which the search is to be conducted; and a subroutine called BOUNDS which contains the boundary limits for each of the variables in the search. These programs are combined using calling statements such that the main program calls PATTERN as a subroutine to initialize the search process. PATTERN then repeatedly calls PROC to calculate the current value of the objective function using the adjusted independent variables. PATTERN also calls BOUNDS to confirm whether or not the adjusted variables are within the specified boundaries.³⁰

The structure of the PATTERN program allows its usage for searches in a great number of different applications. One must be careful, however, in using the program for functions which exhibit more than one minimum. In cases of this type, the search will terminate at different minimum

values depending upon the initial starting point of the search. This can also leave the user with the problem of trying to decide if the minimum found is just a local minimum point for the given function. This problem can be somewhat overcome by running PATTERN with several different initial guesses and selecting the result which provides the lowest value of the minimizing function.³⁰

III. PROJECT DEFINITION

Process Description

Due to the high concentrations of contaminants in the wastewater from a synthetic-fuels plant and the great number of chemical compounds in the water, it is possible that a Phenosolvan type process would not remove the contaminants down to a level where biological oxidation can be employed. One potentially attractive addition to the Phenosolvan process would be a Wet Air Oxidation, (WAO), process. This combination offers the advantages of removal and recovery of phenols as a by-product and substantial reductions in many of the contaminants shown in Table 1 on page 4 which may not be removed by the Phenosolvan process. A block representation of such a process is shown in Figure 5.

In the combined process the contaminated water after being processed to remove ammonia and sulphur is contacted countercurrently in an extraction tower with a solvent to extract phenols from the aqueous stream. The extract from this tower is processed through the primary solvent recovery system to separate the extracted phenols from the solvent. The raffinate stream is processed through a secondary solvent recovery system consisting of a preheater, distillation tower, cooler and decanter to recover residual solvent from the dephenolated water. The solvent recovered in both the primary and secondary recovery systems is recycled back to the extraction tower where it is mixed with fresh solvent to make up for solvent losses. The water from the secondary recovery loop still contains many compounds not removed in the extraction system and also contains residual solvent and phenols. This water is

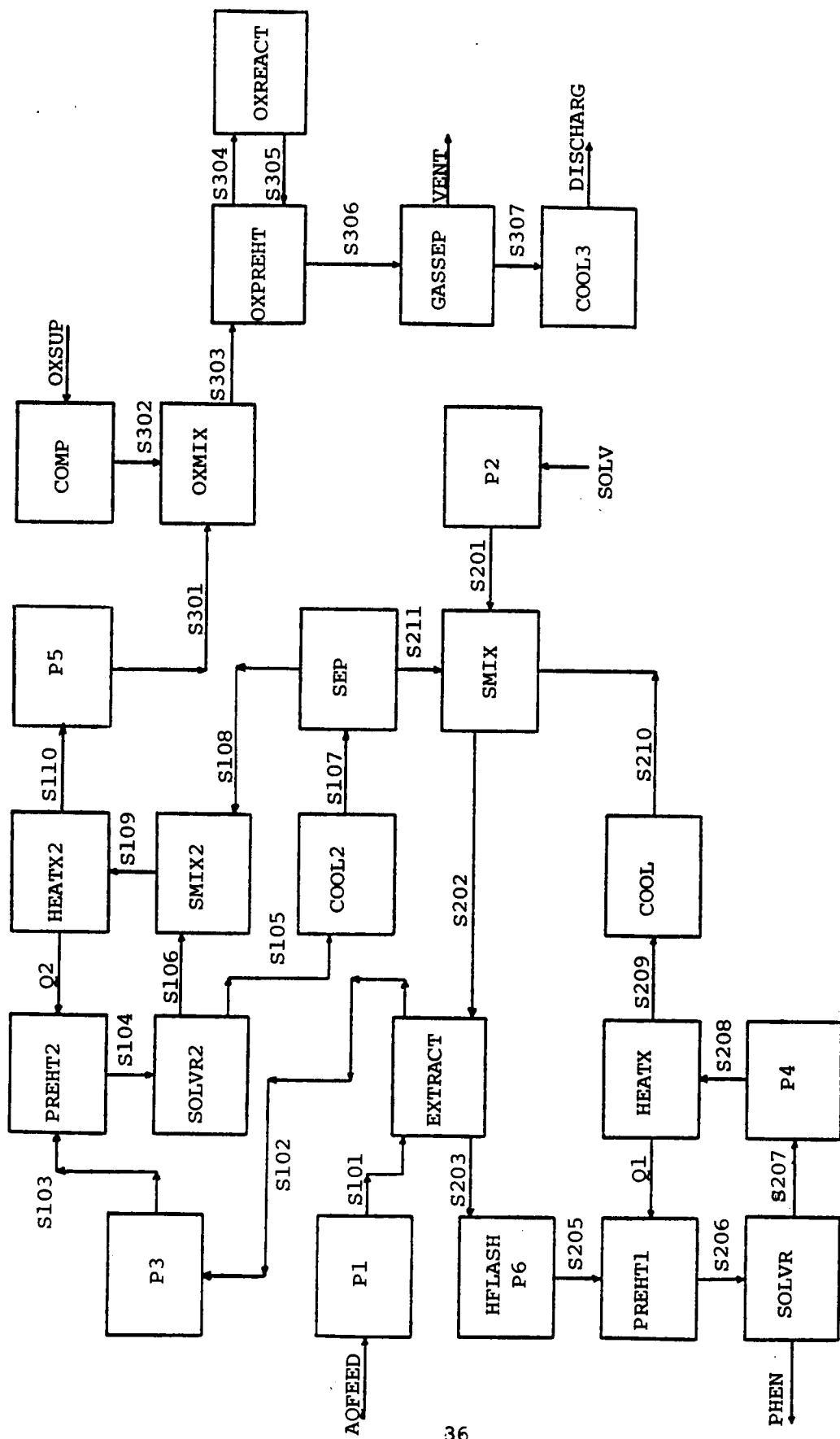


Figure 5. Block Representation of a Combined Phenolsolvan--WAO Treatment System.

further processed through a wet air oxidation system as described in Section 2. The oxidized water discharged from this system will still contain residual concentrations of many organics and may require further processing through a biological oxidation system if the water is to be discharged to the environment.

Design Considerations

In designing a chemical process, many parameters must be evaluated to determine equipment sizes and ultimately the overall cost of the process. These parameters include plant capacity, physical properties, operating parameters, and product purities. Often, selection of optimum values for these various parameters involves trade-offs between the parameters. This means that great care must be taken in the initial phases of design in order to identify these trade-offs.

Process Capacity

One of the most important parameters in a chemical process is the operating capacity. This parameter, more than any other, contributes to the determination of the size and cost of the equipment.

For a synthetic-fuels plant, the capacity is typically described in units of tons of coal throughput per day. For the purpose of this study, a plant processing 20,000 tons/day of coal was selected as a basis since several plants of this size are being considered.⁸ Since the proposed process is concerned with clean up of the wastewater produced from such a plant, a relationship between the amount of wastewater produced per ton of coal feed must be determined. Based on information from Jahnig²² and from Beychok⁶, a ratio of one pound of water per pound of coal was

selected as a good approximation. Therefore, for a 20,000 ton/day synthetic-fuels plant, a wastewater flow of 1.67×10^6 lb/hr, (3330 GPM), results.

This water contains a great many organic and inorganic contaminants which must to be removed to a varying degree. However, it is somewhat impractical to try to consider all of the contaminants for a preliminary design estimate. For this reason, only those contaminants which could significantly effect the size and operating parameters of the equipment are examined. This assumption limited the contaminants considered for this study to phenols.

Phenols were selected for several reasons including; 1) their relative abundance with respect the the other contaminants; 2) their recoverability as a valuable by-product; 3) their solubility in water and, 4) the low discharge concentrations required by Federal and state laws. To date, no Federal regulations concerning wastewater from a synthetic-fuels plant have been established but a close approximation may be estimated using the new source BATEA, (Best Available Technology Experimentally Achievable), standards established by the Environmental Protection Agency for coke-oven wastewater.²⁶ These standards are given as a 30-day average limit for some specific contaminants in pounds per pound of coke produced. The value for phenol translates to 130 ppb for treated condensate wastewater on an undiluted basis.⁴⁸ A survey of literature values for the concentration of phenols in the wastewater resulted in an assumed initial concentration of 7000 ppm.^{7,16,48,53} Given this value, the BATEA standard represents a > 99.99% removal of phenol.

A further assumption was made concerning phenols due to the lack of adequate liquid-liquid equilibrium information for the various phenolic compounds. This assumption is that pure phenol is representative of all the phenolic compounds with respect to its solubility in the wastewater. The validity of this assumption depends upon the relative concentration of phenol with respect to other phenolic compounds in the water. As was stated before, this is a function of the type of gasifier, operating conditions and type of coal used as feedstock. Even if the other phenolic compounds are more prevalent than phenol, the assumption is probably valid if these phenols are mainly monohydric and have distribution coefficients similar to that of phenol in the selected solvent.

Given these assumptions, the wastewater feed to the treatment system was characterized as having 1.655×10^6 lb/hr of H_2O and 1.167×10^4 lb/hr of phenol.

Physical Properties

One of the most important steps in the design of a chemical process is the gathering and selection of physical properties for the compounds of interest. The basis for almost all of the calculations in the design of process equipment is derived from physical-property information for the chemicals involved. Without this information, many of the calculations cannot be performed. The accuracy of this information is also a matter of great importance from a design standpoint. Inaccurate property data or estimations may produce erroneous results leading to misleading or false conclusions. This means that great care should be taken to gather accurate data on the chemicals used in the process and,

to use reliable models to simulate the physical properties which are needed.

For the simulation at hand, the task of gathering physical-property data has been greatly reduced through the use of ASPEN's physical-property data base. Pure component data for all of the chemicals in the wastewater treatment process are included in the data base of ASPEN. Given this data, however, selection of property models to accurately simulate the necessary physical properties at the conditions of the process had to be resolved. This too was simplified through the use of property option sets in ASPEN. These option sets combine property models together to calculate all of the physical properties required by the equipment models. Different option sets can be selected depending upon the expected degree of interaction between chemicals and the operating conditions of the process such as component non-ideality, temperature and pressure ranges, etc.

For the simulation of the Phenosolvan and WAO process, two different option sets were selected. This was due to the very different types of operations and conditions in the two processes. The two option sets selected were SYSOP5 and SYSOP11. A summary of the property models which make up these option sets is shown in Table 9.¹⁸

SYSOP5 focuses primarily on a modified version of the Benedict-Webb-Rubin, (BWR), equation of state to predict most of the major thermodynamic properties. A general description of this equation and its use can be found in the "ASPEN Technical Reference Manual".²⁸ SYSOP5 has been described as the most accurate system option for pure component properties by Seader.⁴⁵ This option set was selected for the

Table 9
ASPEN System Option Sets

Property	Option Set Number	
	5	11
Vapor Fugacity Coefficient	BWR Conformal Sol'n	Redlick-Kwong
Liquid Fugacity Coefficient	BWR Conformal Sol'n	UNIQUAC Symmetric Extended Antoine
Vapor Mixture Enthalpy	BWR Conformal Sol'n	Redlick-Kwong
Liquid Mixture Enthalpy	BWR Conformal Sol'n	Cavett
Vapor Mixture Free Energy	BWR Conformal Sol'n	Redlick-Kwong
Liquid Mixture Free Energy	BWR Conformal Sol'n	UNIQUAC Symmetric Extended Antoine
Vapor Mixture Entropy	BWR Conformal Sol'n	Redlick-Kwong
Liquid Mixture Entropy	BWR Conformal Sol'n	Cavett UNIQUAC Symmetric Extended Antoine
Vapor Mixture Volume	BWR Conformal Sol'n	Redlick-Kwong
Liquid Mixture Volume	BWR Conformal Sol'n	Rackett
Vapor Mixture Viscosity	Dean-Stiel	Dean-Stiel
Liquid Mixture Viscosity	Letsow-Stiel	Letsow-Stiel
Vapor Mixture Thermal Cond.	Stiel-Thodos	Stiel-Thodos
Liquid Mixture Thermal Cond.	Sato-Reidel, Vredeveld	Sato-Reidel, Vredeveld
Vapor Mix. Diffusion Coeff.	Dawson-Khoury- Kobayashi	Dawson-Khoury- Kobayashi
Liquid Mix. Diffusion Coeff.	Wilke-Chang-Vignes	Wilke-Chang-Vignes
Surface Tension	Hakim-Steinberg- Stiel, Power Law Mixing	Hakim-Steinberg- Stiel, Power Law Mixing

simulation of properties in the WAO section of the process because the streams contain essentially pure water or oxygen.

SYSOP11 focuses on the use of the UNIQUAC activity coefficient model and the Redlick-Kwong equation of state to account for highly non-ideal interactions which may occur in component mixtures. This option set was selected for the simulation of properties in the Phenosolvan section of the process primarily because of the ability of the UNIQUAC equation to accurately predict phase separation.²¹ This was considered very important since the primary operation in the Phenosolvan process is a solvent extraction process involving two separate phases.

Binary interaction parameters are required by both the BWR equation of state and the UNIQUAC equation to accurately account for the non-idealities which occur in a component mixture. The binary parameters for the BWR equation of state were assumed to be zero because of the relatively pure component process streams where the equation was applied. This was not true, however, for the UNIQUAC equation. In order to properly simulate the equilibrium phases in the solvent extraction process, binary interaction parameters for the UNIQUAC equation had to be regressed from liquid-liquid equilibrium data.

To regress the necessary interaction parameters, the Data Regression System, (DRS), of ASPEN was initially selected. Several attempts were made with different equilibrium data using the DRS but, results could not be obtained due to severe errors which occurred during the regression. These problems seemed to be inherent in the system itself and could not be resolved for this work. This prompted the development of a separate computer model of the UNIQUAC equation in

coordination with the search program PATTERN to regress the interaction parameters.

The UNIQUAC computer model was developed from the equations presented by Henley and Seader.²¹ In order to develop a basis from which the UNIQUAC interaction parameters could be regressed, an expression was needed to relate the activity coefficient determined using the UNIQUAC equation to the experimentally measured equilibrium data. This expression can be found by using the definition of liquid equilibrium. For multiple-liquid phases to be in equilibrium, the activity of each substance in a given phase must equal the activity of the same substance in all the other phases. For a two phase system, this translates to

$$a_i^I = a_i^{II} \quad (10)$$

where

a_i^I = activity of component i in phase I.

a_i^{II} = activity of component i in phase II.

Given that

$$a_i = \gamma_i x_i \quad (11)$$

where

γ_i = activity coefficient of component i.

this can be rewritten as

$$\gamma_i^I x_i^I = \gamma_i^{II} x_i^{II} \quad (12)$$

or

$$\frac{x_i^I}{x_i^{II}} = \frac{\gamma_i^{II}}{\gamma_i^I} = K_{Di} \quad (13)$$

where

K_{Di} = distribution coefficient of component i.

x_i = mole fraction of component i in the given phase.²¹

Given equation 12 or 13, one can relate the experimentally measured values of X_i to the estimated values of x_i and minimize the difference by adjusting the interaction parameters of the UNIQUAC equation.

This was accomplished through the combination of several programs which are shown in Appendix A.* The first program shown is called MAIN and is the main program from which subsequent subroutines are called. This program retrieves information from the user which is needed to perform the calculations of the UNIQUAC model. This information includes the number of components and their names; van der Waals volume and molecular surface area parameters for each component; and the number of equilibrium data points and the initial mole fractions of each component in their respective phases. The main program supplies initial guesses for the regression parameters and regression control information for the PATTERN subroutine. The main program calls PATTERN to start the regression. PATTERN first checks the regression parameters to see if they are within specific limits by calling the subroutine BOUNDS. This subroutine contains limits for each of the regression parameters and passes an error code back to PATTERN if the limits are violated. If the parameters are within the limits, PATTERN begins its regression algorithm and repeatedly calls the subroutine PROC, which contains the UNIQUAC model, to evaluate the minimization parameter called COST. COST is calculated in PROC as the summation of the absolute value of the differences between the calculated activity of all components in phase I

*The PATTERN search program was not included in this listing.

with respect to the same components in phase II for all the given equilibrium data points. I.e.:

$$\text{COST} = \sum_{j=1}^N \sum_{i=1}^C (\gamma_i^I x_{ij} - \gamma_i^{II} x_{ij}) \quad (14)$$

where

N = Number of equilibrium data points

C = Number of chemical components.

Equation 12 was used as a basis for the minimization rather than equation 13 because the differences in the activities for a component were more of the same magnitude as differences in the activities for the other components.* Once the PATTERN subroutine has determined that varying the parameters does not improve the value of cost, the resulting values are passed back to the main program where they are printed.

Several problems exist with using this method of data regression which are mainly due to the crude nature in which PATTERN works. The major problem is that there are many local minimums in the cost function of equation 14. One minimum for the equation can be found such that it is always zero and thus the value of cost will be zero. This trivial solution, however, is not desired. The minimum that is found depends

*Several different methods were attempted to normalize the difference such as dividing the activity difference by the average value of the activity in the two phases for a given component. The results from using these methods were inconclusive because it was difficult to conclude whether the values obtained were any better than the unnormalized values.

strongly upon the initial starting values of the parameters. This problem necessitates a trial and error process in which the initial starting parameters are varied and the resulting costs at each minimum are with the lowest cost considered to be the "best" result. Of course, since there is an infinite number of starting points, it is impossible to state that the parameters resulting from the lowest cost value represent the absolute minimum value of the function.

The regressed binary parameters from the PATTERN search provides an opportunity to generate an estimated equilibrium curve using a liquid-liquid flash model in ASPEN called FLASH3. This is done by supplying the FLASH3 block with input streams of varying concentrations of Phenol, Water and solvent at a given temperature and pressure while setting the input heat and outlet vapor streams equal to zero. The block calculates the equilibrium concentrations of the two phases using the supplied UNIQUAC parameters and generates separate streams for each phase as output. These data can then be plotted against the experimental equilibrium curve to compare the results.

Operating Parameter Selection

Before one can design a process unit, it is necessary to determine the conditions under which the equipment is to operate. Primarily this includes temperature, pressure and throughput. Additional parameters, such as residence time, reflux ratios, separation purities and solvent ratios, may be needed depending on the type of equipment used. This information may be extracted in some cases from a survey of literature information or evaluated using fundamental design criteria as a basis.

Phenosolvan process. The primary items of process equipment in the Phenosolvan process are the solvent extraction tower and the primary and secondary solvent recovery towers. The remaining flowsheet equipment operating conditions depend directly on these three process units.

The solvent extraction tower is the central piece of process equipment in the Phenosolvan process. Its operating conditions ultimately determine the conditions for the remaining equipment in the process. There are many parameters to consider in the operation of a solvent extraction tower. The most important of these parameters is the solvent which is to be employed in the tower. An ideal solvent should possess the following characteristics:⁵⁰

1. The solvent should be relatively insoluble in the feed to reduce solvent losses.
2. The selectivity of the solvent should be high for the solute. I.e., the ratio of the amount of extractant in the solvent phase with respect to the extract in the feed phase should be high.
3. The distribution coefficient of the extractant should be high. This is desirable because it decreases the amount of solvent needed for the extraction.
4. The solvent should be easily separable from the extractant and raffinate so that it can be recovered for reuse with minimal losses.
5. The difference in the densities of the two phases should be large to allow for better phase separation.
6. The solvent should be chemically stable and not react with the other components in the system.

7. The solvent should be nontoxic, nonflammable and have a low unit cost

It is generally very difficult to find a solvent which possesses all of the desirable characteristics mentioned above so, some trade-offs must usually be made.

For phenol, there are several solvents which have desirable characteristics. Kiezyk and Mackey²³ present a brief listing of solvents suitable for phenol removal. Saha⁴³ and Abrams¹ also give information on distributions of phenols in aqueous solutions with respect to several solvents. Although there is a large number of solvents which are suitable for phenol removal, reliable liquid-liquid equilibrium data on the solvents with respect to phenol and water is limited and difficult to find. This reduced the selection of possible solvents down to three candidates; n-butyl acetate, (NBA), methyl isobutyl ketone, (MIBK), and isoamyl acetate.^{31 32} Of these, n-butyl acetate was selected due to its low solubility in water, high distribution coefficient, high selectivity and moderate cost. It was also selected because it is believed to be one of the components of the solvent used in the Phenosolvan process.

The binary interaction parameters for the UNIQUAC equation were evaluated from equilibrium data presented by Narasimham, Reddy and Chari³² using the PATTERN search program. The results of this regression are presented in Appendix A. These parameters were then used in an ASPEN program to generate the equilibrium phase curve estimated using the UNIQUAC equation. The results from this program are shown graphically in Figure 6 along with the experimental equilibrium curve and are also summarized in Table 10.

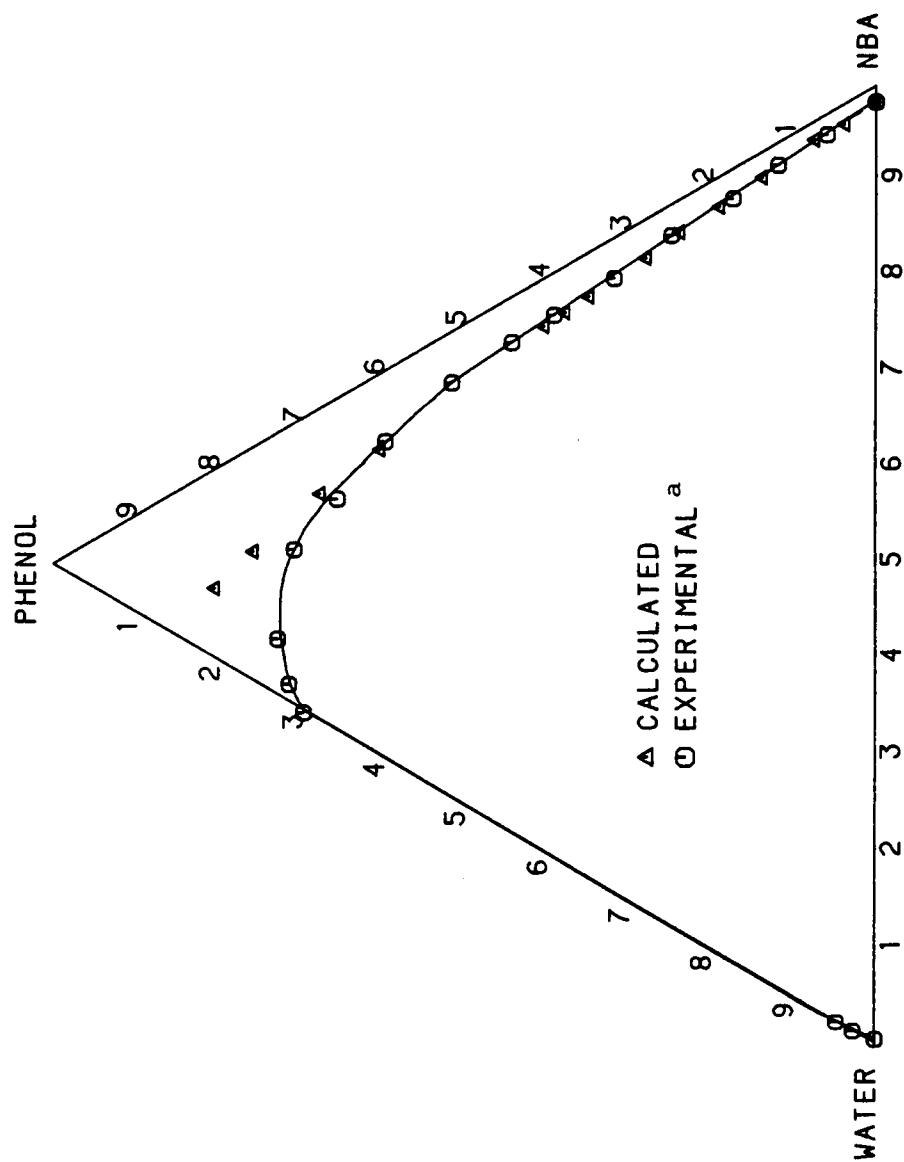


Figure 6. Experimental and Predicted Equilibrium Curves for the Phenol--Water--N-Butyl Acetate System at 30 °C

^aAll values are shown in weight percent

Table 10

Equilibrium Data for the Ternary System
Phenol--Water--N-Butyl Acetate at 30 °C^a

Literature Data					
Extract Phase			Raffinate Phase		
Phenol	Water	NBA	Phenol	Water	NBA
-	0.0145	0.9855	-	0.9971	0.0029
0.0600	0.0150	0.9259	0.0015	0.9955	0.0030
0.1200	0.0175	0.8625	0.0020	0.9950	0.0030
0.1725	0.0275	0.8000	0.0040	0.9930	0.0030
0.2456	0.0300	0.7250	0.0070	0.9900	0.0030
0.3175	0.0375	0.6450	0.0130	0.9840	0.0030
0.3900	0.0400	0.5700	0.0185	0.9785	0.0030
0.4400	0.0425	0.5175	0.0245	0.9730	0.0025
0.5125	0.0500	0.4375	0.0325	0.9655	0.0020
0.5925	0.0700	0.3375	0.0450	0.9530	0.0020
0.6505	0.1020	0.2475	0.0570	0.9415	0.0015
0.7050	0.1300	0.1656	0.0676	0.9320	0.0010
0.7250	0.2150	0.0600	0.0725	0.9265	0.0010
0.7120	0.2680	0.0200	0.0775	0.9215	0.0060
0.6939	0.3061	-	0.0880	0.9120	-
UNIQUAC Estimation					
Phenol	Water	NBA	Phenol	Water	NBA
-	0.0102	0.9898	-	0.9953	0.0047
0.0383	0.0129	0.9488	0.0014	0.9941	0.0045
0.0735	0.0155	0.9110	0.0028	0.9929	0.0043
0.1363	0.0266	0.8432	0.0055	0.9964	0.0040
0.1897	0.0253	0.7850	0.0083	0.9880	0.0037
0.2366	0.0296	0.7338	0.0108	0.9857	0.0035
0.2779	0.0335	0.6886	0.0134	0.9833	0.0032
0.3465	0.0406	0.6128	0.0184	0.9788	0.0028
0.3751	0.0438	0.5811	0.0208	0.9766	0.0026
0.4031	0.0467	0.5520	0.0231	0.9744	0.0025
0.6046	0.0740	0.3214	0.0484	0.9504	0.0012
0.6736	0.0867	0.2396	0.0610	0.9382	0.0008
0.7569	0.1074	0.1357	0.0796	0.9206	0.0004
0.8026	0.1252	0.0721	0.0919	0.9080	0.0002

^aAll values are given in weight percent

Another important parameter in the design of an extraction system is the water to solvent ratio. This parameter is related to the distribution coefficient. The larger the K_D value for the solute, the greater the water to solvent ratio. Since operating information about the NBA--Phenol--Water system was not found, guesses about the water to solvent ratio had to be made based on studies of similar systems. One recent study by Medzadourian, Durlofsky and Thomason²⁷ involved an extraction study with MIBK as the solvent. In the study, water to solvent ratios of five and ten to one were investigated. Since NBA and MIBK have similar distributions with respect to phenol, both of these ratios were considered. The initial tests using ASPEN with these ratios indicated that removal efficiencies were essentially the same for both. It was surmized that at ratios much greater than ten, a physical limitation of being able to properly disperse the solvent within the feed phase would be encountered. Since operating costs would be lower for a water to solvent ratio of ten rather than five due to the decrease in the amount of heat required to volatilize the solvent in the recovery tower, a water to solvent ratio of ten was selected.

It was decided to operate the column at atmospheric pressure since pressure has very little effect on liquid-liquid equilibrium. This was also advantageous since it reduced the cost of the vessel. The temperature was determined based on an assumption concerning cooling water temperature. It was desirable to operate the extraction tower at low temperatures because the solubility of NBA in water increases with temperature. Since the majority of the solvent would be recycle solvent from the recovery system, the operating temperature of the extraction

system was determined by the lowest temperature that the condensate from the primary solvent recovery tower could be cooled to without refrigeration. This temperature was limited to the temperature of the cooling water, (which was assumed to be 80 °F, plus a reasonable temperature approach of 10 °F in the solvent cooler. This resulted in assuming an operating temperature of 90 °F for the extraction tower.

The only remaining parameter needed for the simulation of the extractor was the number of equilibrium stages. This was primarily determined by the overall removal efficiency for the Phenosolvan system. Given a desired removal efficiency of around 99.5%^{25, 48}, the number of stages was adjusted by trial and error until a conservative estimate for the removal efficiency of approximately 99.3% was obtained. This resulted in 15 equilibrium stages for the extraction tower.

The primary solvent recovery system was modeled using a distillation tower. The major information needed to design the tower was the reflux ratio, the number of equilibrium stages, the bottoms or overhead rate and the operating pressure. It was found that there was no advantage to operating the tower at pressures above atmospheric so a pressure of 14.7 psia was assumed. Initial values for the remaining parameters were determined using a short-cut distillation routine in ASPEN called DSTWU which employs the Fenske-Winn-Underwood design equations.^{21, 28, 33, 50} The input required for this model were the separation efficiencies for the light and heavy keys in the feed and a q value for the feed which indicates the feed thermal condition, (i.e. a subcooled liquid, saturated liquid, saturated vapor or superheated vapor). Given this information, the routine calculated the number of

equilibrium stages and the minimum and actual reflux ratios. An initial guess for the phenol product purity of 95% was assumed for this calculation.

The information obtained from this calculation was then used as input to a more rigorous distillation design routine in ASPEN called RADFRC. This routine required the number of equilibrium stages, the reflux ratio and the bottoms or overhead production rates. When the distillation tower was coupled with the solvent extraction tower, it was found that the initial parameters obtained from DSTWU had to be changed somewhat. This was due to the sensitivity of the overall removal efficiency to the amount of phenol distilled in the primary solvent recovery tower. The amount of phenol removed in the tower bottoms significantly affected the quantity of phenol in the column overhead and thus the phenol in the recycle loop between the extraction tower and the solvent recovery tower. This in turn affected the ability of the solvent to remove from the wastewater stream because solvent was already partially laden with phenol.

Since the total amount of phenol in the feed was 123.97 lbmole/hr, the total bottoms rate from the recovery tower was taken to be 123.8 lbmole/hr assuming that most of the phenol was extracted from the waste stream and that it could be separated from the solvent to a significant degree. The reflux ratio was set to one based on the calculations taken from the DSTWU simulation. The number of equilibrium stages and the feed location were adjusted in order to increase the overall removal efficiency to the desired level. It was found that the stripping section of the tower was much more critical than the enriching section in terms

of phenol removal. This resulted in moving the feed location towards the top of the tower in order to allow more stripping stages. These adjustments resulted in a total of 30 equilibrium stages with the feed located 10 stages from the top of the tower.

A similar procedure was used to evaluate the design parameters in the secondary recovery system. The operating pressure of the tower was increased to 60 psia in order to reduce the number of pumps required in the secondary recovery system. The criterion for this tower was to recover as much of the solvent as possible without volatilizing significant quantities of water. This was accomplished by increasing the number of stages in the stripping section of the tower until the additional stages provided no significant decrease in the amount of solvent lost to the tower bottoms. These adjustments resulted in setting the number of theoretical stages equal to 10 with the feed located four stages from the top of the tower. The reflux ratio was set to one and the bottoms rate was set to 91500 lbmoles/hr.

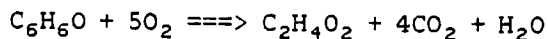
Wet Air Oxidation process. The primary focus of interest in the WAO process was in the design of the oxidation reactor. The major parameters needed to simulate a reactor with most of the ASPEN reactor models are the operating temperature, pressure and stoichiometry and extent, (i.e. conversion efficiencies), of the reactions.

The operating temperature was selected based on the types of material to be oxidized. Some contaminants are much more difficult to oxidize than others as is shown in Tables 3 and 4 on pages 15 and 16. Since many different organic materials may appear in the wastewater after extraction, it was felt that the operating conditions of the reactor

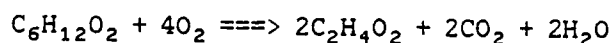
should reflect the possibility of having to remove materials which are difficult to oxidize. In reviewing the literature reports on contaminant removal efficiencies for WAO, it was observed that the majority of compounds had removal efficiencies of 99% or better at temperatures approaching 600 °F.^{9,37} Based on this review, 580 °F was assumed to be the reactor temperature.

The pressure of the reactor is directly related to the reactor temperature. To maintain the water in the liquid state, the reactor pressure must be at least the saturation pressure for water at the given temperature. In addition to this, an over-pressure is required in order to decrease the volatilization of water caused by the effect of the volatile gases produced from the reactions and to increase the concentration of oxygen in the liquid phase to improve reaction rates. At 580 °F, the saturation pressure is 1326.1 psia. This led to selection of 1580 psia as the operating pressure of the reactor.

The next set of parameters needed is the stoichiometry of the oxidation reactions. To determine this, one must first determine the reaction products for a given reactant. The major end products of WAO are assumed for this study to be CO₂, H₂O, and acetic acid, (C₂H₄O₂) based on information presented by Randall³⁸ and Baillod⁴. Given this assumption, the stoichiometric equation for phenol was taken as



A similar reaction equation was assumed for the residual NBA which was soluble in the wastewater. The reaction equation assumed was



Conversion efficiencies of 99.5% were assumed for both of these reactions.

Both of these reactions are exothermic but the concentration of phenol and NBA are so low that the energy required to heat the water up to the reactor temperature is much greater than the energy gained by the heat of reaction. To minimize the amount of energy required, the reactor effluent was used to heat up the incoming wastewater. An approach temperature of 30 °F was assumed for this exchanger to try to reduce the heat transfer area required. The addition heat required to maintain the reactor temperature was gained by employing a fired furnace using natural gas.

IV. ASPEN SIMULATION PROGRAMS FOR THE WATER TREATMENT PLANT

Mass and Energy Balance Simulation

The ASPEN simulation program for the block diagram shown in Figure 5 on page 36 is tabulated in Figure 7. What follows is a description of the simulation techniques employed in the program.

The first section of the program involves the simulation of the physical properties which were used. As stated before, SYSOP11 and SYSOP5 were used to simulate the flowsheet sections SOLVAN and WETAIR respectively. The pure component property parameters for H_2O , NBA, C_6H_6O , O_2 , $C_2H_4O_2$ and CO_2 were recalled from the ASPEN's physical-property data base to use in the declared option sets. In addition, O_2 and CO_2 were declared as super-critical components and their chemical formulae were specified to allow the use of special parameters in the property models.

The next section of the program describes the equipment flowsheet and material streams in the waste treatment plant. The flowsheet was partitioned into two sections to allow the use of different physical-property options for each section. The flowsheet section SOLVAN contains the primary equipment involved in the Phenosolvan simulation. The flowsheet WETAIR contains the equipment used to simulate the WAO process. In addition to this, information about the calculation sequence had to be supplied for this simulation. This was done because the calculation sequence generated by ASPEN's sequence generator was impractical for this simulation and would not converge. To implement the specified calculation sequence, information identifying the tear streams

```

NEW
IN-UNITS ENG
OUT-UNITS ENG
RUN-CONTROL MAX-TIME=600
TITLE 'PHENOSOLVAN -- WET-AIR OXIDATION SIMULATION'
DESCRIPTION " THIS IS A SIMULATION PROGRAM FOR A COMBINED PHENOSLOVAN
      WET-AIR OXIDATION PROCESS USING PHENOL CONTAMINATED WASTE WATER
      FROM A 20,000 TON/DAY COAL SYN-FUELS PLANT."
HISTORY MSG-LEVEL PROPERTIES=2
;
;*****
;*
;*          PHYSICAL PROPERTY AND COMPONENT LIST          *
;*
;*****
;
PROPERTIES SYSOP11 SOLVAN/SYSOP5 WETAIR
COMPONENTS H2O WATER/NBA C6H12O2-1/C6H6O PHENOL/O2 O2/C2H4O2 C2H4O2-1 &
      /CO2 CO2
SC-COMPS SCCOMP O2 CO2
FORMULA O2 O2 /CO2 CO2
;
;*****
;*
;*          FLOWSHEET SECTION          *
;*
;*****
;
FLOWSHEET SOLVAN
BLOCK P1          IN= AQFEED          OUT= S101
BLOCK EXTRACT     IN= S202 S101        OUT= S102 S203
BLOCK P3          IN= S102             OUT= S103
BLOCK PREHT2      IN= S103 Q2          OUT= S104
BLOCK SOLVR2      IN= S104             OUT= S105 S106
BLOCK COOL2       IN= S105             OUT= S107
BLOCK SEP         IN= S107             OUT= VAP S108 S211
BLOCK SMIX2       IN= S108 S106        OUT= S109
BLOCK HEATX2      IN= S109             OUT= S110
BLOCK P5          IN= S110             OUT= S301
BLOCK P2          IN= SOLV             OUT= S201
BLOCK SMIX1       IN= S201 S210 S211   OUT= S202
BLOCK HFLASH      IN= S203             OUT= VAP2 S204
BLOCK P6          IN= S204             OUT= S205
BLOCK PREHT1      IN= S205 Q1          OUT= S206
BLOCK SOLVR       IN= S206             OUT= S207 PHEN
BLOCK P4          IN= S207             OUT= S208
BLOCK HEATX       IN= S208             OUT= S209 Q1
BLOCK COOL        IN= S209             OUT= S210

```

Figure 7. ASPEN Mass and Energy Balance Simulation Program

```

FLOWSHEET   WETAIR
BLOCK COMP      IN= OXSUP              OUT= S302
BLOCK OXMIX      IN= S301 S302          OUT= S303
BLOCK OXPREHT    IN= S305 S303          OUT= S306 S304
BLOCK OXREACT    IN= S304              OUT= S305
BLOCK GASSEP     IN= S306              OUT= VENT S307
BLOCK COOL3      IN= S307              OUT= DISCHARGE
;
;*****
;*
;*      LOOP CONVERGENCE AND SEQUENCE SPECIFICATIONS      *
;*
;*****
;
CONVERGENCE CA WEGSTEIN
  TEAR S202
;
CONVERGENCE APR WEGSTEIN
  TEAR Q1
;
CONVERGENCE AOX WEGSTEIN
  TEAR S304
;
CONVERGENCE ANB ONE-VAR
  SPEC NBAIN
;
CONVERGENCE BPR WEGSTEIN
  TEAR Q2
;
SEQUENCE FLOW P1 CA ANB P2 SMIX1 (RETURN ANB) EXTRACT P3 BPR PREHT2 &
  SOLVR2 COOL2 SEP SMIX2 HEATX2 (RETURN BPR) HFLASH P6 &
  APR PREHT1 SOLVR P4 HEATX (RETURN APR) COOL (RETURN CA) P5 OXIN &
  COMP OXMIX AOX OXREACT OXPREHT (RETURN AOX) GASSEP COOL3
;
;*****
;*
;*      USER SUPPLIED PHYSICAL PROPERTIES      *
;*
;*****
;
;      UNIQUAC BINARY INTERACION PARAMETERS
;
PROP-DATA
  PROP-LIST
    BPVAL  H2O      NBA      GMUQA  /  GMUQB
    BPVAL  NBA      H2O      0.0    /  -169.6
    BPVAL  NBA      C6H6O    0.0    /  -499.4
    BPVAL  NBA      C6H6O    0.0    /  150.3

```

Figure 7 (Continued)

```

BPVAL  C6H6O  NBA          0.0    /  -22.4
BPVAL  C6H6O  H2O          0.0    /   71.0
BPVAL  H2O    C6H6O        0.0    / -213.5
;
;  UNIQUAC PURE COMPONENT PARAMETERS
;
PROP-LIST                      GMUQR    /  GMUQQ
PVAL   H2O                    0.9200   /  1.4000
PVAL   NBA                    4.8274   /  4.1960
PVAL   C6H6O                  3.5517   /  2.6800
;
;  INPUT CORRECTION TO DATA BANK VALUES FOR PHENOL.
;
PROP-LIST                      VC /  ZC
PVAL   C6H6O                  .264 /  .287
;
;*****
;*
;*  INPUT SOURCE STREAMS AND INITIAL GUESSES FOR TEAR STREAMS  *
;*
;*****
;
DEF-STREAMS HEAT Q1 Q2
;
;***  TEAR STREAM INITIAL GUESSES
;
STREAM Q1
  STREAM-ATTR HEAT DUTY=.78348D7
;
STREAM Q2
  STREAM-ATTR HEAT DUTY= 3.7032D8
;
STREAM S202  TEMP=90 PRES=30
  MOLE-FLOW H2O 157.3350/NBA 1427.2145/C6H6O 2.5912
;
STREAM S210  TEMP=90 PRES=40
  MOLE-FLOW H2O 152.1318/NBA 1353.8975/C6H6O 2.5736
;
STREAM S211  TEMP=90 PRES=40
  MOLE-FLOW H2O 5.217/NBA 72.27/C6H6O .0174
;
STREAM S304  TEMP=560 PRES=1615
  MOLE-FLOW H2O 91869.5926/NBA .321/C6H6O .8948/O2 7.485
;
;***  FEED STREAM SPECIFICATIONS
;

```

Figure 7 (Continued)

```

; ASSUME A RATIO OF 1LB H2O PRODUCED PER 1LB OF COAL
;
STREAM AQFEED TEMP=90 PRES=14.7
  MOLE-FLOW C6H6O 123.9652/H2O 91868.9078
;
STREAM SOLV TEMP=80 PRES=14.7
  MOLE-FLOW NBA 1.0529
;
STREAM OXSUP TEMP=80 PRES=14.7
  MOLE-FLOW O2 7.485
;
;*****
;*
;* INPUT EQUIPMENT PARAMETERS FOR THE SOLVAN FLOWSHEET *
;*
;*****
;
BLOCK P1 PUMP
  PARAM PRES=30
;
F-BLOCK EXTRACT EXTRC
  PARAM 1 15 0 0 0 50 0 90 90 14.7 14.7 1.D-3
  BLOCK-OPTIONS HMB-RESULTS=2
;
BLOCK P3 PUMP
  PARAM PRES=80
;
BLOCK PREHT2 HEATER
  PARAM PRES=-10
;
BLOCK SOLVR2 RADFR3
  PARAM 10 ALGOPT=2
  FEEDS S104 4
  PRODUCTS S105 1 0/S106 10 0
  COL-SPECS RDV=0.0 RR=1.00 B=91500.0
  P-SPEC 1 60/ 10 60
;
CBLOCK TOWER2 COST-TRAYTWR
  HMB-REFERENCE SOLVR2
  PARAM CONDENSER=1 REBOILER=1 NSCT=1 NCND=1 NRBL=1
  SINP NFD= 1 * * * PRES=60 CNDU=200 REBU=300 CTIN=80 CTOT=100 &
    ACCV=1000 COND-AREA=468.705 REBOIL-AREA=713.557 RTIN=445
;
BLOCK COOL2 HEATER
  PARAM PRES=-10 TEMP=90

BLOCK SEP FLASH3
  PARAM PRES=-10 Q=0 KEY=H2O L-KODE=1

```

Figure 7 (Continued)

```

;
BLOCK SMIX2 MIXER
  PARAM PRES=-10 NPK=1 KPH=2
;
BLOCK HEATX2 HEATER
  PARAM PRES=-10 TEMP=110
;
BLOCK P5 PUMP
  PARAM PRES=1635
;
BLOCK P2 PUMP
  PARAM PRES=40
;
BLOCK SMIX1 MIXER
  PARAM PRES=-10 NPK=1 KPH=2
;
; ADDED FLASH TO ELIMINATE PROBLEM OF FLOWTRAN BLOCK NOT CALCULATING
; OUTLET STREAM PROPERTIES.
;
BLOCK HFLASH FLASH2
  PARAM PRES=14.7 Q=0.00
;
BLOCK P6 PUMP
  PARAM PRES=45
;
BLOCK P4 PUMP
  PARAM PRES=60
;
BLOCK PREHT1 HEATER
  PARAM PRES=-10
;
BLOCK SOLVR RADFR3
  PARAM 30 ALGOPT=2
  FEEDS S206 10
  PRODUCTS S207 1 0/PHEN 30 0
  COL-SPECS RDV=0.0 RR=1.00 B=123.8
  P-SPEC 1 14.7/ 30 14.7
  T-EST 1 210/ 30 360
;
CBLOCK TOWER1 COST-TRAYTWR
  HMB-REFERENCE SOLVR
  PARAM CONDENSER=1 REBOILER=1 NSCT=1 NCND=1 NRBL=1
  SINP NFD= 1 * * * PRES=14.7 CNDU=200 REBU=300 CTIN=80 CTOT=100 &
    ACCV=1000 COND-AREA=2636.2 REBOIL-AREA=2292.3 RTIN=445
;
BLOCK HEATX HEATER
  PARAM PRES=-10 TEMP=100

```

Figure 7 (Continued)

```

;
BLOCK COOL HEATER
  PARAM PRES=-10 TEMP=90
;
;*****
;*
;* INPUT EQUIPMET PARAMETERS FOR THE WETAIR FLOWSHEET *
;*
;*****
;
BLOCK COMP COMPR
  PARAM 2 1635
;
BLOCK OXMIX MIXER
  PARAM PRES=-10
;
BLOCK OXPREHT HEATX
  PARAM 2 0 TCOLD=550 DPH=-10 DPC=-10
;
BLOCK OXREACT RSTOIC
  PARAM TEMP=580 PRES=-15
  STOICH 1 MIXED C6H6O -1/02 -5/H2O 1/C2H4O2 1/CO2 4
  STOICH 2 MIXED NBA -1/02 -4/C2H4O2 2/H2O 2/CO2 2
  CONV 1 MIXED C6H6O .995
  CONV 2 MIXED NBA .995
;
BLOCK GASSEP FLASH2
  PARAM PRES=25
;
BLOCK COOL3 HEATER
  PARAM PRES=14.7 TEMP=100
;
;*****
;*
;* INPUT SECTION FOR DESIGN SPECIFICATIONS AND FORTRAN BLOCKS *
;*
;*****
;
; DESIGN SPECIFICATION TO CONTROL THE WATER TO SOLVENT RATIO
;
DES-SPEC NBAIN
  DEFINE A1IN MASS-FLOW STREAM=AQFEED COMPONENT=H2O
  DEFINE A2IN MASS-FLOW STREAM=S202 COMPONENT=H2O
  DEFINE SOUT MASS-FLOW STREAM=S202 COMPONENT=NBA
F  RATIO=((A1IN + A2IN)/SOUT)/10
  SPEC RATIO TO 1.00
  TOL-SPEC 1.D-5
  VARY MOLE-FLOW STREAM=SOLV COMPONENT=NBA

```

Figure 7 (Continued)

```

LIMITS 0 10
;
;   FORTRAN BLOCK TO CALCULATE THE OXYGEN NEEDED IN THE REACTOR
;
FORTRAN OXIN
  DEFINE PIN    MOLE-FLOW STREAM=S301  COMPONENT=C6H6O
  DEFINE FNBA   MOLE-FLOW STREAM=S301  COMPONENT=NBA
  DEFINE ANEED  MOLE-FLOW STREAM=OXSUP  COMPONENT=O2
F    ANEED= 1.3*(PIN*5 +FNBA*4)
  EXECUTE BEFORE OXMIX
;
SIMULATE

```

Figure 7 (Continued)

for the convergence loops and the convergence methods for these loops was specified.

The physical-property section of the program was used to enter the necessary parameters required by the UNIQUAC equation and the corrections for the phenol critical constants. The parameters entered here include the UNIQUAC binary interaction parameters GMUQA and GMUQB which were obtained by regression using PATTERN. GMUQA was set equal to zero so that the one-parameter model for the UNIQUAC equation was used rather than the two-parameter model. Also entered were the pure component van der Waals volume and surface area parameters GMUQR and GMUQQ.

The stream section of the program was used to enter initial guesses for the convergence loop tear streams to reduce the calculation time and to define all material feed streams. Streams Q1 and Q2 were defined as heat streams to allow the passage of energy rather than material through these streams. The initial guesses for all of the convergence streams were obtained from previous calculations made by the program.

The equipment section of the program contains the specifications for each piece of equipment in the process. For most of the equipment listed, the specifications required are the operating pressure and/or temperature. Negative values given for pressure represent pressure drops across the equipment. All other pressures shown are specified in psia. All of the other parameters specified are given in standard engineering units.

Certain simulation techniques were employed in order to reduce possible convergence problems and eliminate calculation errors. One

such technique involved the simulation of feed preheaters as two general heaters with a heat stream flowing between them. This was done to improve the convergence. The advantage of doing this over using the heat-exchanger model, HEATX, in ASPEN is the separation of the convergence loop which is formed between the stream being heated, the stream supplying the heat and the energy flowing from one stream to the other. In the HEATX model, the energy stream flowing between the material streams is an internal stream and cannot be used as a tearing point for the convergence loop. This may result in slow or no convergence within the loop.

Another aspect of the equipment flowsheet which requires some explanation is the addition of an equipment block called HFLASH. This block is a result of a problem which was encountered when the FLOWTRAN extractor model was invoked. This extractor model was used because the ASPEN extractor model was not completed by MIT. The FLOWTRAN routine, however, was not properly interfaced with the ASPEN simulation models. The result was that the density and entropy of the outlet material streams were not calculated. This caused an error to occur when the outlet streams were used as input to a unit operation model which did not perform a flash on the inlet stream. To resolve the error, the two-phase flash model of ASPEN was used to calculate the necessary properties of the stream.

The final section of the simulation program defines the design specifications and FORTRAN blocks which allow manipulation of stream or block parameters to achieve a specific goal. In this simulation, both a design specification and a FORTRAN block were used.

The design specification was used in order to set the water to solvent weight ratio in the extraction equal to ten. This was achieved by retrieving the amount by weight of water entering the extraction tower from the wastewater feed and solvent recycle streams and, the amount by weight of solvent entering the tower from recycle and solvent makeup streams. The water to solvent weight ration was then calculated and divided by 10. This was done to try and make the specification value equal to one, allowing greater accuracy. The resulting value was then compared to the desired value of one and adjustments were made to the solvent makeup stream until the calculated ratio was within the specified tolerance of the specified ratio.

The FORTRAN block, OXIN, was used to calculate the amount of oxygen needed for the WAO reactions. Since the amounts of phenol and NBA entering the reactor were known from the SOLVAN flowsheet calculations, the amount of oxygen needed was calculated based on the stoichiometry of the reaction equations. In order to insure that enough oxygen exists in the reactor for the reactions, a 30% excess in stoichiometry was assumed. The resulting quantity, ANEED, was then defined as the amount of oxygen in the supply stream OXSUP.

The results of the simulation program were returned from ASPEN in a tabulated report section. This report is shown in Appendix B. The concentration of phenol leaving the Phenosolvan section of the flowsheet was around 18 ppm. This phenol was further treated in the WAO section to produce a phenol discharge rate of 0.0044 lbmoles/hr,

(0.41 lb/hr), at a concentration of 0.25 ppm in the discharge stream. Since the removal efficiencies for phenol in both the Phenosolvan and the

WAO steps of the process were greater than 99%, it was felt that this value represented the lowest reasonable concentration that could be obtained without additional treatment. This value, however, is still greater than the desired limit of 0.13 ppm.

Capital Cost Simulation

Once the mass and energy balance for the treatment process was completed, a capital cost estimate was performed. This was accomplished using ASPEN's sizing and cost estimation routines. The ASPEN simulation program for this is shown in Figure 8. To obtain the simulation program, many assumptions had to be made for a wide range of material and economic factors which affect the overall capital cost.

Purchased Equipment Cost Simulation Parameters

To evaluate the purchased equipment cost for each piece of equipment, it was necessary to determine the parameters required by ASPEN's size and cost estimation routines. Much of this information was directly obtained from the mass and energy balance simulation. However, parameters such as material of construction, heat-transfer coefficients, cooling water inlet and outlet temperatures, residence times and in some cases the actual cost of the equipment were not available from previous calculations and had to be determined by other means.

In general, the material of construction was taken to be carbon steel. It was judged that the wastewater streams were not very corrosive since their composition was mainly water at low temperatures and pressures and a pH of around 8. There were some exceptions to this, however, where it was believed that stainless steel would be beneficial

```

NEW COST
RUN-CONTROL MAX-TIME=300
UTILITY ELECT ELECTRICITY
    PRICE PRICE=.05 [$/KWHR] INDX=286 TYPE=2
    GENORIMPORT FLAG=2
;
UTILITY HSTEAM STEAM
    PRICE PRICE=.0056 [$/LB] INDX=286 TYPE=2
    PARAM CAP=780.4 [BTU/LB] PRES=400
    GENORIMPORT FLAG=1
;
UTILITY COLD WATER
    PRICE PRICE=.024 [$/TON] PARAM CAP=1 [BTU/LB] TEMP=80 RTEMP=100
    GENORIMPORT FLAG=1
;
RAW-MATERIAL NBA
    FLOW-RATE MASS-FLOW=122.305 [LB/HR] DENSITY= 54.444 [LB/CUFT]
    PRICE PRICE=.50 [$/LB] INDX=286 TYPE=2
    PHASE CODE=2
;
PRODUCT PHENOL
    FLOW-RATE MASS-FLOW=11667.25 [LB/HR] DENSITY= 57.639 [LB/CUFT]
    PRICE PRICE= .29 [$/LB] PHASE CODE=2
;
CBLOCK PUMP1 COST-PUMP
    PARAM NSB=1 PTYP=4 MTYP=2 SRSY=2 MAT=1
    SIZE-INPUT VFW=27042.3 DENS=61.63 PRS=15.3
    UTILITY ELECT
    INST-MAT-FAC PIPF=.302 CONF=.04 INTF=.03 ELEF=.31 INSF=.025 &
        PNTF=.008
    INST-LAB-FAC OVRF=.071
;
CBLOCK PUMP2 COST-PUMP
    PARAM NSB=1 PTYP=4 MTYP=2 SRSY=2 MAT=1
    SIZE-INPUT VFW=2.2474 DENS=54.44 PRS=25.3
    UTILITY ELECT
    INST-MAT-FAC PIPF=.302 CONF=.04 INTF=.03 ELEF=.31 INSF=.025 &
        PNTF=.008
    INST-LAB-FAC OVRF=.071
;
CBLOCK PUMP3 COST-PUMP
    PARAM NSB=1 PTYP=4 MTYP=2 SRSY=2 MAT=1
    SIZE-INPUT VFW=27042.6 DENS=61.52 PRS=65.3
    UTILITY ELECT
    INST-MAT-FAC PIPF=.302 CONF=.04 INTF=.03 ELEF=.31 INSF=.025 &
        PNTF=.008
    INST-LAB-FAC OVRF=.071
;

```

Figure 8. ASPEN Capital Investment Simulation Program

```

CBLOCK PUMP4 COST-PUMP
  PARAM NSB=1 PTYP=4 MTYP=2 SRSY=2 MAT=1
  SIZE-INPUT VFW=3159.47 DENS=50.72 PRS=45.3
  UTILITY ELECT
  INST-MAT-FAC PIPF=.302 CONF=.04 INTF=.03 ELEF=.31 INSF=.025 &
    PNTF=.008
  INST-LAB-FAC OVRF=.071
;
CBLOCK PUMP5A COST-PUMP
  PARAM NSB=1 PTYP=6 MTYP=2 SRSY=2 MAT=1
  SIZE-INPUT VFW=27187.7 DENS=60.878 PRS=540.0
  UTILITY ELECT
  INST-MAT-FAC PIPF=.302 CONF=.04 INTF=.03 ELEF=.31 INSF=.025 &
    PNTF=.008
  INST-LAB-FAC OVRF=.071
;
CBLOCK PUMP5B COST-PUMP
  PARAM NSB=0 PTYP=6 MTYP=2 SRSY=2 MAT=1
  SIZE-INPUT VFW=27187.7 DENS=60.878 PRS=540.0
  UTILITY ELECT
  INST-MAT-FAC PIPF=.302 CONF=.04 INTF=.03 ELEF=.31 INSF=.025 &
    PNTF=.008
  INST-LAB-FAC OVRF=.071
;
CBLOCK PUMP5C COST-PUMP
  PARAM NSB=0 PTYP=6 MTYP=2 SRSY=2 MAT=1
  SIZE-INPUT VFW=27187.7 DENS=60.878 PRS=535.0
  UTILITY ELECT
  INST-MAT-FAC PIPF=.302 CONF=.04 INTF=.03 ELEF=.31 INSF=.025 &
    PNTF=.008
  INST-LAB-FAC OVRF=.071
;
CBLOCK PUMP6 COST-PUMP
  PARAM NSB=1 PTYP=4 MTYP=2 SRSY=2 MAT=1
  SIZE-INPUT VFW=3114.77 DENS=55.20 PRS=30.3
  UTILITY ELECT
  INST-MAT-FAC PIPF=.302 CONF=.04 INTF=.03 ELEF=.31 INSF=.025 &
    PNTF=.008
  INST-LAB-FAC OVRF=.071
;
CBLOCK COMP COST-COMPRESS
  PARAM NSB=0 CTYP=2
  SIZE-INPUT TVF=2993.24 GHP=59.6543
  COST-INPUT BCST=42507 PCST=54834 INDX=340.4 TYPE=2
  UTILITY ELECT
  INST-MAT-FAC PIPF=.18 CONF=.10 INTF=.04 ELEF=.02 INSF=.000
  INST-LAB-FAC OVRF=.071

```

Figure 8 (Continued)

```

;
; HEAT EXCHANGER FOR TOWER (PREHT1-HEATX)
CBLOCK HEATX1 COST-HEATEX
  PARAM NSB=0 TYPE=1 SMAT=1 TMAT=2 SPRS=35 TPRS=50
  SIZE-INPUT HDTY=7.831D6 HTRC=150 TI1=90.314 TO1=183.3778 &
    TI2=198.2786 TO2=100 NTPS=1
  INST-MAT-FAC PIPF=.456 CONF=.051 INTF=.102 STLf=.031 ELEF=.02 &
    INSF=.049 PNTF=.005
  INST-LAB-FAC OVRF=.071
;
; HEAT EXCHANGER FOR OXPREHT
CBLOCK HEATX2 COST-HEATEX
  PARAM NSB=0 TYPE=1 SMAT=1 TMAT=2 SPRS=1600 TPRS=1625
  SIZE-INPUT HDTY=8.13159D8 HTRC=200 TI1=580 TO1=201.99 TI2=167.738 &
    TO2=550 NTPS=1
  INST-MAT-FAC PIPF=.456 CONF=.051 INTF=.102 STLf=.031 ELEF=.02 &
    INSF=.049 PNTF=.005
  INST-LAB-FAC OVRF=.071
;
; HEAT EXHANGER FOR COOL
CBLOCK HEATX3 COST-HEATEX
  PARAM NSB=0 TYPE=1 SMAT=1 TMAT=1 SPRS=30 TPRS=50
  SIZE-INPUT HDTY=7.5767D5 HTRC=100 TI1=80 TO1=90 TI2=100 TO2=90 NTPS=1
  UTILITY WATER=COLD
  INST-MAT-FAC PIPF=.456 CONF=.051 INTF=.102 STLf=.031 ELEF=.02 &
    INSF=.049 PNTF=.005
  INST-LAB-FAC OVRF=.071
;
; HEAT EXHANGER FOR COOL2
CBLOCK HEATX4 COST-HEATEX
  PARAM NSB=0 TYPE=1 SMAT=1 TMAT=1 SPRS=30 TPRS=50
  SIZE-INPUT HDTY=1.9473D6 HTRC=100 TI1=80 TO1=100 TI2=252.8084 &
    TO2=90 NTPS=1
  UTILITY WATER=COLD
  INST-MAT-FAC PIPF=.456 CONF=.051 INTF=.102 STLf=.031 ELEF=.02 &
    INSF=.049 PNTF=.005
  INST-LAB-FAC OVRF=.071
;
; HEAT EXCHANGER FOR COOL3
CBLOCK HEATX5 COST-HEATEX
  PARAM NSB=0 TYPE=1 SMAT=1 TMAT=1 SPRS=30 TPRS=25
  SIZE-INPUT HDTY=2.1288D8 HTRC=150 TI1=80 TO1=100 TI2=204.3494 &
    TO2=100 NTPS=1
  UTILITY WATER=COLD
  INST-MAT-FAC PIPF=.456 CONF=.051 INTF=.102 STLf=.031 ELEF=.02 &
    INSF=.049 PNTF=.005
  INST-LAB-FAC OVRF=.071

```

Figure 8 (Continued)

```

; HEAT EXCHANGER FOR TOWER (PREHT2-HEATX2)
CBLOCK HEATX6 COST-HEATEX
  PARAM NSB=0 TYPE=1 SMAT=1 TMAT=2 SPRS=30 TPRS=80
  SIZE-INPUT HDTY=3.7032D8 HTRC=150 TI1=90.2019 TO1=283.7816 &
    TI2=292.1773 TO2=110 NTPS=1
  INST-MAT-FAC PIPF=.456 CONF=.051 INTF=.102 STLF=.031 ELEF=.02 &
    INSF=.049 PNTF=.005
  INST-LAB-FAC OVRF=.071
;
;
CBLOCK FURNACE COST-FURNACE
  PARAM TYPE=1 NSB=0 MAT=2
  SIZE-INPUT QTOT=7.1997D7 PRES=1610
  INST-MAT-FAC PIPF=.18 CONF=.10 INTF=.04 ELEF=.02
  INST-LAB-FAC OVRF=.071
;
CBLOCK REACT COST-VERVSSL
  PARAM NVS=6 NSB=0 MAT=1 MFCT=2.25
  SIZE-INPUT DIA=12 LGTH=75 PRES=1600
  INST-MAT-FAC PIPF=.596 CONF=.099 STLF=.079 INTF=.115 ELEF=.049 &
    INSF=.08 PNTF=.013
  INST-LAB-FAC OVRF=.071
;
;
CBLOCK EXTRACT COST-VERVSSL
  PARAM NVS=3 NSB=0 MAT=2 AFCT=3.8236
  SIZE-INPUT DIA=11 LGTH=30 PRES=14.7
  INST-MAT-FAC PIPF=.60 CONF=.10 STLF=.08 INTF=.115 ELEF=.05 &
    INSF=.08 PNTF=.013
  INST-LAB-FAC OVRF=.071
;
;
CBLOCK SEP COST-VERVSSL
  PARAM NVS=1 NSB=0 MAT=1
  SIZE-INPUT DIA=2 LGTH=7 PRES=40
  INST-MAT-FAC PIPF=.60 CONF=.10 STLF=.08 INTF=.115 ELEF=.05 &
    INSF=.08 PNTF=.013
  INST-LAB-FAC OVRF=.071
;
;
CBLOCK GASSEP COST-VERVSSL
  PARAM NVS=1 NSB=0 MAT=1
  SIZE-INPUT DIA=8 LGTH=28 PRES=25
  INST-MAT-FAC PIPF=.60 CONF=.10 STLF=.08 INTF=.115 ELEF=.05 &
    INSF=.08 PNTF=.013
  INST-LAB-FAC OVRF=.071

```

Figure 8 (Continued)

```

;
CBLOCK TOWER1 COST-TRAYTWR
  COST-INPUT BCST=5.113D5 PCST=8.617D5 INDX=257.4 TYPE=2
  INST-MAT-FAC PIPF=.596 CONF=.099 STLf=.079 INTF=.115 ELEF=.049 &
  INSF=.08 PNTF=.013
  INST-LAB-FAC OVRF=.071
;
CBLOCK TOWER2 COST-TRAYTWR
  COST-INPUT BCST=1.163D5 PCST=1.774D5 INDX=257.4 TYPE=2
  INST-MAT-FAC PIPF=.60 CONF=.10 STLf=.08 INTF=.115 ELEF=.05 &
  INSF=.08 PNTF=.013
  INST-LAB-FAC OVRF=.071
;
COST-SECTION PHSOLVAN
  DESCRIPTION 'COST SUMMARY FOR THE PHENOLSOLVAN SECTION OF THE PLANT'
  BLOCKS PUMP1 PUMP2 PUMP3 PUMP4 PUMP6 HEATX1 HEATX3 HEATX4 &
    TOWER1 TOWER2 SEP EXTRACT
  BLDG-MAT      FACTOR= 0.00
  BLDG-LAB      FACTOR= 0.00
  TESTING-MAT   FACTOR= 0.001
  TESTING-LAB   FACTOR= 0.002
  EQUIPMENT     UNLISTED= 0.05 SPARES= 0.01
;
COST-SECTION WETAIR
  DESCRIPTION 'COST SUMMARY FOR THE WAO SECTION OF THE PLANT'
  BLOCKS PUMP5A PUMP5B PUMP5C COMP HEATX2 HEATX5 HEATX6 REACT GASSEP &
    FURNACE
  BLDG-MAT      FACTOR= 0.00
  BLDG-LAB      FACTOR= 0.00
  TESTING-MAT   FACTOR= 0.001
  TESTING-LAB   FACTOR= 0.002
  EQUIPMENT     UNLISTED= 0.05 SPARES= 0.01
;
COST-INDEX
  CONSTR-YEAR   1984 1
  PLANT-INDEX   321.0
  EQUIP-INDEX   340.4
  LABOR-INDEX   268.5
  INSTAL-INDEX  374.1
  RAWMAT-INDEX  286.0 ;NO CHANGE FROM 82 VALUES
;
CAPITAL-INVESTMENT ASPEN
  LABOR RATE=15 [$/HR]
  INVENTORY RMD= 30.0 PRD= 15.0
  LIQ-STORE STYP=3
  SERV-BLDG MCF= 0.075 LHF= 0.03
  SERV-SYS MCF=.06 LHF=.02
  SITE-DEVELOPMENT MCF= 0.01 LHF= 0.002

```

Figure 8 (Continued)

CONTINGENCY PRCS= 0.1 PRJD= 0.1
LAND FACT= 0.01
FREIGHT FACT= 0.02
INDIRECTS FMCf= 0.27 FLCF= 0.50 PRJM= 0.06 ENCF= 0.05 FECF= 0.08
;
ECONOMICS
DESCRIPTION 'WASTEWATER TREATMENT FOR A COAL CONVERSION FACILITY'
EXECUTION 1 1

Figure 8 (Continued)

or required. This included the heat exchanger tubes for all of the preheaters in the process, (i.e. SOLVR preheater HEATX1, SOLVR2 preheater HEATX6 and the oxidation reactor preheater HEATX2), the extraction tower and the oxidation reactor. This was decided because the elevated temperatures that occur in most of these pieces of equipment could led to stress and corrosion problems. It was assumed that stainless-steel cladding in the WAO reactor would be adequate for corrosion resistance rather than using all stainless-steel construction. The vessel itself was costed as a carbon-steel vessel. The resulting cost was multiplied by a material of construction factor of 2.25 obtained from Guthrie¹⁹ to account for the stainless-steel cladding.

Heat-transfer coefficients for the various heat exchangers used in the process ranged from 100 to 200 BTU/FT²°F based on estimates given in Perry's Chemical Engineer's Handbook.³³ Higher average temperatures in the exchanger resulted in a higher estimate for the heat-transfer coefficient. The heat-transfer coefficients for the reboilers in the distillation towers were assumed to be 300 BTU/FT²°F

The reactor vessel, the solvent-water separator in the secondary solvent recovery system and the gas separator after the oxidation reactor were sized based on a residence time for either liquid or vapor hold up. For the reactor, a residence time of 1 hour was suggested from the literature data.^{9, 37} The actual time used was 1.33 hours as a conservative estimate. A residence time of 5 minutes was selected for the solvent-water separator to allow sufficient time for phase separation. A residence time of around 3 minutes was selected for the

WAO gas separator to allow time for the vapor to separate from the liquid phase.

The solvent recovery towers were sized and costed using the heat and material balance reference mode, (HMB), in the simulation program shown in Figure 7 on page 58. Carbon-steel construction was used with bubble-cap trays and a kettle reboiler. Tray efficiencies for each tower were calculated by the tower costing model. However, due to a missing report writing routine for this cost model, the sizing and costing results were not printed in the report section. Most of the results from this routine are printed in the ASPEN history section which allows the user to see the intermediate calculation results and in general examine the progression of the simulation from the start to the end of the convergence calculations. The information obtained from this section for both solvent recovery towers is summarized in Table 11. To include the tower costs in the capital investment estimate, the costing results in Table 11 were used as input to the cost simulation program of Figure 8.

For some pieces of equipment, there were no costing routines in ASPEN or the routines produced erroneous results. This was the case for the extraction tower and the oxygen supply compressor.

The extraction tower had to be sized before a cost estimate for the equipment could be made. This was done using the sizing procedure described by Henley and Seader.²¹ After performing the necessary calculations, three extraction towers were selected with a diameter of 11 feet and a height of 30 feet. Given this information, the cost of the unit was found using data developed by Rush for Dupont.⁴⁰ This value was

Table 11

Cost and Sizing Results for the Solvent Recovery Towers

	Tower 1 (SOLVR)	Tower 2 (SOLVR2)
Diameter	15ft	11ft
Height	205ft	48ft
Vessel Costs: ^a		
Base Cost	\$ 197,700	\$ 35,930
Purchased Cost	\$ 236,400	\$ 52,510
Overhead Condenser		
Base Cost	\$ 31,130	\$ 10,190
Purchased Cost	\$ 31,130	\$ 10,190
Reboiler:		
Base Cost	\$ 28,010	\$ 12,880
Purchased Cost	\$ 28,010	\$ 12,880
Accumulator Tank:		
Base Cost	\$ 13,320	\$ 13,320
Purchased Cost	\$ 14,920	\$ 14,920
Distillate Pump:		
Number Required	1	1
Base Cost	\$ 2,054	\$ 1,055
Purchased Cost	\$ 2,288	\$ 1,215
Power Usage	26.643 hp	5.4947 hp
Bottoms Pump:		
Number Required	1	5
Base Cost	\$ 986	\$ 2,172
Purchased Cost	\$ 1,142	\$ 2,429
Power Usage	3.2247 hp	126.7783 hp
Total Tower Cost (including Trays):		
Base Cost	\$ 511,300	\$ 116,300
Purchased Cost	\$ 861,700	\$ 177,400

^aAll dollar amounts are shown in 1979 dollars.

then used to develop an adjustment factor for the vertical-vessel costing routine so that the cost of the extractor could be included in the capital cost analysis. The factor was calculated by taking the ratio of the extractor cost found from the Dupont data and the vessel cost calculated by the ASPEN cost model. The factor determined was 3.82.

The results calculated from ASPEN's compressor costing routine were not reasonable values for the oxygen supply compressor. As a result, a cost for this compressor was obtained from information in Guthrie.¹⁹ This value was then used as direct cost input to the compressor routine.

Capital Investment Simulation Parameters

Once the purchased equipment cost for the equipment was found, the overall capital investment was determined using factors to account for installation, buildings, testing laboratory, engineering costs, land, contingencies and many other capital costs.

The first step in the the evaluation of these capital cost was to estimate commodity material and labor-hour installation factors to allow calculation of the installed equipment cost. These factors can be entered into ASPEN based on the cost of individual equipment items or on the total equipment cost. Since data were available on an equipment item basis, this method was selected to improve the accuracy of the estimate. The necessary commodity material item factors were obtained from data given by Guthrie.¹⁹ The overall labor-hour factor used was calculated based on the labor factors from the same source. The value had to be calculated to convert the labor-dollar factors in Guthrie to an overall labor-hour factor. This was done by calculating the number of labor hours need to generate the total labor-dollar cost calculated by hand

from the Guthrie labor factor estimates using ASPEN's base labor rate of \$13.20/hr. Given the total labor hours, the overall labor-hour factor was calculated as a fraction of the overall material costs.

The remaining factors used in the capital investment section were either default values in ASPEN or were determined from values taken from Guthrie.²⁰ The resulting dollar values were compared to estimates obtained using different estimation methods with reasonable agreement.

The installed equipment costs were separated into two sections to allow better comparison between the costs of the Phenosolvan equipment and the costs of the WAO equipment. All of the dollar amounts were indexed to the first quarter of the construction year, 1984, using the Chemical Engineering cost indices taken from Chemical Engineering Magazine.¹²

Additional information is shown in the simulation program concerning utilities such as electricity, steam and cooling water, and raw material and product data. This information was originally supplied to allow ASPEN to calculate the operating costs of the treatment plant. This calculation could not be performed with the current version of ASPEN due to errors in the routines used to calculate the operating costs. The data, however, were left in the simulation as reference information. Raw material and product flowrates were used in the estimation to determine the capital costs associated with storage of these materials.

The results obtained from the simulation program shown in Figure 8 can be found in Appendix B. These results indicate that a capital

investment of about \$126 MM*, (1984), would be required for the combined treatment process. The Phenosolvan section equipment costs was about \$35 MM, (1984), while the WAO section was about \$91 MM, (1984).

Operating Cost Estimation

When it was determined that ASPEN could not perform the operating cost estimation due to errors in the ASPEN code, another option had to be considered. Given that most of the information needed to calculate to calculate the operating costs was readily available from the two simulations performed by ASPEN, it was decided to calculate the costs by hand.

Before the operating costs could be evaluated, the unit costs for the commodities consumed or produced had to be determined. The values required for this process were found in several literature sources and are summarized in Table 12.^{2,13,24,34,51} In some cases, the literature values were indexed to the construction year using an appropriate cost index or factor. The actual selling price for phenol was found to be between \$0.36 - 0.395 /lb.¹³ This value was reduced to \$0.29/lb because the phenol produced is a mixture of various phenols and may require further processing.²⁷ Using this information and the commodity consumption and production rates determined from the ASPEN simulations, the annual costs for each commodity were evaluated assuming 100% on-stream efficiency. These results are shown in Table 13.

* MM in the body of this report means million and M means thousand.

Table 12

Unit Costs for Commodities Consumed or Produced
in the Phenosolvan--WAO Treatment System

Commodity	Characteristics	Unit	Unit Value ^a
Steam	400 psia @445 °F	1000 lb	\$ 5.60
Cooling Water	Supplied at 80 ° F	1000 Gal	\$ 0.10
Electricity	-	Kw-hr	\$ 0.05
Oxygen	Industrial Grade	Ton	\$40.00
Natural Gas	Pipeline	1*10 ⁶ BTU	\$ 4.41
N-Butyl Acetate	-	lb	\$ 0.50
Phenol	-	lb	\$ 0.29
Labor	Construction	hr	\$15.00
Labor	Chemical	hr	\$16.00

^aAll dollar amounts are shown in 1984 dollars.

Table 13
Annual Operating Cost Summary

Commodity Consumption Costs		
Commodity	Consumption Rate	Annual Cost ^a
Steam	1,026,826 Mlb/yr	\$ 5,750,226
Cooling Water	15,136,217 MGal/yr	\$ 1,513,622
Electricity	31,892,532 Kw-hr	\$ 1,594,613
Sub-Total		\$ 8,858,461
Oxygen	1,049 Ton/yr	\$ 41,952
Natural Gas	630,694 MMBTU/yr	\$ 2,781,359
N-Butly Acetate	1,071,392 lb/yr	\$ 535,696
Sub-Total		\$ 3,359,007
Phenol Credit	102,205,110 lb/yr	\$ 29,639,482
Labor and Overhead Expenses		
Operating Labor: 6 workers @ \$16.00 /hr 3 Shifts		\$ 280,320
Supervision: 3 workers @ \$27,000 /yr 1 per Shift		\$ 81,000
Maintenance: 5% of Investment + Contingency		\$ 6,222,469
Insurance: 1% of Investment + Contingency		\$ 1,244,494
Operating Supplies: 15% of Maintenance		\$ 933,370
Laboratory: 15% of Labor + Supervision		\$ 54,198
Overhead: 70% of Maintenance+Labor+Supervision		\$ 4,608,652
Administration: 30% of Maintenance+Labor+Supervision		\$ 1,975,137
Total		\$ 15,399,640

^aAll dollar values are given in 1984 dollars.

Profitability Analyses

Once the major operating expenses had been determined, a profitability analysis was performed for the treatment system. The objective of this study was to determine the unit cost to clean up the water given a 15% return on equity investment, (ROI), and a 12% interest rate on debt. This cost can be expressed in many different ways. The cost estimates in this study were expressed in term of 1) Dollars per thousand gallons of water cleaned at a fixed phenol price of \$0.29/lb of phenol; 2) The phenol selling price in \$/lb of phenol with no water credit; 3) The incremental price increase of synthetic-natural gas, (SNG), which results from addition of the wastewater treatment process.

The cost estimates were performed using the PRPIEA profitability analysis program described in Section 2 on pages 27-32. A summary of the major parameters entered into the program is shown in Table 14. The selection of many of these parameters was based on estimates taken from several different sources.^{20,33,34}

Due to the nature of this wastewater treatment plant, there were several options available concerning the independence of the process. One could approach the problem by considering the treatment system as an independent plant and therefore develop revenues by charging a price for treating the wastewater. Another approach would be to consider the treatment process as part of a larger complex allowing the use of negative Federal income taxes.

Assuming the wastewater treatment plant was an independent entity, two cost analyses were made. For these analyses, negative Federal income

Table 14

PRPIEA Input Parameter Summary

Project Life:	20 years
Construction Period:	3 years
Start-up Year:	4
Depreciation Life:	17 years
Depreciation Method:	Sum of the Year Digits
Salvage Value:	\$ 0.0
Capital Investment:	\$ 103,707,815
Investment Schedule During Construction:	1/30% - 2/50% - 3/20%
Investment Tax Credit:	7% of Investment + Contingency
Investment Tax Credit Carry Forward:	7 years
Investment Financing:	70% Debt 30% Equity
Annual Interest Rate:	12%
Federal Income Tax Rate:	46%
Tax Loss Carry Forward:	5 Years
State Income Tax Rate:	3%
Property Tax Rate:	2% Capital Investment + Contingency
Property Insurance:	1% Capital Investment + Contingency
Plant Efficiency:	1 - 50%, 2 - 80%, 3- 90% ,Remaining Years 95%
Proportional Annual Operating Expenses: Utilities	\$ 8,858,461
Contingency:	20% of Capital Investment
Maintenance:	5% Capital Investment + Contingency
Maintenance Material/Labor Split:	50% Material, 50% Labor
Labor:	\$ 8,008,684
Start Up Costs: .01% Invest. + Cont. + 20 days Raw-Materials Supply	
Working Capital:	30 days Raw-Material & 14 days Product Storage 25 days Receivables and 1% Invest. + Cont.
Land:	\$ 1,269,679

taxes were not allowed. The first analysis done was to determine the selling price required for the water in order to recover the capital investment plus interest and provide a 15% ROI. The resulting price from this analysis was \$20.80/MGAL. The second analysis was done on the same basis as the first except that the selling price of phenol was calculated assuming the water wasn't being sold. The resulting price for phenol was \$0.64/lb. This value is far above the estimated selling price of \$0.29/lb.

Both of the above analyses were repeated except that the plant was not consider to be completely independent, allowing for negative Federal income taxes. These analyses gave a water selling price of \$17.58/MGAL and a phenol selling price of \$0.59/lb. This indicates a substantial advantage is obtained from the negative income taxes.

In the analyses performed to calculate the cost of treating the wastewater in \$/MGAL of treated water, it was assumed that the water was being sold and thus taxes were applied to the profits generated from the sales. This, however, is not really the case. For this reason, it was deemed necessary to incorporate the wastewater treatment plant into the analysis of a synthetic-gas producing plant to determine its effect on the SNG price. In order to do this, it was necessary to evaluate the price of SNG from a 20,000 ton/day coal gasification facility without water treatment.

A study was performed by C. F. Braun⁸ for the Tennessee Valley Authority to compare cost of three different gasification processes at a capacity of 20,000 ton/day of coal. This study provided capital investment and operating costs for the three processes. As a basis for

the current study, the cost data for a Babcock and Wilcox type gasification facility were used to evaluate the SNG price using the PRPIEA program. The cost data obtained from the Braun study was indexed to the 1st quarter of 1984 using the Chemical Engineering Plant index and are summarized in Table 15. The SNG price derived from this data was \$8.24/MMBTU. This analysis used the same basic parameters as was shown in Table 14 on page 84 except for those values shown in Table 15. Along with the SNG produced, a by-product sulfur was also produced and sold at a price of \$110/ton.¹³

The effect of adding the combined Phenosolvan--WAO waste treatment system on the SNG price as evaluated simply by adding in the operating and capital costs for the treatment system shown in Table 14 to the gasification profitability analysis program. The resulting SNG price was \$8.52/MMBTU.

These results indicate that the addition of the combined treatment process to the gasification plant caused a \$0.28/MMBTU increase in the price of the SNG. This value was converted to a \$/MGAL of water treated basis by multiplying it by the total amount of gas produced over the project life and dividing that result by the total amount of water treated over the project life. The final value was a cost of \$20.52/MGAL of treated water.

A separate set of analyses were performed assuming that only the Phenosolvan process was employed to clean up the wastewater to evaluate the viability of the Phenosolvan process. Using the capital investment and operating data calculated from a separate ASPEN simulation, the percent ROI and the effect of the process on the SNG price were evaluated

Table 15

Synthetic Natural Gas Capital Investment and Operating Costs

Raw Material and Product Summary			
Commodity	Annual Rate	Price ^a	Annual Am't \$MM
Coal	7,300,000 tons	\$ 45/ton	328.50
Synthetic Gas	126,003,840 MMBTUs	Calc'ed	Calc'ed
Sulfur	292,000 tons	\$ 110/ton	32.12
Annual Operating Costs			
Commodity		Annual Cost MM\$	
Utilities:			
Electricity		49.09	
Water (estimated)		11.00	
Catalyst and Chemicals		2.84	
Labor and Supervision		53.74	
Maintenance Labor and Materials		108.12	
Operating Supplies		1.42	
Administration and Overhead		27.77	

^aAll dollar values are shown in first quarter, 1984 dollars.

on the same basis as the previous analyses. The percent ROI for an independent treatment plant was determined to be 36% by the PRPIEA program. Combining the Phenosolvan process with the gasification plant provided a reduction in the SNG price of \$0.06/MMBTU. This indicated that the Phenosolvan process is a highly attractive method for water treatment.

All of the PRPIEA analysis programs and results are shown in Appendix C. A summary of the cost analyses produced from these program is shown in Table 16.

Table 16

PRPIEA Profitability Analyses Cost Summary

Process Simulation Description	PRPIEA Estimate ^a
Independent Phenosolvan Process: With Negative Taxes Without Negative Taxes	40.50% ROI 36.32% ROI
Combined Treatment Process: Phenol Price With Negative Taxes Without Negative Taxes Water Price With Negative Taxes Without Negative Taxes	 \$0.5860/lb \$0.6409/lb \$17.5823/MGAL \$20.7983/MGAL
Synthetic Gas Simulation: Indepedent Plant SNG-Phenosolvan-WAO SNG-Phenosolvan	 \$8.2362/MMBTU \$8.5192/MMBTU \$8.1769/MMBTU

^aAll dollar values are shown in first quarter, 1984 dollars.

V. RESULTS DISCUSSION

Given the results presented in Section 3, a determination must be made regarding the accuracy of the ASPEN simulations, the operating cost analysis, the profitability analysis and the viability of the combined Phenosolvan--Wet Air Oxidation process as a wastewater treatment alternative. Since this combined process has not been used before to treat coal-conversion plant wastewaters, a direct comparison can not be made with this study and literature results. However, limited data are available for both the Phenosolvan and WAO processes as individual treatment steps.

Mass and Energy Balance Simulation

Much of the accuracy in the simulation of the mass and energy balance for the combined treatment process depended upon the accuracy of the physical properties used in the simulation. The majority of the data used for this simulation came from the physical-property data base built into ASPEN. Extensive studies of the pure component data in the data base are currently being done by the ASPEN User's Group but have not yet been completed.³ Preliminary comparisons of the results from this study with the data bases from other physical-property studies have yielded good results.

The physical-property models used in ASPEN are in general the most recently developed and widely accepted models. The relative accuracy of these models have been reviewed extensively in the literature^{21, 28, 33, 39}. For this study, the results obtained were found to be relatively good

based on independent calculations which were made from actual physical property values.

The UNIQUAC binary interaction parameters derived using PATTERN also produced good results. Figure 6 and Table 10 on pages 49 and 50 show that the equilibrium data are accurately predicted throughout most of the equilibrium curve. The prediction, however, is not very good approaching the plate point but, since the extraction tower did not operate near the plate point, the deviation should not affect the results. The solvent recovery towers, on the other hand, operate over liquid-phase concentrations covering the entire phase curve. The deviation, however, did not appear to significantly affect the result produced from the tower simulations. This could not be verified, however, due to the lack of vapor-liquid equilibrium data for the Phenol--Water--NBA system.

Given that the physical properties simulated by ASPEN were reasonably accurate, the next major consideration was the accuracy of the unit-operation models. The unit-operation models in ASPEN use design equations which are widely used and are considered to be accurate for a preliminary design estimate.^{21,28,33,39} The results obtained from the major process equipment items such as the extraction tower and the solvent recovery towers were compared to simulations of the same equipment performed using a combination of other models available in ASPEN. The FLOWTRAN model simulation for the extraction tower was in complete agreement with a model comprised of a series of three-phase flash models combined countercurrently in series. The results for the recovery tower simulation produced using the RADFRC simulation model were compared to results produced using the DSTWU simulation model.

These values were also found to be in good agreement with each other. Unfortunately, actual operating data could not be found in the literature to verify the results obtained from the process equipment mentioned above.

The combined process provided a reduction in the phenol concentration from an inlet value of 7000 ppm to a discharge value of 0.25 ppm. This can be expressed in terms of Chemical Oxygen Demand, (COD), using COD ratios given by Verschueren for phenol and NBA.⁵² From this data, the COD concentration was estimated to be reduced from 16,300 ppm in the inlet stream down to 0.72 ppm in the discharge stream. The discharge value for phenol is very close to the BATEA standard of 130 ppb.⁴⁸ It is possible to increase the number of theoretical stages in the extraction tower and the primary-solvent-recovery tower to achieve the BATEA standard. However, it was difficult to justify increasing the phenol removal efficiency above 99.3% when the suggested removal efficiencies given in Section 2 on page 11 were around 99.5% for monohydric phenols and 60% for polyhydric phenols. The same reasoning can be invoked when the reaction-conversion extents for the WAO reactor are examined given a conversion of 99.5% for both the oxidation reactions of phenol and n-butyl acetate. It was surmised that 0.25 ppm was the lowest justifiable concentration for the combined process based on the suggested removal efficiencies.

Assuming this, some form of post treatment must be performed to reduce the phenol concentration to 130 ppb or below. One reasonable process to accomplish this would be a biological oxidation pond. This would offer the advantage of removing residual concentrations of phenol

and many other organics which would exist in the discharge water plus stabilization of minor fluctuations in the phenol concentration to the process.

Capital Investment Simulation

The capital investment simulation performed using ASPEN resulted in an overall capital investment estimate of \$125.7 MM, (first quarter, 1984). As was mentioned before in Section 2, this estimate has an accuracy of $\pm 30\%$. However, this estimate could only be obtained through some major revisions in the ASPEN code to correct or override errors in the cost estimation routines.

Initially, only the purchased equipment costs were generated using ASPEN to allow cross-checking of the results with other published cost data. This was done mainly using Guthrie¹⁹ as a source of accurate cost information. Given the equipment sizes calculated from the ASPEN costing routines, corresponding costs were determined from the Guthrie data for each item costed. The purchased equipment costs generated by ASPEN were found to be in very good agreement with the values estimated using Guthrie's data for all cases except the compressor cost. The ASPEN results were in error by an order of magnitude in the positive direction. Rather than using the ASPEN generated cost value for the compressor, the Guthrie estimated value was used as direct cost input to the compressor costing routine in ASPEN.

Once the purchased equipment costs were determined to be reasonable estimates based on the Guthrie cost data, an attempt was made to calculate the installed costs for the equipment and the overall capital

investment. This estimate initially could not be performed due to a serious error in the ASPEN subroutine CSECT which is responsible for calculating capital costs from the user supplied information. This error was later identified, corrected and submitted to the ASPEN User's Group for verification. This correction allowed the user to enter commodity material and labor-hour factors to correctly calculate the material and labor costs for installing the equipment. This led ultimately to determination of the overall capital cost of the combined process using ASPEN.

The overall capital investment calculated using ASPEN could not be directly compared to existing data for the combined process, however, data were available for both the Phenosolvan and WAO processes as separated treatment systems.⁴⁸ This led to a separate cost analysis for the Phenosolvan section of the process using ASPEN. The resulting capital-investment estimate was \$34.65 MM in first quarter, 1984 dollars. This value was compared to capital-investment data presented by Sing, et. al. for the Phenosolvan process.⁴⁸ Capital-investment data were developed by Sing for plant capacities ranging from 50 to 500 GPM in first quarter, 1978 dollars. These data were fitted to a first-order linear equation to yield

$$\text{Capital Investment [MM\$]} = 0.0051 * \text{GPM} + 1.111$$

with a correlation coefficient of 0.98.

Extrapolating the data to 3330 GPM using this equation and adjusting the value to first quarter, 1984 dollars gives a capital investment of \$27.5 MM. One must note that this value is a result of a very large extrapolation of the given data and therefore should not be considered as

very accurate. However, this does provide some credence to the \$34.65 MM estimate provided by ASPEN.

Limited data were also available on the capital costs for the Wet Air Oxidation process as an independent treatment process.⁴⁸ Sing has presented a capital cost estimation equation based on Chemical Oxygen Demand, (COD), removal per day. This equation is:

$$\text{Capital Cost} = \$7 \times 10^6 \times \frac{(\text{lb COD removed/day})}{21000}$$

This equation is based on a wastewater containing 10,000 ppm of COD at a flow rate of 100 GPM. The dollar values from the equation are given for the first quarter of 1978. He has also suggested an exponent of 0.4 as a scale up factor on plant capacity. In the simulation of the WAO process, approximately 2,010 lb/day of phenol were removed along with 890 lb/day of NBA. Assuming a ratio of 2.33 lb-COD/lb-phenol and 1.72 lb-COD/lb-NBA taken from data given by Verschueren⁵², a COD removal of 6,214 lb-COD/day is found. In addition to phenol and NBA, other organic compounds are expected to be in the wastewater contribution to the COD of the stream. Since the amounts of these materials are not known, it was not possible to calculate a COD removal for these compounds. However, to account for their presence a total COD removal of 10,000 lb/day was assumed. Given the above values and a plant capacity of 3330 GPM, a capital cost estimate of \$18.2 MM, (1978), was obtained. Indexing this value to the first quarter of 1984, a value of \$27.5 MM resulted. This value is clearly inconsistent with the ASPEN estimate of \$91 MM. This inconsistency is most likely due to the extent of extrapolation of the plant capacity. Since the actual process simulation contains six reactors, a more reasonable approach is to assume six plants at 555 GPM

each. Assuming this, a cost estimate of \$81 MM, (1984), is obtained from the same equation. The value is much more in line with the ASPEN estimate.

Due to the uncertainty in the extrapolation of the literature data, Zimpro Inc. was contacted to try and resolve the problem. In the resulting discussion with a Zimpro salesman concerning the process operating conditions, construction materials and contaminate removal values, the salesman indicated that the \$91 MM estimate was possibly a little high but could be considered as a reasonable estimate.⁵⁵ Based on a rough estimate, he suggested a value between \$80 MM and \$85 MM for the capital cost of the plant.

Profitability Analysis

The profitability analyses were performed using the PRPIEA program which has been found to be a reliable program based on prior usage at the the University of Tennessee. The analyses performed using the program provided some interesting results.

In assuming the wastewater treatment system was an independent process plant, a water credit of \$20.80/MGAL was required based upon a 15% ROI. Since data were not available concerning the expected cost per unit quantity of water for wastewaters of this type, it is not known whether this value represents a reasonable cost estimate. When the same plant is combined with the gasification facility, the analysis indicated that treatment of the water imparts a \$0.28/MMBTU cost increase to the SNG produced or in terms of amount of water treated, \$20.52/MGAL. This increase is only a 3.4% increase in the product price.

The small difference of \$0.28/MGAL between the independent water price and the combined SNG-treatment system can be attributed primarily to the relative differences in the amounts of working capital between the two analyses amounting to \$1.598 MM. This difference causes the total amount of interest to be greater for the independent cost analysis and therefore results in a higher selling price.

The major influence in the cost of the treated water is the intensive capital investment required for the WAO step of the combined process. The interest charges on the debt have a considerable impact on the cost to treat the water. This fact is born out by performing a cost analysis of the Phenosolvan process alone. This analysis indicated a surprising 36% ROI for an independent treatment plant. Combining this with the SNG process plant provided a \$0.06/MMBTU decrease on the price of the SNG. From this, one can see that the Phenosolvan process appears to be a highly desirable treatment process for phenolic wastewater. In considering the \$0.06/MMBTU decrease in the SNG price given by the Phenosolvan process, it can be concluded that addition of the WAO process to the treatment system results in a \$0.35/MMBTU increase in the SNG price.

From these results, it would appear that it might be advantageous to investigate another post-treatment process besides WAO. It should be noted, however, that one of the primary reasons for selecting the WAO process was its ability to oxidize a large variety of organic materials. It is quite possible that due to the variety of organic compounds which appear in the wastewater, many processing alternatives such as biological oxidation, carbon adsorption and ozonation might not be

adequate, particularly for compounds such as HCN and some of the phenolic compounds mentioned in Table 1 on page 4.

In this study, many assumptions were made which could have significant adverse impact on the results. One assumption which fits into this category involves the classification of the wastewater. In order to simplify the analysis, many materials were not considered or were lumped into a single compound. This is particularly true for the inorganic salts and oxides which exist in the wastewater stream. These materials could cause significant changes in the process process equipment and operating conditions due to their build up in recycle streams. The result would be a significant increase in the treatment costs. The same can also said for the assumption that phenol is representative of the other phenolic compounds in the wastewater stream with respect to their distribution coefficients. This is probably not the case for polyhydric phenols but, since distribution data are not available for many of the phenols listed in Table 1, it was not possible to evaluate the magnitude of the differences.

Overall the results found in this study appear to be reasonable and fairly consistent with literature data. The overall capital investment for the waste treatment system is less than 10% of the capital investment for the total gasification facility. This percentage is consistent with those of other chemical processes for water treatment.⁵⁶

VI. SUMMARY AND CONCLUSIONS

To summarize the results of this study, it is advantageous to relate the findings in terms of the objectives for the study.

The primary objective of the study was to evaluate the viability of a combined Phenosolvan--Wet Air Oxidation process in treating phenolic wastewater streams from a synthetic-fuel facility processing 20,000 ton/day of coal. In this respect, it was found that employing n-butyl acetate as a solvent in the combined process provided a reduction in the phenol concentration in the wastewater stream from 7000 ppm down to 0.25 ppm. The removal of phenol from the stream produced a COD reduction from 16,300 ppm down to 0.72 ppm. The phenol concentration, however, is still above the expected discharge standard of 0.13 ppm. By using a post-treatment process such as biological oxidation or carbon adsorption, the residual organic concentration can be reduced to acceptable levels.

In terms of economics of the combined process, the capital-investment cost was found to be \$126 MM. A profitability analysis of the process indicated that the cost to treat the wastewater was \$20.52/MGAL. In terms of the selling price of the synthetic natural gas from the conversion plant, this represents a \$0.28/MMBTU price increase.

The secondary objective of the study was to evaluate the performance of the ASPEN simulator in generating a mass and energy balance and a capital cost estimate for the combined process. In this aspect, ASPEN was found to be a very useful tool. It allowed manipulation of many different process parameters with rapid acquisition of results. The

values obtained using ASPEN compared well with the limited literature data available on the Phenosolvan and Wet Air Oxidation processes as separate process units.

The only major difficulties encountered using the program were some logic errors in the ASPEN code which caused program termination during the capital cost simulation. These problems were quickly resolved, however, and the capital cost estimates were calculated using ASPEN. One other problem was encountered when attempts to use the data regression system of ASPEN to estimate the UNIQUAC binary interaction parameters for the liquid-liquid equilibrium of Phenol--Water-N-Butyl Acetate failed. This problem was resolved through the use of the PATTERN search program to estimate the necessary parameters.

Given these results, the following conclusions were made:

1. Usage of the PATTERN search program to derive UNIQUAC binary interaction parameters in order to account for the non-ideal behavior of the chemicals in multi-component mixtures is a viable alternative when the derivatives of the minimizing equations can not be easily evaluated. The results obtained for the Phenol--Water--N-Butyl Acetate system compared very well to the actual equilibrium distribution curve taken from the literature.
2. N-butyl acetate, (NBA), is a very good solvent to use in the extraction of phenol in dilute concentrations from aqueous solutions. The ASPEN simulation indicated that when NBA is employed in a countercurrent extraction tower, a major fraction of the phenol in the aqueous stream is extracted with very low

solvent losses to the aqueous phase. In addition, NBA is also easily recovered from both the extract and raffinate phases allowing further reduction of solvent losses.

3. The combined Phenosolvan--WAO treatment system is very effective in the removal of phenol from coal-conversion wastewaters. The combined process reduced the phenol concentration from 7000 ppm to 0.25 ppm. However, to reduce the residual phenol concentration down to the expected BATEA limit suggested by the Environmental Protection Agency of 0.13 ppm, some form of post treatment will be necessary.
4. The overall economics of the combined process indicates that a water treatment cost of \$20.52/MGAL of treated water can be expected from this process. In terms of a cost to the SNG product from a coal-gasification facility, this value represents a \$0.28/MMBTU increase in the price of the SNG. Overall, this increase is only 3.4% of the total price of the product.
5. The ASPEN program is a valuable tool for the design and economic evaluation of chemical process alternatives. The program itself is a bit overwhelming to the general user due to its complexity and flexibility. However, after working with the language for a short time, the program becomes easy to use. The public version of ASPEN still contains some program logic errors which can cause erroneous results or even termination of program execution. These problems can generally be resolved but may require considerable effort by to user in some cases.

Once a simulation program has been developed, the ASPEN simulator has a significant advantage since the operating parameters of the process can be easily manipulated to evaluate their effect.

VII. RECOMMENDATIONS

Based on the results produced from this study, the following recommendation are made:

1. A more detailed breakdown of the gasification-wastewater analysis should be performed. It is obvious that for this study, the wastewater characterization was greatly simplified due to the lack of adequate information concerning the wastewater composition and its physical properties. This is especially true with respect to the various phenolic compounds which are in the wastewater. Many of these phenolic compounds could have distribution coefficients very different from that of phenol. This could significantly affect the overall removal efficiencies and the resulting equipment costs for the process.
2. Testing of additional solvents concerning their ability to recover phenol from aqueous streams should be performed. It is very possible that many other solvents could be very effective in the extraction of phenols. It is also possible that combinations of several different solvents could be very effective in treating coal-gasification wastewaters where many different phenolic compounds exist.
3. A study of the mechanisms and kinetics of the reactions which take place in the WAO process is recommended. Several studies have been made on the phenol WAO reaction kinetics, however, it was unclear in the literature as to what the reaction products were and their relative concentrations in the treated water. In addition, adequate information to properly design the reactor

was not available for the reactions of other organic materials in the wastewater stream.

4. Investigation of other less energy intensive methods of wastewater treatment as a substitute for the WAO in this treatment process may be advantageous. The WAO process is both highly energy and capital cost intensive. The energy requirements can be reduced depending on the amount of organic material in the stream. However, the capital cost can probably not be significantly reduced. It is possible that a combination of other treatment systems could be more economical in treating wastewaters of this nature.
5. Development of a binary physical-property data base for ASPEN would provide a substantial enhancement to its capabilities. One of the major drawbacks in this study was the lack of binary physical properties required to for the simulation of highly non-ideal processes such as extraction or azeotropic distillation. Inclusion of these data in ASPEN's data structure would greatly enhance its usability for these purposes.

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APPENDIXES

APPENDIX A

PATTERN REGRESSION PROGRAMS FOR THE
UNIQUEAC BINARY-INTERACTION PARAMETERS

```

C
C
C*****
C*
C*   This program is the MAIN program used to enter the necessary
C*   parameters required for the UNIQUAC equation contained in the
C*   subroutine PROC. The necessary parameters are defined below.
C*
C*   Program Name: MAIN
C*
C*   Program Title: MAIN PROGRAM FOR PATTERN-UNIQUAC DATA REGRESSION
C*
C*   Written By: Dale R. Heron           Date Written: 7/20/83
C*
C*   Variables Used:
C*
C*   Variable      Type      Dimension      Description
C*
C*   DISTE          R          3      Experimental Distribution
C*                               Coefficient for Each
C*                               Component.
C*   DISTC          R          3      UNIQUAC Estimated Distrib-
C*                               ution Coefficient for Each
C*                               Component.
C*   GAMMA          R        3,100    Vector of Acticity Coeff-
C*                               icients for Each Compound.
C*   IO             I          --     PATTERN Output Control Param.
C*   NAME           A        20,3     Chemical Component Name
C*   NC             I          --     Number of Components.
C*   NF             I          --     Input Data File Number.
C*   NP             I          --     Number of Search Parameters.
C*   NUMDAT         I          --     Number of Equilibrium Pairs
C*   P              R          6      UNIQUAC Binary Interaction
C*                               Parameters.
C*   Q              R          5      Pure Component van der
C*                               Surface Area Parameters.
C*   R              R          5      Pure Component van der
C*                               Waals Volume Parameters.
C*   RESULT         R          5      Summation of the Differ-
C*                               ences of the Activities
C*                               between Phases.
C*   STEP           R          6      Search Parameter Step Size.
C*   T              R          --     System Temperature
C*   X              R        3,100    Mole Fraction Vector for
C*                               each Component both Phases.
C*   Y              R        3,100    Activity of a given compon-
C*                               ent in the component mixture
C*
C*****
C
C
C

```

```

      IMPLICIT REAL (A-H,L,O-Z)

```

```

        DIMENSION NAME(20,3),P(6),STEP(6),DISTE(3),DISTC(3)
        COMMON X(3,100),GAMMA(3,100),R(5),Q(5),NC,T,NUMDAT,Y(3,100),
1  RESULT
        WRITE (5,1000)
1000  FORMAT(' PLEASE ENTER THE NUMBER OF COMPONENTS?', $)
        READ (5,*)NC
        DO 10 J=1,NC
            WRITE (5,1010) J
1010  FORMAT (' PLEASE ENTER THE NAME OF COMPONENT ',I2,'?', $)
            READ (5,1020) (NAME(I,J),I=1,20)
1020  FORMAT (20A1)
            WRITE (5,1030) (NAME(I,J),I=1,20)
1030  FORMAT(' PLEASE ENTER R AND Q FOR ',20A1,'?', $)
            READ (5,*) R(J),Q(J)
10  CONTINUE
        WRITE (5,1040)
1040  FORMAT(' PLEASE ENTER THE SYSTEM TEMPERATURE, [K]?', $)
        READ (5,*)T
        WRITE (5,1050)
1050  FORMAT (' ENTER THE DATA FILE NUMBER TO BE READ?', $)
        READ (5,*)NF
        WRITE (5,1070)
1070  FORMAT(' PLEASE ENTER THE NUMBER OF DATA GROUPS?', $)
        READ (5,*)NUMDAT
        DO 30 I=1,NUMDAT
            SUMX=0
            READ (NF,*) X(1,I),X(2,I)
C**
C** Calculate the value of the third mole fraction given the first two
C**
            SUMX=X(1,I)+X(2,I)
            X(3,I)=1-SUMX
30  CONTINUE
C**
C** Initial Guesses for the UNIQUAC binay interaction parameters.
C**
        P(1)= -213.5
        P(2)= 150.3
        P(3)= 71.0
        P(4)= -499.4
        P(5)= -22.4
        P(6)= -169.6
C**
C** Enter the search variable step sizes.
C**
        STEP(1)=1
        STEP(2)=1
        STEP(3)=1

```

```

        STEP(4)=1
        STEP(5)=1
        STEP(6)=1
        NP=6
        NPASS=2
        IO=1
C**
C** Create Output file called PHASE.FOR
C**
        OPEN(UNIT=30,FILE='PHASE.FOR')
C**
C** Call the Patern Program to Begin the Search
C**
        CALL PATERN(NP,P,STEP,NPASS,IO,COST)
C**
C** Print out the Results
C**
        WRITE(30,2000)
2000    FORMAT(40X,'UNQUAC PARAMETER DATA REGRESSION')
        WRITE(30,2010)
2010    FORMAT(/,40X,'PHASE EQUILIBRIUM DATA ESTIMATION')
        WRITE(30,2020)
2020    FORMAT(/,1X,'COMPONENT NOMENCLATURE')
        DO 40 I=1,NC
                WRITE(30,1080)I,(NAME(J,I),J=1,20)
1080        FORMAT(1X,'COMPONENT ',I2,' =',20A1)
        40    CONTINUE
C**
C** Print out the UNIQUAC parameters.
C**
        K=0
        WRITE(30,2030)
2030    FORMAT(/,1X,'REGRESSED UNIQUAC BINARY INTERACTION PARAMETERS')
        DO 60 J=1,NC
                DO 50 I=1,NC
                        IF(I.NE.J) GO TO 45
                        GOTO 50
                45        K=K+1
                        WRITE(30,1090)I,J,P(K)
1090    FORMAT(1X,'GMUQB(' ,I2,' ,',I2,' )=' ,F15.7)
                50        CONTINUE
                60        CONTINUE
        WRITE(30,4000)RESULT
4000    FORMAT(/,1X,'SUMMATION OF PHASE ACTIVITY DIFFERENCE=' ,F10.4)
C**
C** Print out a Summary Table
C**
        WRITE(30,2040)
2040    FORMAT(///,30X,'PHASE EQUILIBRIUM AND ACTIVITY COEFFICIENT
1 ESTIMATION SUMMARY')
        WRITE(30,1105)
        WRITE(30,1094)
1094    FORMAT(23X,'PHASE I',36X,'PHASE II')

```

```

        WRITE(30,1105)
        WRITE(30,1095)
1095    FORMAT(10X,'X(I,J)',7X,'GAMMA(I,J)',7X,'ACTIVITY',5X,'X(J,I)',
1      7X,'GAMMA(J,I)',7X,'ACTIVITY',7X,'KEXP',9X,'KEST')
        WRITE(30,1105)
C**
C** Divide the Number of Equilibrium Data Points by two to separate
C** the Extract and Raffinate Data.
C**
        DO 80 I=1,NUMDAT/2
            N=NUMDAT/2+I
            DO 70 J=1,NC
C**
C** Calculate the Expermental Distribution Coefficient from the
C** Given X-X Data pairs.
C**
                DISTE(J)=X(J,I)/X(J,N)
C**
C** Calculate the UNIQUAC Estimation of the Distribution Coefficient
C** Using the ratio of the Activity Coefficients.
C**
                DISTC(J)=GAMMA(J,N)/GAMMA(J,I)
                WRITE(30,1100)X(J,I),GAMMA(J,I),Y(J,I),X(J,N),GAMMA(J
1      ,N),Y(J,N),DISTE(J),DISTC(J)
1100    FORMAT(8F15.5)
        70    CONTINUE
        WRITE(30,1105)
1105    FORMAT(6X,'-----')
1      1-----')
        80    CONTINUE
        STOP
        END

```

```

C*****
C*
C*      This the the subroutine BOUNDS where the Boundries for the *
C*      search parameters are defined. In this case all of the *
C*      UNIQUAC binary interactions parameters were bounded between*
C*      -5000 and +5000. *
C*
C*****
C
C

```

```

      SUBROUTINE BOUNDS (P,IOUT)
      DIMENSION P(6)
      IOUT=0
      IF(ABS(P(1)).GT.5000) IOUT=1
      IF(ABS(P(2)).GT.5000) IOUT=1
      IF(ABS(P(3)).GT.5000) IOUT=1
      IF(ABS(P(4)).GT.5000) IOUT=1
      IF(ABS(P(5)).GT.5000) IOUT=1
      IF(ABS(P(6)).GT.5000) IOUT=1
      RETURN
      END

```

```

C
C*****
C*
C* This is the subroutine PROC which contains the UNIQUAC equation
C* to calculate the activity coefficients for all the components
C* in a given component mixture given the binary interaction
C* parameters.
C*
C* Subroutine Name: PROC
C*
C* Program Function: To Calculate Activity Coefficients and
C*                   Determine the Difference in the calculated
C*                   Activities for Phases in Equilibrium.
C*
C* Variables Used:
C*
C* Variable      Type      Dimension      Description
C*
C* ADIFF          R          --          Absolute value of the Differ-
C*                                     ence Between the Activity of
C*                                     Phases in Equilibrium.
C* COST           R          --          Summation of the Phase
C*                                     Activity Differences.
C* GAMMAC         R          5          Combination Contribution to
C*                                     the Activity Coefficient.
C* GAMMAR         R          5          Residual Contribution to the
C*                                     Activity Coefficient.
C* L              R          5          Script L parameter of the
C*                                     UNIQUAC equation.
C* LT             R          5          Log Term of the Residual
C*                                     Activity Coefficient.
C* NEGT           R          5          Negative Term of the Residual
C*                                     Activity Coefficient.
C* NC             I          --          Number of Components.
C* P              R          6          UNIQUAC Interaction Paramet-
C*                                     ers Calculated By PATTERN
C* PSI            R          5          Fractional Volume Contribution*
C*                                     of a Component to the mixture.*
C* Q              R          5          van der Waals Surface Param.
C* R              R          5          van der Waals Volume Parameter*
C* SUMLX          R          --          Summation of L(J) * X(J,N)
C*                                     for a given Comp. Mixture.
C* SUMQX          R          --          Summation of Q(J) * X(J,N)
C*                                     for a given Comp. Mixture.
C* SUMRX          R          --          Summation of R(J) * X(J,N)
C*                                     for a given Comp. Mixture.
C* T              R          --          System Temperature.
C* TAU            R          5,5        Dimensionless UNIQUAC Param.
C* THETA          R          5          Fractional Surface Area
C*                                     Contribution of a Component
C*                                     to the mixture.
C* Y              R          3,100      Activity of a given Component
C*                                     in the Component Mixture.

```

```

C*
C*****
C
      SUBROUTINE PROC (P,COST)
      IMPLICIT REAL(A-H,L,O-Z)
      REAL*8 NEGTL
      DIMENSION L(5),THETA(5),PSI(5),GAMMAC(5),GAMMAR(5),LT(5),
1  NEGTL(5),TAU(5,5),P(6)
      COMMON X(3,100),GAMMA(3,100),R(5),Q(5),NC,T,NUMDAT,Y(3,100),
1  RESULT
C**
C** Initialize Summers to Zero.
C**
      K=0
      RESULT=0
      DIFF=0
C**
C** Convert the UNIQUAC Parameters to a Dimensionless Value.
C**
      DO 20 J=1,NC
          DO 10 I=1,NC
              IF(I.NE.J) GO TO 5
              TAU(I,J)=1
              GO TO 10
          5
              K=K+1
              TAU(I,J)=EXP(P(K)/T)
          10
              CONTINUE
C**
C** Calculate the Script L Parameter.
      L(J)=5*(R(J)-Q(J))-(R(J)-1)
      20
      CONTINUE
      DO 120 N=1,NUMDAT
C**
C** Set the necessary Summers to Zero.
C**
          SUMRX=0
          SUMQX=0
          SUMLX=0
          DO 30 J=1,NC
              SUMRX= SUMRX + X(J,N)*R(J)
              SUMQX= SUMQX + X(J,N)*Q(J)
              SUMLX= SUMLX + X(J,N)*L(J)
          30
              CONTINUE
              DO 40 J=1,NC
C**
C** Calculate the Fractional Component Volume and Surface Area.
C**
                  THETA(J)=Q(J)*X(J,N)/SUMQX
                  PSI(J)=R(J)*X(J,N)/SUMRX
C**
C** Calculate the Combination Contribution to the Activity Coefficient.
C**

```

```

          GAMMAC(J)=ALOG(PHI(J)/X(J,N))+5*Q(J)*
1  ALOG(THETA(J)/PSI(J))+L(J)-PSI(J)/X(J,N)*SUXLX
40      CONTINUE
          DO 60 J=1,NC
              DO 50 I=1,NC
C**
C**      Calculate the Log Term of the Residual Contribution to the Activity
C**      Coefficient.
C**
          LT(J)=LT(J)+TAU(I,J)*THETA(I)
50      CONTINUE
60      CONTINUE
          DO 80 J=1,NC
              DO 70 I=1,NC
C**
C**      Calculate the Negative Term of the Residual Contribution to the
C**      Activity Coefficient.
C**
          NEGT(J)=NEGT(J)+THETA(I)*TAU(J,I)/LT(I)
70      CONTINUE
80      CONTINUE
          DO 90 J=1,NC
C**
C**      Calculate the Residual Contribution to the Activity Coefficient.
C**
          GAMMAR(J)=Q(J)*(1-ALOG(LT(J))-NEGT(J))
C**
C**      Calculate the Activity Coefficient by Combining the Combinational
C**      and Residual Contribution together.
C**
          GAMMA(J,N)=EXP(GAMMAC(J)+GAMMAR(J))
90      CONTINUE
          DO 110 J=1,NC
C**
C**      Reset Summers to Zero.
C**
          LT(J)=0.00
          NEGT(J)=0.00
110     CONTINUE
120     CONTINUE
          DO 140 I= 1,NUMDAT/2
              N=NUMDAT/2+I
              DO 130 J=1,NC
C**
C**      Calculate the Activity for Each Component in Each Phase in
C**      Equilibrium with Each Other.
C**
          Y(J,I)=X(J,I)*GAMMA(J,I)
          Y(J,N)=X(J,N)*GAMMA(J,N)
C**
C**      Calculate the Absolute value of the Difference between the
C**      Activity of Phases in Equilibrium with One another.
C**

```

```

                                ADIFF=ABS(Y(J,I)-Y(J,N))
C**
C** Sum up the Differences for Each Component over all the Equilibrium
C** Data Points.
C**
                                RESULT=RESULT+ADIFF
130                                CONTINUE
140                                CONTINUE
                                COST=RESULT
                                RETURN

                                END

```


0.48110	0.71520	0.34408	0.00900	39.40214	0.35462	53.45556	55.09270
0.29690	3.33332	0.99560	0.99070	1.00114	0.99182	0.29969	0.29855
0.22200	0.80331	0.17834	0.00030	869.52567	0.26086	740.00584	1082.42693
0.47010	0.82673	0.38864	0.01150	37.28961	0.42883	40.87826	45.10513
0.38500	2.80708	1.08073	0.98830	1.00173	0.99001	0.38956	0.35686
0.14490	0.76682	0.11111	0.00020	779.94900	0.15591	724.51469	1016.59852
0.46450	0.96043	0.44612	0.01360	35.65518	0.48491	34.15441	37.12406
0.44740	2.44033	1.09180	0.98630	1.00232	0.98859	0.45361	0.41073
0.08810	0.69142	0.06091	0.00010	712.62035	0.07126	880.98510	1030.66328
0.38220	1.32082	0.50482	0.01480	34.67703	0.51322	25.82432	26.29409
0.59220	1.85168	1.09657	0.98510	1.00273	0.98179	0.60116	0.54152
0.02560	0.84463	0.02162	0.00010	675.73687	0.06757	255.99560	800.03627
0.33740	1.61950	0.54642	0.01580	33.74210	0.53313	21.35443	20.83493
0.65500	1.63685	1.07214	0.98400	1.00316	0.98711	0.66565	0.61286
0.00760	1.07340	0.00816	0.00020	644.03399	0.12881	37.99933	599.99311

APPENDIX B

ASPEN MASS AND ENERGY BALANCE--CAPITAL COST

ESTIMATION REPORT SECTIONS

=====

R E P O R T

=====

DESCRIPTION

THIS IS A SIMULATION PROGRAM FOR A COMBINED PHENOSLOVAN WET-AIR
OXIDATION PROCESS USING PHENOL CONTAMINATED WASTE WATER FROM A
20,000 TON/DAY COAL SYN-FUELS PLANT.

RUN CONTROL INFORMATION

TYPE OF RUN: NEW

INPUT FILE NAME: JCLINPUT

INPUT PROBLEM DATA FILE NAME: PHENOL.D UPDATE NO. 0

MAIN CALLING PROGRAM NAME: PHENOL.D

SIMULATION REQUESTED FOR ENTIRE FLOWSHEET

FLowsheet SECTION.....	1
FLowsheet CONNECTIVITY BY STREAMS.....	1
FLowsheet CONNECTIVITY BY BLOCKS.....	1
DESIGN-SPEC: NBAIN.....	2
FORTRAN BLOCK: OXIN.....	2
CONVERGENCE BLOCK: CA.....	2
CONVERGENCE BLOCK: APR.....	3
CONVERGENCE BLOCK: AOX.....	3
CONVERGENCE BLOCK: ANB.....	3
CONVERGENCE BLOCK: BPR.....	4
COMPUTATIONAL SEQUENCE.....	4
OVERALL FLOWSHEET BALANCE.....	4
PHYSICAL PROPERTIES SECTION.....	5
COMPONENTS.....	5
OPTION SETS.....	6
U-O-S BLOCK SECTION.....	7
PUMP (PUMP) : P1.....	7
FLOWTRANBLCK (EXTRC) : EXTRACT.....	7
PUMP (PUMP) : P3.....	8
GENERAL-HEAT (HEATER) : PREHT2.....	9
FRACTIONATIO (RADFR) : SOLVR2.....	10
GENERAL-HEAT (HEATER) : COOL2.....	13
FLASH:3-OUTL (FLASH3) : SEP.....	13
MIXER (MIXER) : SMIX2.....	15
GENERAL-HEAT (HEATER) : HEATX2.....	15
PUMP (PUMP) : P5.....	16
PUMP (PUMP) : P2.....	17
MIXER (MIXER) : SMIX1.....	17
FLASH:2-OUTL (FLASH2) : HFLASH.....	18
PUMP (PUMP) : P6.....	19
GENERAL-HEAT (HEATER) : PREHT1.....	20
FRACTIONATIO (RADFR) : SOLVR.....	21
PUMP (PUMP) : P4.....	25
GENERAL-HEAT (HEATER) : HEATX.....	26
GENERAL-HEAT (HEATER) : COOL.....	27
COMPR-TURBIN (COMPR) : COMP.....	27
MIXER (MIXER) : OXMIX.....	28
HEAT-EXCHANG (HEATX) : OXPREHT.....	29
REACT-STOICH (RSTOIC) : OXREACT.....	30
FLASH:2-OUTL (FLASH2) : GASSEP.....	32
GENERAL-HEAT (HEATER) : COOL3.....	33
STREAM SECTION.....	34
DESCRIPTION OF STREAM CLASS HEAT.....	34
DESCRIPTION OF STREAM CLASS CONVEN.....	34

Q1 Q2.....	35
S101 S102 S203 S103 S104.....	36
S105 S106 S107 VAP S108.....	37
S211 S109 S110 S301 S201.....	38
S202 VAP2 S204 S205 S206.....	39
S207 PHEN S208 S209 S210.....	40
AQFEED SOLV S302 S303 S306.....	41
S304 S305 VENT S307 DISCHARG.....	42
OXSUP.....	43

=====

FLOWSHEET CONNECTIVITY BY STREAMS

FLOWSHEET SECTION SOLVAN

STREAM	SOURCE	DEST	STREAM	SOURCE	DEST
S101	P1	EXTRACT	S102	EXTRACT	P3
S203	EXTRACT	HFLASH	S103	P3	PREHT2
S104	PREHT2	SOLVR2	S105	SOLVR2	COOL2
S106	SOLVR2	SMIX2	S107	COOL2	SEP
VAP	SEP	----	S108	SEP	SMIX2
S211	SEP	SMIX1	S109	SMIX2	HEATX2
S110	HEATX2	P5	S301	P5	OXMIX
S201	P2	SMIX1	S202	SMIX1	EXTRACT
VAP2	HFLASH	----	S204	HFLASH	P6
S205	P6	PREHT1	S206	PREHT1	SOLVR
S207	SOLVR	P4	PHEN	SOLVR	----
S208	P4	HEATX	S209	HEATX	COOL
Q1	HEATX	PREHT1	S210	COOL	SMIX1
AQFEED	----	P1	Q2	----	PREHT2
SOLV	----	P2			

FLOWSHEET SECTION WETAIR

STREAM	SOURCE	DEST	STREAM	SOURCE	DEST
S302	COMP	OXMIX	S303	OXMIX	OXPREHT
S306	OXPREHT	GASSEP	S304	OXPREHT	OXREACT
S305	OXREACT	OXPREHT	VENT	GASSEP	----
S307	GASSEP	COOL3	DISCHARG	COOL3	----
OXSUP	----	COMP			

=====

FLOWSHEET CONNECTIVITY BY BLOCKS

FLOWSHEET SECTION SOLVAN

BLOCK	INLETS	OUTLETS
P1	AQFEED	S101
EXTRACT	S202 S101	S102 S203
P3	S102	S103
PREHT2	S103 Q2	S104
SOLVR2	S104	S105 S106
COOL2	S105	S107
SEP	S107	VAP S108 S211
SMIX2	S108 S106	S109
HEATX2	S109	S110
P5	S110	S301
P2	SOLV	S201
SMIX1	S201 S210 S211	S202
HFLASH	S203	VAP2 S204

FLOWSHEET CONNECTIVITY BY BLOCKS (CONTINUED)

FLOWSHEET SECTION WETAIR

BLOCK	INLETS	OUTLETS
P6	S204	S205
PREHT1	S205 Q1	S206
SOLVR	S206	S207 PHEN
P4	S207	S208
HEATX	S208	S209 Q1
COOL	S209	S210
COMP	OKSUP	S302
OXMIX	S301 S302	S303
OXPREHT	S305 S303	S306 S304
OXREACT	S304	S305
GASSEP	S306	VENT S307
COOL3	S307	DISCHARG

DESIGN-SPEC: NBAIN

SAMPLED VARIABLES:

A1IN	IS H2O MASSFLOW	IN STREAM AQFEED	SUBSTREAM MIXED
A2IN	IS H2O MASSFLOW	IN STREAM S202	SUBSTREAM MIXED
SOUT	IS NBA MASSFLOW	IN STREAM S202	SUBSTREAM MIXED

FORTTRAN STATEMENTS:

RATIO=(A1IN + A2IN)/SOUT)/10

SPECIFICATION:

MAKE RATIO APPROACH 1.00
 WITHIN 0.100000-04

MANIPULATED VARIABLES:

VARY THE NBA MOLEFLOW	IN STREAM SOLV	SUBSTREAM MIXED
BETWEEN 0.0	AND	10.00000 LBMOL/HR

FORTTRAN BLOCK: OXIN

SAMPLED VARIABLES:

PIN	IS C6H6O MOLEFLOW	IN STREAM S301	SUBSTREAM MIXED
FNBA	IS NBA MOLEFLOW	IN STREAM S301	SUBSTREAM MIXED
ANEED	IS O2 MOLEFLOW	IN STREAM OKSUP	SUBSTREAM MIXED

FORTTRAN STATEMENTS:

ANEED= 1.3*(PIN*5 +FNBA*4)

CONVERGENCE BLOCK: CA

TEAR STREAMS: S202

MAXIT= 30 WAIT 1 ITERATIONS BEFORE ACCELERATING

ACCELERATE EVERY 0 ITERS. QMAX= 0.0 QMIN= -5.000

CA (WEGSTN) ITER= 1 *** CONVERGED ***

		X	OLD X	ERROR
TOTAL MOLEFLOW	LBMOL/HR	0.1587160D+04	0.1587141D+04	0.1950000D-01
H2O MOLEFLOW	LBMOL/HR	0.1573488D+03	0.1573350D+03	0.1380000D-01
NBA MOLEFLOW	LBMOL/HR	0.1427220D+04	0.1427214D+04	0.5900000D-02
C6H6O MOLEFLOW	LBMOL/HR	0.2591000D+01	0.2591200D+01	-.2000000D-03
O2 MOLEFLOW	LBMOL/HR	0.0	0.0	0.0
C2H4O2 MOLEFLOW	LBMOL/HR	0.0	0.0	0.0
CO2 MOLEFLOW	LBMOL/HR	0.0	0.0	0.0
PRESSURE	PSI	0.3000000D+02	0.3000000D+02	0.2358753D-14
MASS ENTHALPY	BTU/HR	-.1613260D+08	-.1613252D+08	-.7918009D+02

CONVERGENCE BLOCK: APR

TEAR STREAMS: Q1

MAXIT= 30 WAIT 1 ITERATIONS BEFORE ACCELERATING

ACCELERATE EVERY 0 ITERS. QMAX= 0.0 QMIN= -5.000

APR (WEGSTN) ITER= 6 *** CONVERGED ***

		X	OLD X	ERROR
STRM-ATTR-VA		0.2295050D+07	0.2295059D+07	-.9900202D+01

CONVERGENCE BLOCK: AOX

TEAR STREAMS: S304

MAXIT= 30 WAIT 1 ITERATIONS BEFORE ACCELERATING

ACCELERATE EVERY 0 ITERS. QMAX= 0.0 QMIN= -5.000

AOX (WEGSTN) ITER= 2 *** CONVERGED ***

		X	OLD X	ERROR
TOTAL MOLEFLOW	LBMOL/HR	0.9187725D+05	0.9187725D+05	0.0
H2O MOLEFLOW	LBMOL/HR	0.9186855D+05	0.9186855D+05	0.0
NBA MOLEFLOW	LBMOL/HR	0.3210571D+00	0.3210571D+00	0.0
C6H6O MOLEFLOW	LBMOL/HR	0.8944153D+00	0.8944153D+00	0.0
O2 MOLEFLOW	LBMOL/HR	0.7483197D+01	0.7483197D+01	0.0
C2H4O2 MOLEFLOW	LBMOL/HR	0.0	0.0	0.0
CO2 MOLEFLOW	LBMOL/HR	0.0	0.0	0.0
PRESSURE	PSI	0.1595000D+04	0.1595000D+04	0.0
MASS ENTHALPY	BTU/HR	-.5003486D+08	-.5003486D+08	0.0

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=====
CONVERGENCE BLOCK: ANB
  SPECS: NBAIN
  MAXIT= 30 STEP-SIZE= .100D+71 MAX-STEP= .133D-03
  ANB (SECANT ) ITER= 1 *** CONVERGED ***
                                X          OLD X          ERROR
NBA MOLEFLOW    LBMOL/HR    0.1052900D+01 0.1052900D+01 -.3742170D-05
=====
```

```
CONVERGENCE BLOCK: BPR
  TEAR STREAMS: Q2
  MAXIT= 30 WAIT 1 ITERATIONS BEFORE ACCELERATING
  ACCELERATE EVERY 0 ITERS. QMAX= 0.0 QMIN= -5.000
  BPR (WEGSTN ) ITER= 1 *** CONVERGED ***
                                X          OLD X          ERROR
STRM-ATTR-VA                    0.1085301D+09 0.1085301D+09 0.0
=====
```

COMPUTATIONAL SEQUENCE

SEQUENCE USED WAS:
 P1 CA ANB P2 SMIX1 ANB<--- EXTRACT P3 BPR PREHT2 SOLVR2 COOL2 SEP
 SMIX2 HEATX2 BPR<--- HFLASH P6 APR PREHT1 SOLVR P4 HEATX APR<--- COOL
 CA <--- P5 OXIN COMP OXMIX AOX OXREACT OXPREHT AOX<--- GASSEP
 COOL3 TOWER1 TOWER2

OVERALL FLOWSHEET BALANCE

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*** MASS AND ENERGY BALANCE ***
```

	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	91868.9	91870.1	-.127873D-04
NBA	1.05290	.729957	.306717
C6H6O	123.965	123.076	.717201D-02
O2	7.48320	1.75567	.765385
C2H4O2	.0	1.52885	-1.00000
CO2	.0	4.19868	-1.00000
TOTAL BALANCE			
MOLE(LBMOL/HR)	92001.4	92001.4	.404926D-06
MASS(LB/HR)	.166705D+07	.166704D+07	.402178D-05
ENTHALPY(BTU/HR)	-.109351D+11	-.113966D+11	.405017D-01

COMPONENTS

ID	TYPE	FORMULA	NAME OR ALIAS
H2O	C	MISSING	WATER
NBA	C	MISSING	C6H12O2-1
C6H6O	C	MISSING	PHENOL
O2	C	O2	O2
C2H4O2	C	MISSING	C2H4O2-1
CO2	C	CO2	CO2

LISTID	SUPERCRITICAL COMPONENT LIST
SCCOMP	O2 CO2

=====

OPTION SETS

KEY TO OPTION SET TABLES:

OPTION SET ID
 MP KEYWORD MP ROUTE ID

SYSOP0		SYSOP11		SYSOP5	
PHILMX	PHILMX00	PHIVMX	PHIVMX01	PHIVMX	PHIVMX05
HVMX	HVMX00	PHILMX	PHILMX11	PHILMX	PHILMX05
HLMX	HLMX00	HVMX	HVMX01	HVMX	HVMX05
GVMX	GVMX00	HLMX	HLMX02	HLMX	HLMX05
GLMX	GLMX00	GVMX	GVMX01	GVMX	GVMX05
SVMX	SVMX00	GLMX	GLMX09	GLMX	GLMX05
SLMX	SLMX00	SVMX	SVMX01	SVMX	SVMX05
VVMX	VVMX00	SLMX	SLMX02	SLMX	SLMX05
VLMX	VLMX01	VVMX	VVMX01	VVMX	VVMX05
MUVMX	MUVMX01	VLMX	VLMX01	VLMX	VLMX05
MULMX	MULMX01	MUVMX	MUVMX02	MUVMX	MUVMX02
KVMX	KVMX01	MULMX	MULMX02	MULMX	MULMX02
KLMX	KLMX01	KVMX	KVMX02	KVMX	KVMX02
DVMX	DVMX01	KLMX	KLMX01	KLMX	KLMX01
DLMX	DLMX01	DVMX	DVMX02	DVMX	DVMX02
SIGLMX	SIGLMX01	DLMX	DLMX01	DLMX	DLMX01
PHIV	PHIV00	SIGLMX	SIGLMX01	SIGLMX	SIGLMX01
PHIL	PHIL00				
HV	HV00				
HL	HL00				
GV	GV00				
GL	GL00				
SV	SV00				
SL	SL00				
VV	VV00				
VL	VL01				
MUV	MUV01				
MUL	MUL01				
KV	KV01				
KL	KL01				
SIGL	SIGL01				
PHISMV	PHISMV02				
HSMX	HSMX02				
GSMX	GSMX02				
SSMX	SSMX01				
VSMX	VSMX02				
KSMX	KSMX01				
PHIS	PHIS02				
HS	HS02				
GS	GS02				
SS	SS02				
VS	VS01				
KS	KS01				

PUMP (PUMP) : P1
 INLET = AQFEED OUTLET = S101 &b& b
 PROPERTY OPTION SET SYSOP11

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	91868.9	91868.9	.0
NBA	.0	.0	.0
C6H6O	123.965	123.965	.0
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	91992.9	91992.9	.0
MASS(LB/HR)	.166669D+07	.166669D+07	.0
ENTHALPY(BTU/HR)	-.113051D+11	-.113050D+11	-.802753D-05

*** INPUT DATA ***

TYPE OF PUMP: 1=CENTRIFUGAL PUMP;	
2, SLURRY PUMP; 3, POSITIVE DISPLACEMENT PUMP	***
REQUIRED EXIT PRESSURE ,PSI	30.0000
PUMP EFFICIENCY ,	MISSING
DRIVER EFFICIENCY ,	1.00000
SOLID FLOW RATE ,LB/HR	0.0

*** RESULTS ***

TYPE OF PUMP, (CAL)	1
VOLUMETRIC FLOW RATE CUFT/HR	27,042.3
DELTA PRESSURE PSI	15.3000
FLUID POWER REQUIREMENTHP	30.0907
BRAKE POWER REQUIREMENTHP	35.6792
ELECTRICITY REQUIREMENT HP	35.6792
PUMP EFFICIENCY (CAL)	0.84337

FLOWTRANBLCK (EXTRC) : EXTRACT
 INPUT: S202 S101
 OUTPUT: S102 S203
 PROPERTY OPTION SET SYSOP11

FLOWTRANBLCK (EXTRC): EXTRACT (CONTINUED)

*** MASS AND ENERGY BALANCE ***			
	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	92026.2	92026.3	-.149957D-06
NBA	1427.21	1427.22	-.413391D-05
C6H6O	126.556	126.556	.158032D-05
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	93580.0	93580.0	-.208378D-06
MASS(LB/HR)	.183555D+07	.183555D+07	-.498558D-06
ENTHALPY(BTU/HR)	-.116483D+11	-.116484D+11	.804998D-05

*** PARAMETERS ***					
1	15.000	.0	.0	.0	50.000
6	.0	90.000	90.000	14.700	14.700
11	.10000D-02				

*** RETENTIONS ***		
1	91947.	91894.

=====

PUMP (PUMP): P3
 INLET = S102 OUTLET = S103 &b& b
 PROPERTY OPTION SET SYSOP11

*** MASS AND ENERGY BALANCE ***			
	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	91873.8	91873.8	.0
NBA	72.5861	72.5861	.0
C6H6O	.912193	.912193	.0
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	91947.3	91947.3	.0
MASS(LB/HR)	.166362D+07	.166362D+07	.0
ENTHALPY(BTU/HR)	-.113142D+11	-.113138D+11	-.342342D-04

*** INPUT DATA ***

TYPE OF PUMP: 1=CENTRIFUGAL PUMP;	
2, SLURRY PUMP; 3, POSITIVE DISPLACEMENT PUMP	***
REQUIRED EXIT PRESSURE ,PSI	80.0000
PUMP EFFICIENCY ,	MISSING
DRIVER EFFICIENCY ,	1.00000
SOLID FLOW RATE ,LB/HR	0.0

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 PHENOSOLVAN -- WET-AIR OXIDATION SIMULATION
 U-O-S BLOCK SECTION

PUMP (PUMP) : P3 (CONTINUED)

*** RESULTS ***

TYPE OF PUMP, (CAL)	1
VOLUMETRIC FLOW RATE CUFT/HR	27,042.6
DELTA PRESSURE PSI	65.3000
FLUID POWER REQUIREMENTHP	128.428
BRAKE POWER REQUIREMENTHP	152.279
ELECTRICITY REQUIREMENT HP	152.279
PUMP EFFICIENCY (CAL)	0.84337

=====

GENERAL-HEAT (HEATER): PREHT2
 INPUT STREAM: S103 Q2
 OUTPUT STREAM: S104
 PROPERTY OPTION SET SYSOP11

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	91873.8	91873.8	.0
NBA	72.5861	72.5861	.0
C6H6O	.912193	.912193	.0
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	91947.3	91947.3	.0
MASS(LB/HR)	.166362D+07	.166362D+07	.0
ENTHALPY(BTU/HR)	-.109435D+11	-.109435D+11	.606644D-08

*** INPUT DATA ***

TWO PHASE PQ FLASH	
PRESSURE DROP PSI	-10.000
SPECIFIED HEAT DUTY BTU/HR	.10000D+71
MAXIMUM ITERATION NO.	30
CONVERGENCE TOLERANCE	.10000D-03

*** RESULTS ***

OUTPUT TEMPERATURE F	283.78
OUTPUT PRESSURE PSI	70.000
HEAT DUTY BTU/HR	.0
VAPOR FRACTION	.0

V-L PHASE EQUILIBRIUM :

COMPONENT	F(I)	X(I)	Y(I)	K(I)
H2O	.99920	.99920	.88089	.88159
NBA	.78943D-03	.78943D-03	.11909	150.86
C6H6O	.99208D-05	.99208D-05	.23801D-04	2.3992

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FRACTIONATIO (RADFRC): SOLVR2
 INLETS - S104 STAGE 4
 OUTLETS - S105 STAGE 1
 S106 STAGE 10
 PROPERTY OPTION SET SYSOP11

	*** MASS AND ENERGY BALANCE ***		
	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	91873.8	91873.8	.111080D-09
NBA	72.5861	72.5861	-.140570D-06
C6H6O	.912193	.912193	-.214365D-08
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE (LBMOL/HR)	91947.3	91947.3	.0
MASS (LB/HR)	.166362D+07	.166362D+07	-.602036D-09
ENTHALPY (BTU/HR)	-.109435D+11	-.109262D+11	-.158073D-02

FRACTIONATIO (RADFRC): SOLVR2 (CONTINUED)

*** INPUT DATA ***

**** INPUT PARAMETERS ****

NUMBER OF THEORETICAL STAGES	10
MAXIMUM OUTSIDE LOOPS	25
MAXIMUM INSIDE LOOPS PER OUTSIDE LOOP	10
ALGORITHM OPTION	2
BOUNDED WEGSTEIN MODULUS FOR OUTSIDE LOOPS	1
MAXIMUM ITERATIONS FOR FEED FLASH	30
ENTHALPY BALANCE OPTION CODE	0
DESIGN SPECS METHOD OPTION CODE	1
KB UPDATING OPTION CODE	1
INSIDE LOOP ITERATION METHOD CODE	0
KB WEIGHTING OPTION CODE	2
MAX COMP BALANCES FOR NONIDEAL OPTION	10
OUTSIDE LOOP CONVERGENCE TOLERANCE	0.100000-03
MINIMUM INSIDE LOOP CONV TOLERANCE	0.300000-05
INITIAL INSIDE LOOP CONV TOLERANCE	0.0100000
INSIDE LOOP CONV TOL REDUCTION FACTOR	0.30000
INSIDE LOOP RMS ERR FOR JACOBIAN UPDATE	1.000000-06
COMPONENT MASS BALANCE CONV TOLERANCE	0.100000-06
COMP MASS BAL FACTOR FOR NONIDEAL CASE	0.0100000
COMP MASS BAL TOL FOR NONIDEAL CASE	0.100000-04
SIMPLE MODEL SLOPE PARAMETER UPDATE TOLERANCE	0.050000
QMIN FOR BOUNDED WEGSTEIN IN OUTSIDE LOOP	0.0
QMAX FOR BOUNDED WEGSTEIN IN OUTSIDE LOOP	0.50000
BOUNDED WEGSTEIN SLOPE TOLERANCE	0.050000
FACTOR 1 FOR PERFORMANCE SPEC CONV ALG	0.100000
FACTOR 2 FOR PERFORMANCE SPEC CONV ALG	0.20000
FEED FLASH CONVERGENCE TOLERANCE	1.000000-06
QMIN FOR BOUNDED WEGSTEIN IN INSIDE LOOP	0.0
QMAX FOR BOUNDED WEGSTEIN IN INSIDE LOOP	0.50000
LIMITING MF FOR NONIDEAL SIMPLE MODEL PARAMS	0.99000
MAX FOR NONIDEAL SIMPLE MODEL PARAMS AT X=0	1.50000
MAX FOR NONIDEAL SIMPLE MODEL PARAMS AT X=.5	1.30000
MIDDLE LOOP CONVERGENCE TOLERANCE	0.20000
MANIPULATED VARIABLE FACTOR	0.100000-04

**** COL-SPECS ****

VAPOR DISTILLATE / TOTAL DISTILLATE	0.0
REFLUX RATIO	1.00000
BOTTOMS RATE	LBMOL/HR 91,500.0

**** PROFILES ****

P-SPEC	STAGE 1 PRES, PSI	60.0000
	10	60.0000

FRACTIONATIO (RADFRC): SOLVR2 (CONTINUED)

*** RESULTS ***		
TOP TRAY TEMPERATURE	F	252.808
BOTTOM TRAY TEMPERATURE	F	292.986
TOP TRAY LIQUID FLOW	LBMOL/HR	447.269
BOTTOM TRAY LIQUID FLOW	LBMOL/HR	91,500.0
TOP TRAY VAPOR FLOW	LBMOL/HR	0.0
BOTTOM TRAY VAPOR FLOW	LBMOL/HR	1,985.17
CONDENSER DUTY	BTU/HR	-0.152426+08
REBOILER DUTY	BTU/HR	0.325412+08

**** PROFILES ****

STG	TEMP F	PRESS PSI	L-FLOW LBMOL/HR	V-FLOW LBMOL/HR
1	252.81	60.000	447.27	.0
2	281.72	60.000	457.57	894.54
3	287.54	60.000	460.92	904.84
4	287.58	60.000	92850.	908.19
5	290.74	60.000	93221.	1349.6
6	292.27	60.000	93402.	1720.5
7	292.78	60.000	93462.	1901.6
8	292.93	60.000	93480.	1962.2
9	292.97	60.000	93485.	1980.1
10	292.99	60.000	91500.	1985.2

****		X-PROFILE		****
STAGE	H2O	NBA	C6H6O	
1	.83774	.16222	.40859D-04	
2	.99886	.11194D-02	.18037D-04	
3	.99949	.49695D-03	.11383D-04	
4	.99950	.49473D-03	.10034D-04	
5	.99979	.20320D-03	.10127D-04	
6	.99992	.65268D-04	.10173D-04	
7	.99997	.18956D-04	.10189D-04	
8	.99998	.52987D-05	.10194D-04	
9	.99999	.14136D-05	.10172D-04	
10	.99999	.31434D-06	.97696D-05	

****		Y-PROFILE		****
STAGE	H2O	NBA	C6H6O	
1	.58683	.41317	.22831D-05	
2	.83774	.16222	.40859D-04	
3	.91922	.80754D-01	.29318D-04	
4	.91983	.80144D-01	.25899D-04	
5	.96596	.34015D-01	.27934D-04	
6	.98898	.10993D-01	.29153D-04	
7	.99678	.31908D-02	.29583D-04	
8	.99908	.88820D-03	.29760D-04	
9	.99973	.23563D-03	.29804D-04	
10	.99992	.52080D-04	.28715D-04	

=====

GENERAL-HEAT (HEATER): COOL2
 INPUT STREAM: S105
 OUTPUT STREAM: S107
 PROPERTY OPTION SET SYSOP11

*** MASS AND ENERGY BALANCE ***

CONVENTIONAL COMPONENTS	IN (LBMOL/HR)	OUT	RELATIVE DIFF.
H2O	374.693	374.693	.0
NBA	72.5573	72.5573	.0
C6H6O	.182748D-01	.182748D-01	.0
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	447.269	447.269	.0
MASS(LB/HR)	15180.1	15180.1	.0
ENTHALPY(BTU/HR)	-.605875D+08	-.625348D+08	.311400D-01

*** INPUT DATA ***

TWO PHASE TP FLASH
 SPECIFIED TEMPERATURE F 90.000
 PRESSURE DROP PSI -10.000
 MAXIMUM ITERATION NO. 30
 CONVERGENCE TOLERANCE .10000D-03
 TP FLASH, NO INITIAL GUESSES ARE REQUIRED.

*** RESULTS ***

OUTPUT TEMPERATURE F 90.000
 OUTPUT PRESSURE PSI 50.000
 HEAT DUTY BTU/HR -1.9473D+07
 VAPOR FRACTION .0

V-L PHASE EQUILIBRIUM :

COMPONENT	F(I)	X(I)	Y(I)	K(I)
H2O	.83774	.83774	.43929	.18664D-01
NBA	.16222	.16222	.56071	.12302
C6H6O	.40859D-04	.40859D-04	.45803D-06	.39900D-03

=====

FLASH:3-OUTL (FLASH3): SEP
 INPUT STREAM: S107
 OUTPUT STREAM: VAP S108 S211
 PROPERTY OPTION SET SYSOP11

FLASH:3-OUTL (FLASH3): SEP (CONTINUED)

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	374.693	374.693	.291549D-07
NBA	72.5573	72.5573	-.150522D-06
C6H6O	.182748D-01	.182748D-01	-.146278D-06
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE (LBMOL/HR)	447.269	447.269	.0
MASS (LB/HR)	15180.1	15180.1	-.706248D-07
ENTHALPY (BTU/HR)	-.625348D+08	-.625348D+08	.181353D-07

*** INPUT DATA ***

THREE PHASE PQ FLASH
 PRESSURE DROP PSI -10.000
 SPECIFIED HEAT DUTY BTU/HR .0
 MAXIMUM ITERATION NO. 30
 CONVERGENCE TOLERANCE .10000D-03
 KEY COMPONENT ID H2O
 LIQUID STREAM SPECIFICATION CODE 1
 SOLID SPLIT STREAM SPECIFICATION CODE 1
 LIQUID ENTRAINMENT OF 1ST LIQUID .0
 LIQUID ENTRAINMENT OF 2ND LIQUID .0
 SOLID SPLIT FRACTIONS:
 SUBSTREAM NO. = 1 MIXED SUBSTREAM, NO SOLID SPLITS.

*** RESULTS ***

OUTPUT TEMPERATURE F 90.000
 OUTPUT PRESSURE PSI 40.000
 HEAT DUTY BTU/HR .0
 VAPOR FRACTION .0
 1ST LIQUID/TOTAL LIQUID .82673

V-L1-L2 PHASE EQUILIBRIUM :

COMP	F(I)	X1(I)	X2(I)	Y(I)	K1(I)	K2(I)
H2O	.838	.999	.673D-01	.726	.728	.490D-01
NBA	.162	.790D-03	.932	.274	145.	.172D+06
C6H6O	.409D-04	.135D-05	.229D-03	.372D-05	1.76	299.

MIXER (MIXER): SMIX2
 INLET STREAM(S): S108 S106
 OUTLET STREAM: S109
 PROPERTY OPTION SET SYSOP11

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	91868.6	91868.6	-.422926D-13
NBA	.321057	.321057	-.477885D-10
C6H6O	.894415	.894415	-.522410D-12
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	91869.8	91869.8	-.426088D-13
MASS(LB/HR)	.165513D+07	.165513D+07	-.431862D-13
ENTHALPY(BTU/HR)	-.109111D+11	-.109111D+11	.897639D-13

*** INPUT DATA ***

PRESSURE DROP ,PSI	10.00000
TYPE OF FLASH - ONE PHASE - LIQUID	
MAXIMUM NUMBER OF ITERATIONS IN FLASH	30
CONVERGENCE TOLERANCE FOR FLASH	0.100000-03

GENERAL-HEAT (HEATER): HEATX2
 INPUT STREAM: S109
 OUTPUT STREAM: S110
 PROPERTY OPTION SET SYSOP11

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	91868.6	91868.6	.0
NBA	.321057	.321057	.0
C6H6O	.894415	.894415	.0
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	91869.8	91869.8	.0
MASS(LB/HR)	.165513D+07	.165513D+07	.0
ENTHALPY(BTU/HR)	-.109111D+11	-.112589D+11	.308934D-01

*** INPUT DATA ***

TWO PHASE TP FLASH	
SPECIFIED TEMPERATURE F	110.00
PRESSURE DROP PSI	-10.000
MAXIMUM ITERATION NO.	30
CONVERGENCE TOLERANCE	.10000D-03
TP FLASH, NO INITIAL GUESSES ARE REQUIRED.	

ASPEN VERSION ONE RELEASE 4 DATE: 11/08/84 PAGE 16
 PHENOSOLVAN -- WET-AIR OXIDATION SIMULATION
 U-O-S BLOCK SECTION

GENERAL-HEAT (HEATER): HEATX2 (CONTINUED)

*** RESULTS ***
 OUTPUT TEMPERATURE F 110.00
 OUTPUT PRESSURE PSI 20.000
 HEAT DUTY BTU/HR -.34783D+09
 VAPOR FRACTION .0

V-L PHASE EQUILIBRIUM :

COMPONENT	F(I)	X(I)	Y(I)	K(I)
H2O	.99999	.99999	.99819	.64250D-01
NBA	.34947D-05	.34947D-05	.17886D-02	32.942
C6H6O	.97357D-05	.97357D-05	.16717D-04	.11052

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PUMP (PUMP) : P5
 INLET = S110 OUTLET = S301 &b& b
 PROPERTY OPTION SET SYSOP11

*** MASS AND ENERGY BALANCE ***

CONVENTIONAL COMPONENTS	IN (LBMOL/HR)	OUT	RELATIVE DIFF.
H2O	91868.6	91868.6	.0
NBA	.321057	.321057	.864506D-16
C6H6O	.894415	.894415	.310321D-16
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	91869.8	91869.8	.0
MASS(LB/HR)	.165513D+07	.165513D+07	-.422015D-15
ENTHALPY(BTU/HR)	-.112589D+11	-.112494D+11	-.844851D-03

*** INPUT DATA ***

TYPE OF PUMP: 1=CENTRIFUGAL PUMP;
 2, SLURRY PUMP; 3, POSITIVE DISPLACEMENT PUMP
 REQUIRED EXIT PRESSURE ,PSI ***
 PUMP EFFICIENCY , 1,615.00
 DRIVER EFFICIENCY , MISSING
 SOLID FLOW RATE ,LB/HR 1.00000
 0.0

*** RESULTS ***

TYPE OF PUMP, (CAL)	1
VOLUMETRIC FLOW RATE CUFT/HR	27,187.5
DELTA PRESSURE PSI	1,595.00
FLUID POWER REQUIREMENTHP	3,153.74
BRAKE POWER REQUIREMENTHP	3,738.39
ELECTRICITY REQUIREMENT HP	3,738.39
PUMP EFFICIENCY (CAL)	0.84361

PUMP (PUMP): P2
 INLET = SOLV OUTLET = S201 &b& b
 PROPERTY OPTION SET SYSOP11

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	.0	.0	.0
NBA	1.05290	1.05290	.0
C6H6O	.0	.0	.0
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	1.05290	1.05290	.0
MASS(LB/HR)	122.305	122.305	.0
ENTHALPY(BTU/HR)	-239381.	-239346.	-.148580D-03

*** INPUT DATA ***

TYPE OF PUMP: 1=CENTRIFUGAL PUMP;	
2, SLURRY PUMP; 3, POSITIVE DISPLACEMENT PUMP	***
REQUIRED EXIT PRESSURE ,PSI	40.0000
PUMP EFFICIENCY ,	MISSING
DRIVER EFFICIENCY ,	1.00000
SOLID FLOW RATE ,LB/HR	0.0

*** RESULTS ***

TYPE OF PUMP, (CAL)	1
VOLUMETRIC FLOW RATE CUFT/HR	2.24610
DELTA PRESSURE PSI	25.3000
FLUID POWER REQUIREMENTHP	0.0041328
BRAKE POWER REQUIREMENTHP	0.013978
ELECTRICITY REQUIREMENT HP	0.013978
PUMP EFFICIENCY (CAL)	0.29566

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MIXER (MIXER): SMIX1
 INLET STREAM(S): S201 S210 S211
 OUTLET STREAM: S202
 PROPERTY OPTION SET SYSOP11

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	157.703	157.335	.233260D-02
NBA	1427.22	1427.21	.659831D-05
C6H6O	2.59013	2.59120	-.412591D-03
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	1587.52	1587.14	.236977D-03
MASS(LB/HR)	168871.	168863.	.451246D-04
ENTHALPY(BTU/HR)	-.343291D+09	-.343243D+09	-.139312D-03

*** INPUT DATA ***

PRESSURE DROP ,PSI 10.00000

TYPE OF FLASH - ONE PHASE - LIQUID

MAXIMUM NUMBER OF ITERATIONS IN FLASH 30

CONVERGENCE TOLERANCE FOR FLASH 0.100000-03

=====

FLASH:2-OUTL (FLASH2): HFLASH
 INPUT STREAM(S): S203
 OUTPUT STREAM(S): VAP2 S204
 PROPERTY OPTION SET SYSOP11

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	152.486	152.486	.139791D-15
NBA	1354.63	1354.63	.839244D-16
C6H6O	125.644	125.644	.113104D-15
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	1632.76	1632.76	.696284D-16
MASS(LB/HR)	171926.	171926.	.0
ENTHALPY(BTU/HR)	-.334180D+09	-.334180D+09	-.125209D-12

ASPEN VERSION ONE RELEASE 4 DATE: 11/08/84 PAGE 19
 PHENOSOLVAN -- WET-AIR OXIDATION SIMULATION
 U-O-S BLOCK SECTION

FLASH:2-OUTL (FLASH2): HFLASH (CONTINUED)

*** INPUT DATA ***

TWO PHASE PQ FLASH
 SPECIFIED PRESSURE PSI 14.700
 SPECIFIED HEAT DUTY BTU/HR .0
 MAXIMUM ITERATION NO. 30
 CONVERGENCE TOLERANCE .10000D-03
 LIQUID ENTRAINMENT .0
 SOLID SPLIT FRACTIONS:
 SUBSTREAM NO. = 1 MIXED SUBSTREAM, NO SOLID SPLITS.

*** RESULTS ***

OUTPUT TEMPERATURE F 90.000
 OUTPUT PRESSURE PSI 14.700
 HEAT DUTY BTU/HR .0
 VAPOR FRACTION .0

V-L PHASE EQUILIBRIUM :

COMPONENT	F(I)	X(I)	Y(I)	K(I)
H2O	.93391D-01	.93391D-01	.70054	7.5011
NBA	.82966	.82966	.29891	.36028
C6H6O	.76952D-01	.76952D-01	.54764D-03	.71166D-02

PUMP (PUMP): P6
 INLET = S204 OUTLET = S205 &b& b
 PROPERTY OPTION SET SYSOP11

*** MASS AND ENERGY BALANCE ***

CONVENTIONAL COMPONENTS	IN (LBMOL/HR)	OUT	RELATIVE DIFF.
H2O	152.486	152.486	.0
NBA	1354.63	1354.63	.0
C6H6O	125.644	125.644	.0
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	1632.76	1632.76	.0
MASS(LB/HR)	171926.	171926.	.0
ENTHALPY(BTU/HR)	-.334180D+09	-.334155D+09	-.758092D-04

*** INPUT DATA ***

TYPE OF PUMP: 1=CENTRIFUGAL PUMP;
 2, SLURRY PUMP; 3, POSITIVE DISPLACEMENT PUMP ***
 REQUIRED EXIT PRESSURE ,PSI 45.0000
 PUMP EFFICIENCY , MISSING
 DRIVER EFFICIENCY , 1.00000
 SOLID FLOW RATE ,LB/HR 0.0

PUMP (PUMP): P6 (CONTINUED)

*** RESULTS ***

TYPE OF PUMP, (CAL)	1
VOLUMETRIC FLOW RATE CUFT/HR	3,114.69
DELTA PRESSURE PSI	30.3000
FLUID POWER REQUIREMENTHP	6.86364
BRAKE POWER REQUIREMENTHP	9.95662
ELECTRICITY REQUIREMENT HP	9.95662
PUMP EFFICIENCY (CAL)	0.68935

=====

GENERAL-HEAT (HEATER): PREHT1
 INPUT STREAM: S205 Q1
 OUTPUT STREAM: S206
 PROPERTY OPTION SET SYSOP11

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	152.486	152.486	.0
NBA	1354.63	1354.63	.0
C6H6O	125.644	125.644	.0
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	1632.76	1632.76	.0
MASS(LB/HR)	171926.	171926.	.0
ENTHALPY(BTU/HR)	-.326324D+09	-.326324D+09	-.144707D-06

*** INPUT DATA ***

TWO PHASE PQ FLASH	
PRESSURE DROP PSI	-10.000
SPECIFIED HEAT DUTY BTU/HR	.10000D+71
MAXIMUM ITERATION NO.	30
CONVERGENCE TOLERANCE	.10000D-03

*** RESULTS ***

OUTPUT TEMPERATURE F	183.38
OUTPUT PRESSURE PSI	35.000
HEAT DUTY BTU/HR	.0
VAPOR FRACTION	.0

V-L PHASE EQUILIBRIUM :

COMPONENT	F(I)	X(I)	Y(I)	K(I)
H2O	.93391D-01	.93391D-01	.66927	7.1663
NBA	.82966	.82966	.32943	.39707
C6H6O	.76952D-01	.76952D-01	.12925D-02	.16796D-01

=====

FRACTIONATIO (RADFRC): SOLVR
 INLETS - S206 STAGE 10
 OUTLETS - S207 STAGE 1
 PHEN STAGE 30
 PROPERTY OPTION SET SYSOP11

	*** MASS AND ENERGY BALANCE ***		
	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	152.486	152.486	-.390692D-08
NBA	1354.63	1354.63	-.387921D-08
C6H6O	125.644	125.644	.465654D-07
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE (LBMOL/HR)	1632.76	1632.76	.696284D-16
MASS (LB/HR)	171926.	171926.	-.410174D-09
ENTHALPY (BTU/HR)	-.326324D+09	-.324070D+09	-.690721D-02

FRACTIONATIO (RADFRC): SOLVR (CONTINUED)

*** INPUT DATA ***

**** INPUT PARAMETERS ****

NUMBER OF THEORETICAL STAGES	30
MAXIMUM OUTSIDE LOOPS	25
MAXIMUM INSIDE LOOPS PER OUTSIDE LOOP	10
ALGORITHM OPTION	2
BOUNDED WEGSTEIN MODULUS FOR OUTSIDE LOOPS	1
MAXIMUM ITERATIONS FOR FEED FLASH	30
ENTHALPY BALANCE OPTION CODE	0
DESIGN SPECS METHOD OPTION CODE	1
KB UPDATING OPTION CODE	1
INSIDE LOOP ITERATION METHOD CODE	0
KB WEIGHTING OPTION CODE	2
MAX COMP BALANCES FOR NONIDEAL OPTION	10
OUTSIDE LOOP CONVERGENCE TOLERANCE	0.100000-03
MINIMUM INSIDE LOOP CONV TOLERANCE	0.300000-05
INITIAL INSIDE LOOP CONV TOLERANCE	0.0100000
INSIDE LOOP CONV TOL REDUCTION FACTOR	0.30000
INSIDE LOOP RMS ERR FOR JACOBIAN UPDATE	1.000000-06
COMPONENT MASS BALANCE CONV TOLERANCE	0.100000-06
COMP MASS BAL FACTOR FOR NONIDEAL CASE	0.0100000
COMP MASS BAL TOL FOR NONIDEAL CASE	0.100000-04
SIMPLE MODEL SLOPE PARAMETER UPDATE TOLERANCE	0.050000
QMIN FOR BOUNDED WEGSTEIN IN OUTSIDE LOOP	0.0
QMAX FOR BOUNDED WEGSTEIN IN OUTSIDE LOOP	0.50000
BOUNDED WEGSTEIN SLOPE TOLERANCE	0.050000
FACTOR 1 FOR PERFORMANCE SPEC CONV ALG	0.100000
FACTOR 2 FOR PERFORMANCE SPEC CONV ALG	0.20000
FEED FLASH CONVERGENCE TOLERANCE	1.000000-06
QMIN FOR BOUNDED WEGSTEIN IN INSIDE LOOP	0.0
QMAX FOR BOUNDED WEGSTEIN IN INSIDE LOOP	0.50000
LIMITING MF FOR NONIDEAL SIMPLE MODEL PARAMS	0.99000
MAX FOR NONIDEAL SIMPLE MODEL PARAMS AT X=0	1.50000
MAX FOR NONIDEAL SIMPLE MODEL PARAMS AT X=.5	1.30000
MIDDLE LOOP CONVERGENCE TOLERANCE	0.20000
MANIPULATED VARIABLE FACTOR	0.100000-04

**** COL-SPECS ****

VAPOR DISTILLATE / TOTAL DISTILLATE	0.0
REFLUX RATIO	1.00000
BOTTOMS RATE	LBMOL/HR 123.800

**** PROFILES ****

TEMP-EST	STAGE	1	TEMP, F	210.000
		30		360.000
P-SPEC	STAGE	1	PRES, PSI	14.7000
		30		14.7000

FRACTIONATIO (RADFRC): SOLVR (CONTINUED)

*** RESULTS ***		
TOP TRAY TEMPERATURE	F	197.818
BOTTOM TRAY TEMPERATURE	F	359.299
TOP TRAY LIQUID FLOW	LBMOL/HR	1,508.96
BOTTOM TRAY LIQUID FLOW	LBMOL/HR	123.800
TOP TRAY VAPOR FLOW	LBMOL/HR	0.0
BOTTOM TRAY VAPOR FLOW	LBMOL/HR	3,021.97
CONDENSER DUTY	BTU/HR	-0.566825+08
REBOILER DUTY	BTU/HR	0.589354+08

**** PROFILES ****

STG	TEMP F	PRESS PSI	L-FLOW LBMOL/HR	V-FLOW LBMOL/HR
1	197.82	14.700	1509.0	.0
2	254.04	14.700	1719.2	3017.9
3	264.77	14.700	1537.7	3228.2
4	286.31	14.700	1372.6	3046.7
5	308.07	14.700	1325.4	2881.5
6	318.17	14.700	1315.1	2834.4
7	321.60	14.700	1312.3	2824.1
8	322.67	14.700	1311.5	2821.3
9	323.00	14.700	1311.4	2820.5
10	323.09	14.700	3210.5	2820.3
11	348.10	14.700	3251.7	3086.7
12	354.96	14.700	3229.0	3127.9
13	357.22	14.700	3207.0	3105.2
14	358.13	14.700	3191.4	3083.2
15	358.57	14.700	3180.6	3067.6
16	358.81	14.700	3173.0	3056.8
17	358.96	14.700	3167.4	3049.2
18	359.05	14.700	3163.1	3043.6
19	359.11	14.700	3159.7	3039.3
20	359.16	14.700	3157.1	3035.9
21	359.19	14.700	3154.9	3033.3
22	359.22	14.700	3153.1	3031.1
23	359.24	14.700	3151.5	3029.3
24	359.25	14.700	3150.2	3027.7
25	359.27	14.700	3149.1	3026.4
26	359.28	14.700	3148.1	3025.3
27	359.28	14.700	3147.2	3024.3
28	359.29	14.700	3146.5	3023.4
29	359.29	14.700	3145.8	3022.7
30	359.30	14.700	123.80	3022.0

FRACTIONATIO (RADFRC): SOLVR (CONTINUED)

STAGE	H2O	NBA	X-PROFILE C6H6O
1	.10105	.89724	.17047D-02
2	.49184D-02	.97104	.24043D-01
3	.25505D-02	.85971	.13774
4	.26582D-02	.62275	.37459
5	.24582D-02	.45156	.54598
6	.23091D-02	.38238	.61531
7	.22532D-02	.35938	.63836
8	.22355D-02	.35219	.64557
9	.22302D-02	.35000	.64777
10	.22286D-02	.34935	.64842
11	.76656D-04	.19245	.80747
12	.25582D-05	.11932	.88068
13	.85538D-07	.81771D-01	.91823
14	.28786D-08	.60152D-01	.93985
15	.97459D-10	.46502D-01	.95350
16	.33160D-11	.37253D-01	.96275
17	.11327D-12	.30640D-01	.96936
18	.38812D-14	.25705D-01	.97429
19	.13333D-15	.21898D-01	.97810
20	.45900D-17	.18879D-01	.98112
21	.15829D-18	.16431D-01	.98357
22	.54667D-20	.14407D-01	.98559
23	.18904D-21	.12707D-01	.98729
24	.65441D-23	.11259D-01	.98874
25	.22675D-24	.10011D-01	.98999
26	.78639D-26	.89458D-02	.99105
27	.27293D-27	.80262D-02	.99197
28	.94789D-29	.72233D-02	.99278
29	.32939D-30	.65147D-02	.99349
30	.11438D-31	.58833D-02	.99412

STAGE	H2O	NBA	Y-PROFILE C6H6O
1	.69740	.30257	.23776D-04
2	.10105	.89724	.17047D-02
3	.49857D-01	.93654	.13601D-01
4	.51347D-01	.87830	.70350D-01
5	.54202D-01	.76654	.17926
6	.54967D-01	.68892	.25612
7	.55089D-01	.65758	.28733
8	.55117D-01	.64717	.29772
9	.55126D-01	.64392	.30095
10	.55131D-01	.64294	.30193
11	.23180D-02	.36313	.63455
12	.79691D-04	.19984	.80008
13	.26603D-05	.12384	.87615
14	.88973D-07	.84819D-01	.91518
15	.29948D-08	.62342D-01	.93766
16	.10141D-09	.48147D-01	.95185
17	.34506D-11	.38527D-01	.96147
18	.11788D-12	.31647D-01	.96835
19	.40393D-14	.26513D-01	.97349

FRACTIONATIO (RADFRC): SOLVR (CONTINUED)

STAGE	H2O	NBA	Y-PROFILE	C6H6O
20	.13877D-15	.22551D-01	.97745	
21	.47774D-17	.19410D-01	.98059	
22	.16475D-18	.16862D-01	.98314	
23	.56902D-20	.14755D-01	.98524	
24	.19677D-21	.12986D-01	.98701	
25	.68118D-23	.11479D-01	.98852	
26	.23603D-24	.10180D-01	.98982	
27	.81858D-26	.90712D-02	.99093	
28	.28410D-27	.81139D-02	.99189	
29	.98667D-29	.72782D-02	.99272	
30	.34242D-30	.65406D-02	.99346	

PUMP (PUMP): P4
 INLET = S207 OUTLET = S208 &b& b
 PROPERTY OPTION SET SYSOP11

*** MASS AND ENERGY BALANCE ***			
	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	152.486	152.486	.0
NBA	1353.91	1353.91	.0
C6H6O	2.57235	2.57235	.0
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	1508.96	1508.96	.0
MASS(LB/HR)	160259.	160259.	.0
ENTHALPY(BTU/HR)	-.317467D+09	-.317428D+09	-.120768D-03

*** INPUT DATA ***

TYPE OF PUMP: 1=CENTRIFUGAL PUMP;
 2, SLURRY PUMP; 3, POSITIVE DISPLACEMENT PUMP ***
 REQUIRED EXIT PRESSURE ,PSI 60.0000
 PUMP EFFICIENCY , MISSING
 DRIVER EFFICIENCY , 1.00000
 SOLID FLOW RATE ,LB/HR 0.0

*** RESULTS ***

TYPE OF PUMP, (CAL) 1
 VOLUMETRIC FLOW RATE CUFT/HR 3,159.15
 DELTA PRESSURE PSI 45.3000
 FLUID POWER REQUIREMENTHP 10.4080
 BRAKE POWER REQUIREMENTHP 15.0681
 ELECTRICITY REQUIREMENT HP 15.0681
 PUMP EFFICIENCY (CAL) 0.69073

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GENERAL-HEAT (HEATER): HEATX
 INPUT STREAM: S208
 OUTPUT STREAM: S209 Q1
 PROPERTY OPTION SET SYSOP11

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	152.486	152.486	.0
NBA	1353.91	1353.91	.0
C6H6O	2.57235	2.57235	.0
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	1508.96	1508.96	.0
MASS(LB/HR)	160259.	160259.	.0
ENTHALPY(BTU/HR)	-.317428D+09	-.317428D+09	-.187774D-15

*** INPUT DATA ***

TWO PHASE TP FLASH
 SPECIFIED TEMPERATURE F 100.00
 PRESSURE DROP PSI -10.000
 MAXIMUM ITERATION NO. 30
 CONVERGENCE TOLERANCE .10000D-03
 TP FLASH, NO INITIAL GUESSES ARE REQUIRED.

*** RESULTS ***

OUTPUT TEMPERATURE	F	100.00
OUTPUT PRESSURE	PSI	50.000
HEAT DUTY	BTU/HR	-.78310D+07
VAPOR FRACTION		.0

V-L PHASE EQUILIBRIUM :

COMPONENT	F(I)	X(I)	Y(I)	K(I)
H2O	.10105	.10105	.74609	.24059
NBA	.89724	.89724	.25390	.92220D-02
C6H6O	.17047D-02	.17047D-02	.89303D-05	.17072D-03

=====

GENERAL-HEAT (HEATER): COOL
 INPUT STREAM: S209
 OUTPUT STREAM: S210
 PROPERTY OPTION SET SYSOP11

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	152.486	152.486	.0
NBA	1353.91	1353.91	.0
C6H6O	2.57235	2.57235	.0
O2	.0	.0	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	1508.96	1508.96	.0
MASS(LB/HR)	160259.	160259.	.0
ENTHALPY(BTU/HR)	-.325259D+09	-.326017D+09	.232402D-02

*** INPUT DATA ***

TWO PHASE TP FLASH
 SPECIFIED TEMPERATURE F 90.000
 PRESSURE DROP PSI -10.000
 MAXIMUM ITERATION NO. 30
 CONVERGENCE TOLERANCE .10000D-03
 TP FLASH, NO INITIAL GUESSES ARE REQUIRED.

*** RESULTS ***

OUTPUT TEMPERATURE	F	90.000
OUTPUT PRESSURE	PSI	40.000
HEAT DUTY	BTU/HR	-.75767D+06
VAPOR FRACTION		.0

V-L PHASE EQUILIBRIUM :

COMPONENT	F(I)	X(I)	Y(I)	K(I)
H2O	.10105	.10105	.74848	.23056
NBA	.89724	.89724	.25151	.87257D-02
C6H6O	.17047D-02	.17047D-02	.79659D-05	.14546D-03

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COMPR-TURBIN (COMPR): COMP
 INPUT STREAMS: OXSUP
 OUTPUT STREAMS: S302
 PROPERTY OPTION SET SYSOP5

COMPR-TURBIN (COMPR): COMP (CONTINUED)

*** MASS AND ENERGY BALANCE ***			
	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	.0	.0	.0
NBA	.0	.0	.0
C6H6O	.0	.0	.0
O2	7.48320	7.48320	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE (LBMOL/HR)	7.48320	7.48320	.0
MASS (LB/HR)	239.455	239.455	.0
ENTHALPY (BTU/HR)	132.942	151919.	-.999125

*** INPUT DATA ***

TYPE : POLYTROPIC POSITIVE DISPLACEMENT COMPRESSOR	
OUTLET PRESSURE , PSI	1,635.00
POLYTROPIC EFFICIENCY	0.72000
MECHANICAL EFFICIENCY	1.00000
CLEARANCE FRACTION	1.00000

*** RESULTS ***

INDICATED HORSEPOWER REQUIREMENT, HP	59.6543
BRAKE HORSEPOWER REQUIREMENT, HP	59.6543
NET WORK, HP	59.6543
CALCULATED OUTLET TEMP, F	2,586.44
VOLUMETRIC EFFICIENCY	-9.24862
DISPLACEMENT , CUFT/HR	-318.553

MIXER (MIXER): OXMIX
 INLET STREAM(S): S301 S302
 OUTLET STREAM: S303
 PROPERTY OPTION SET SYSOP5

*** MASS AND ENERGY BALANCE ***

CONVENTIONAL COMPONENTS	IN (LBMOL/HR)	OUT	RELATIVE DIFF.
H2O	91868.6	91868.6	.0
NBA	.321057	.321057	.0
C6H6O	.894415	.894415	.0
O2	7.48320	7.48320	.0
C2H4O2	.0	.0	.0
CO2	.0	.0	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	91877.3	91877.3	.0
MASS(LB/HR)	.165537D+07	.165537D+07	.281303D-15
ENTHALPY(BTU/HR)	-.112493D+11	-.112493D+11	-.124317D-08

*** INPUT DATA ***

PRESSURE DROP ,PSI	10.00000
TYPE OF FLASH - TWO PHASE	
MAXIMUM NUMBER OF ITERATIONS IN FLASH	30
CONVERGENCE TOLERANCE FOR FLASH	0.100000-03

HEAT-EXCHANG (HEATX): OXPREHT
 INPUT STREAM S305 S303
 OUTPUT STREAM S306 S304
 PROPERTY OPTION SET SYSOP5

*** MASS AND ENERGY BALANCE ***

CONVENTIONAL COMPONENTS	IN (LBMOL/HR)	OUT	RELATIVE DIFF.
H2O	183739.	183739.	.0
NBA	.322662	.322662	.0
C6H6O	.898887	.898887	.0
O2	9.23887	9.23887	.0
C2H4O2	1.52885	1.52885	.0
CO2	4.19868	4.19868	.0
TOTAL BALANCE			
MOLE(LBMOL/HR)	183755.	183755.	.0
MASS(LB/HR)	.331075D+07	.331075D+07	.0
ENTHALPY(BTU/HR)	-.216131D+11	-.216131D+11	-.143984D-09

HEAT-EXCHANG (HEATX): OXPREHT (CONTINUED)

*** INPUT DATA ***

OPTION NUMBER		2
FLOW DIRECTION CODE		0
SPECIFIED HOT STREAM OUTLET TEMPERATURE		MISSING
SPECIFIED COLD STREAM OUTLET TEMPERATURE		550.000
SPECIFIED HEAT DUTY		MISSING
SPECIFIED TEMP APPROACH, HOT STREAM OUTLET		MISSING
SPECIFIED HEAT TRANSFER AREA		MISSING
SPECIFIED HOT STREAM PRESSURE DROP		-10.00000
SPECIFIED COLD STREAM PRESSURE DROP		-10.00000
SPECIFIED HEAT TRANSFER COEFFICIENTS:		
HOT STREAM PHASE	COLD STREAM PHASE	
LIQUID	LIQUID	149.694
BOILING LIQUID	LIQUID	149.694
VAPOR	LIQUID	149.694
LIQUID	BOILING LIQUID	149.694
BOILING LIQUID	BOILING LIQUID	149.694
VAPOR	BOILING LIQUID	149.694
LIQUID	VAPOR	149.694
BOILING LIQUID	VAPOR	149.694
VAPOR	VAPOR	149.694

*** RESULTS ***

HOT STREAM INLET TEMPERATURE	580.000
HOT STREAM OUTLET TEMPERATURE	202.002
COLD STREAM INLET TEMPERATURE	167.713
COLD STREAM OUTLET TEMPERATURE	550.000
EXCHANGER HEAT DUTY	0.813329+09
HEAT TRANSFER AREA	169,279.

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REACT-STOICH (RSTOIC): OXREACT
 INPUT STREAM: S304
 OUTPUT STREAM: S305
 PROPERTY OPTION SET SYSOP5

*** MASS AND ENERGY BALANCE ***

CONVENTIONAL COMPONENTS	IN (LBMOL/HR)	OUT	RELATIVE DIFF.
H2O	91868.6	91870.1	-.166414D-04
NBA	.321057	.160529D-02	.995000
C6H6O	.894415	.447208D-02	.995000
O2	7.48320	1.75567	.765385
C2H4O2	.0	1.52885	-1.00000
CO2	.0	4.19868	-1.00000
TOTAL BALANCE			
MOLE (LBMOL/HR)	91877.3	91877.6	-.347693D-05
MASS (LB/HR)	.165537D+07	.165537D+07	.923566D-09
ENTHALPY (BTU/HR)	-.104359D+11	-.103639D+11	-.690575D-02

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 PHENOSOLVAN -- WET-AIR OXIDATION SIMULATION
 U-O-S BLOCK SECTION

REACT-STOICH (RSTOIC): OXREACT (CONTINUED)

*** INPUT DATA ***

STOICHIOMETRIC MATRIX:

REACTION # 1:					
SUBSTREAM MIXED :					
H2O	C6H6O	O2	C2H4O2	CO2	
1.0000	-1.0000	-5.0000	1.0000	4.0000	
REACTION # 2:					
SUBSTREAM MIXED :					
H2O	NBA	O2	C2H4O2	CO2	
2.0000	-1.0000	-4.0000	2.0000	2.0000	

REACTION CONVERSION SPECS: NUMBER= 2

REACTION # 1:		
SUBSTREAM:MIXED	KEY COMP:C6H6O	CONV FRAC: .9950
REACTION # 2:		
SUBSTREAM:MIXED	KEY COMP:NBA	CONV FRAC: .9950

REACTION EXTENT SPECS: NUMBER= 0

TWO PHASE TP FLASH
 SPECIFIED TEMPERATURE F 580.00
 PRESSURE DROP PSI -15.000
 MAXIMUM ITERATION NO. 30
 CONVERGENCE TOLERANCE .10000D-03
 TP FLASH, NO INITIAL GUESSES ARE REQUIRED.

*** RESULTS ***

OUTPUT TEMPERATURE	F	580.00
OUTPUT PRESSURE	PSI	1580.0
HEAT DUTY	BTU/HR	.72063D+08
VAPOR FRACTION		.0

V-L PHASE EQUILIBRIUM :

COMPONENT	F(I)	X(I)	Y(I)	K(I)
H2O	.99992	.99992	.99154	.97134
NBA	.17472D-07	.17472D-07	.90217D-08	.50579
C6H6O	.48674D-07	.48674D-07	.24731D-08	.49770D-01
O2	.19109D-04	.19109D-04	.61108D-02	313.26
C2H4O2	.16640D-04	.16640D-04	.11908D-04	.70101
CO2	.45699D-04	.45699D-04	.23417D-02	50.194

FLASH:2-OUTL (FLASH2): GASSEP
 INPUT STREAM(S): S306
 OUTPUT STREAM(S): VENT S307
 PROPERTY OPTION SET SYSOP5

*** MASS AND ENERGY BALANCE ***

CONVENTIONAL COMPONENTS	IN (LBMOL/HR)	OUT	RELATIVE DIFF.
H2O	91870.1	91870.1	.0
NBA	.160529D-02	.160529D-02	.675395D-16
C6H6O	.447208D-02	.447208D-02	.0
O2	1.75567	1.75567	-.118758D-12
C2H4O2	1.52885	1.52885	.290473D-15
CO2	4.19868	4.19868	-.112432D-12
TOTAL BALANCE			
MOLE (LBMOL/HR)	91877.6	91877.6	.0
MASS (LB/HR)	.165537D+07	.165537D+07	-.140651D-15
ENTHALPY (BTU/HR)	-.111772D+11	-.111772D+11	.271721D-08

*** INPUT DATA ***

TWO PHASE PQ FLASH
 SPECIFIED PRESSURE PSI 25.000
 SPECIFIED HEAT DUTY BTU/HR .0
 MAXIMUM ITERATION NO. 30
 CONVERGENCE TOLERANCE .10000D-03
 LIQUID ENTRAINMENT .0
 SOLID SPLIT FRACTIONS:
 SUBSTREAM NO. = 1 MIXED SUBSTREAM, NO SOLID SPLITS.

*** RESULTS ***

OUTPUT TEMPERATURE F 204.33
 OUTPUT PRESSURE PSI 25.000
 HEAT DUTY BTU/HR .0
 VAPOR FRACTION .13152D-03

V-L PHASE EQUILIBRIUM :

COMPONENT	F(I)	X(I)	Y(I)	K(I)
H2O	.99992	.99998	.52612	.52613
NBA	.17472D-07	.17463D-07	.88122D-07	5.0464
C6H6O	.48674D-07	.48681D-07	.48826D-10	.10030D-02
O2	.19109D-04	.37961D-08	.14526	.38269D+08
C2H4O2	.16640D-04	.16641D-04	.95131D-05	.57167
CO2	.45699D-04	.24811D-05	.32860	.13245D+06

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GENERAL-HEAT (HEATER): COOL3
 INPUT STREAM: S307
 OUTPUT STREAM: DISCHARG
 PROPERTY OPTION SET SYSOP5

*** MASS AND ENERGY BALANCE ***

	IN	OUT	RELATIVE DIFF.
CONVENTIONAL COMPONENTS (LBMOL/HR)			
H2O	91863.7	91863.7	.0
NBA	.160422D-02	.160422D-02	.0
C6H6O	.447208D-02	.447208D-02	.0
O2	.348733D-03	.348733D-03	.0
C2H4O2	1.52873	1.52873	.0
CO2	.227928	.227928	.0
TOTAL BALANCE			
MOLE (LBMOL/HR)	91865.5	91865.5	.0
MASS (LB/HR)	.165503D+07	.165503D+07	.0
ENTHALPY (BTU/HR)	-.111759D+11	-.113887D+11	.186880D-01

*** INPUT DATA ***

TWO PHASE TP FLASH
 SPECIFIED TEMPERATURE F 100.00
 SPECIFIED PRESSURE PSI 14.700
 MAXIMUM ITERATION NO. 30
 CONVERGENCE TOLERANCE .10000D-03
 TP FLASH, NO INITIAL GUESSES ARE REQUIRED.

*** RESULTS ***

OUTPUT TEMPERATURE	F	100.00
OUTPUT PRESSURE	PSI	14.700
HEAT DUTY	BTU/HR	-.21283D+09
VAPOR FRACTION		.16014D-05

V-L PHASE EQUILIBRIUM :

COMPONENT	F(I)	X(I)	Y(I)	K(I)
H2O	.99998	.99998	.57021D-01	.57023D-01
NBA	.17463D-07	.17463D-07	.82202D-07	4.7074
C6H6O	.48681D-07	.48681D-07	.17048D-11	.35020D-04
O2	.37961D-08	.12316D-11	.23697D-02	.19241D+10
C2H4O2	.16641D-04	.16641D-04	.14292D-05	.85887D-01
CO2	.24811D-05	.97482D-06	.94061	.96491D+06

DESCRIPTION OF STREAM CLASS HEAT

STREAM CLASS : HEAT
STREAM ATTR : HEAT

DESCRIPTION OF STREAM CLASS CONVEN

STREAM CLASS : CONVEN
SUBSTREAMS : MIXED
SUBSTRM CLASS: MIXED

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Q1 Q2

STREAM ID	Q1	Q2
FROM :	HEATX	
TO :	PREHT1	PREHT2
CLASS:	HEAT	HEAT

STREAM ATTRIBUTES:

HEAT

Q	BTU/HR	.78310+07	.37032+09
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S101 S102 S203 S103 S104

STREAM ID	S101	S102	S203	S103	S104
FROM :	P1	EXTRACT	EXTRACT	P3	PREHT2
TO :	EXTRACT	P3	HFLASH	PREHT2	SOLVR2
CLASS:	CONVEN	CONVEN	CONVEN	CONVEN	CONVEN

SUBSTREAM: MIXED		STRUCTURE: CONVENTIONAL				
H2O	LBMOL/HR	.91869+05	.91874+05	152.4862	.91874+05	.91874+05
NBA	LBMOL/HR	0.0	72.5860	1354.6343	72.5860	72.5860
C6H6O	LBMOL/HR	123.9652	0.9121	125.6440	0.9121	0.9121
O2	LBMOL/HR	0.0	0.0	0.0	0.0	0.0
C2H4O2	LBMOL/HR	0.0	0.0	0.0	0.0	0.0
CO2	LBMOL/HR	0.0	0.0	0.0	0.0	0.0
TOTAL	LBMOL/HR	.91993+05	.91947+05	1632.7646	.91947+05	.91947+05
TEMP	F	90.0472	90.0000	90.0000	90.2019	283.7816
PRES	PSI	30.0000	14.7000	14.7000	30.0000	70.0000
ENTHALPY	BTU/LBMOL	-.12289+06	-.12305+06	-.20467+06	-.12305+06	-.11902+06
VFRAC		0.0	0.0	0.0	0.0	0.0
LFRAC		1.0000	1.0000	1.0000	1.0000	1.0000
ENTROPY	BTU/LBMOL-R	-46.4920	-.63241+61	MISSING	-.63241+61	-.63241+61
DENSITY	LBMOL/CUFT	3.4017	3.4000	MISSING	3.3997	3.0121
AVG MW		18.1175	18.0932	105.2975	18.0932	18.0932

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S105 S106 S107 VAP S108

STREAM ID		S105	S106	S107	VAP	S108
FROM :		SOLVR2	SOLVR2	COOL2	SEP	SEP
TO :		COOL2	SMIX2	SEP		SMIX2
CLASS:		CONVEN	CONVEN	CONVEN	CONVEN	CONVEN

SUBSTREAM: MIXED		STRUCTURE: CONVENTIONAL				
H2O	LBMOL/HR	374.6930	.91499+05	374.6930	0.0	369.4764
NBA	LBMOL/HR	72.5573	0.0287	72.5573	0.0	0.2922
C6H6O	LBMOL/HR	0.0182	0.8939	0.0182	0.0	.49754-03
O2	LBMOL/HR	0.0	0.0	0.0	0.0	0.0
C2H4O2	LBMOL/HR	0.0	0.0	0.0	0.0	0.0
CO2	LBMOL/HR	0.0	0.0	0.0	0.0	0.0
TOTAL	LBMOL/HR	447.2686	.91500+05	447.2686	0.0	369.7691
TEMP	F	252.8084	292.9864	90.0000	MISSING	90.0000
PRES	PSI	60.0000	60.0000	50.0000	MISSING	40.0000
ENTHALPY	BTU/LBMOL	-.13546+06	-.11875+06	-.13981+06	MISSING	-.12305+06
VFRAC		0.0	0.0	0.0	MISSING	0.0
LFRAC		1.0000	1.0000	1.0000	MISSING	1.0000
ENTROPY	BTU/LBMOL-R	-.12996+64	-.25182+58	-.12996+64	MISSING	-.63325+61
DENSITY	LBMOL/CUFT	1.5119	3.0083	1.7219	MISSING	3.4001
AVG MW		33.9395	18.0157	33.9395	MISSING	18.0926

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S211 S109 S110 S301 S201

STREAM ID	S211	S109	S110	S301	S201
FROM :	SEP	SMIX2	HEATX2	P5	P2
TO :	SMIX1	HEATX2	P5	OXMIX	SMIX1
CLASS:	CONVEN	CONVEN	CONVEN	CONVEN	CONVEN

SUBSTREAM: MIXED		STRUCTURE: CONVENTIONAL				
H2O	LBMOL/HR	5.2165	.91869+05	.91869+05	.91869+05	0.0
NBA	LBMOL/HR	72.2650	0.3210	0.3210	0.3210	1.0529
C6H6O	LBMOL/HR	0.0177	0.8944	0.8944	0.8944	0.0
O2	LBMOL/HR	0.0	0.0	0.0	0.0	0.0
C2H4O2	LBMOL/HR	0.0	0.0	0.0	0.0	0.0
CO2	LBMOL/HR	0.0	0.0	0.0	0.0	0.0
TOTAL	LBMOL/HR	77.4994	.91870+05	.91870+05	.91870+05	1.0529
TEMP	F	90.0000	292.1773	110.0000	114.9830	80.6402
PRES	PSI	40.0000	30.0000	20.0000	1615.0000	40.0000
ENTHALPY	BTU/LBMOL	-.21980+06-	.11877+06-	.12255+06-	.12245+06-	.22732+06
VFRAC		0.0	0.0	0.0	0.0	0.0
LFRAC		1.0000	1.0000	1.0000	1.0000	1.0000
ENTROPY	BTU/LBMOL-R	-.74699+64-	.27996+59-	.27996+59-	.27996+59-	.80109+64
DENSITY	LBMOL/CUFT	0.4973	3.0100	3.3791	3.3696	0.4685
AVG MW		109.5486	18.0160	18.0160	18.0160	116.1600

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S202 VAP2 S204 S205 S206

STREAM ID	S202	VAP2	S204	S205	S206
FROM :	SMIX1	HFLASH	HFLASH	P6	PREHT1
TO :	EXTRACT		P6	PREHT1	SOLVR
CLASS:	CONVEN	CONVEN	CONVEN	CONVEN	CONVEN

SUBSTREAM: MIXED		STRUCTURE: CONVENTIONAL				
H2O	LBMOL/HR	157.3350	0.0	152.4862	152.4862	152.4862
NBA	LBMOL/HR	1427.2145	0.0	1354.6343	1354.6343	1354.6343
C6H6O	LBMOL/HR	2.5912	0.0	125.6440	125.6440	125.6440
O2	LBMOL/HR	0.0	0.0	0.0	0.0	0.0
C2H4O2	LBMOL/HR	0.0	0.0	0.0	0.0	0.0
CO2	LBMOL/HR	0.0	0.0	0.0	0.0	0.0
TOTAL	LBMOL/HR	1587.1407	0.0	1632.7646	1632.7646	1632.7646
TEMP	F	90.0000	MISSING	89.9999	90.3144	183.3778
PRES	PSI	30.0000	14.7000	14.7000	45.0000	35.0000
ENTHALPY	BTU/LBMOL	-.21626+06	MISSING	-.20467+06	-.20466+06	-.19986+06
VFRAC		0.0	MISSING	0.0	0.0	0.0
LFRAC		1.0000	MISSING	1.0000	1.0000	1.0000
ENTROPY	BTU/LBMOL-R	-.72037+64	MISSING	-.66463+64	-.66463+64	-.66463+64
DENSITY	LBMOL/CUFT	0.5137	MISSING	0.5242	0.5241	0.4924
AVG MW		106.3947	MISSING	105.2975	105.2975	105.2975

S207 PHEN S208 S209 S210

STREAM ID	S207	PHEN	S208	S209	S210
FROM :	SOLVR	SOLVR	P4	HEATX	COOL
TO :	P4		HEATX	COOL	SMIX1
CLASS:	CONVEN	CONVEN	CONVEN	CONVEN	CONVEN

SUBSTREAM: MIXED		STRUCTURE: CONVENTIONAL				
H2O	LBMOL/HR	152.4862	.14160-29	152.4862	152.4862	152.4862
NBA	LBMOL/HR	1353.9060	0.7283	1353.9060	1353.9060	1353.9060
C6H6O	LBMOL/HR	2.5723	123.0716	2.5723	2.5723	2.5723
O2	LBMOL/HR	0.0	0.0	0.0	0.0	0.0
C2H4O2	LBMOL/HR	0.0	0.0	0.0	0.0	0.0
CO2	LBMOL/HR	0.0	0.0	0.0	0.0	0.0
TOTAL	LBMOL/HR	1508.9646	123.8000	1508.9646	1508.9646	1508.9646
TEMP	F	197.8183	359.2993	198.2786	100.000	90.0000
PRES	PSI	14.7000	14.7000	60.0000	50.0000	40.0000
ENTHALPY	BTU/LBMOL	-.21039+06-	.53339+05-	.21036+06-	.21555+06-	.21605+06
VFRAC		0.0	0.0	0.0	0.0	0.0
LFRAC		1.0000	1.0000	1.0000	1.0000	1.0000
ENTROPY	BTU/LBMOL-R	-.71877+64-	.47131+62-	.71877+64-	.71877+64-	.71877+64
DENSITY	LBMOL/CUFT	0.4776	0.6116	0.4774	0.5115	0.5148
AVG MW		106.2045	94.2427	106.2045	106.2045	106.2045

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AQFEED SOLV S302 S303 S306

STREAM ID	AQFEED	SOLV	S302	S303	S306
FROM :			COMP	OXMIX	OXPREHT
TO :	P1	P2	OXMIX	OXPREHT	GASSEP
CLASS:	CONVEN	CONVEN	CONVEN	CONVEN	CONVEN

SUBSTREAM: MIXED		STRUCTURE: CONVENTIONAL				
H2O	LBMOL/HR	.91869+05	0.0	0.0	.91869+05	.91870+05
NBA	LBMOL/HR	0.0	1.0529	0.0	0.3210	0.0016
C6H6O	LBMOL/HR	123.9652	0.0	0.0	0.8944	0.0044
O2	LBMOL/HR	0.0	0.0	7.4831	7.4831	1.7556
C2H4O2	LBMOL/HR	0.0	0.0	0.0	0.0	1.5288
CO2	LBMOL/HR	0.0	0.0	0.0	0.0	4.1986
TOTAL	LBMOL/HR	.91993+05	1.0529	7.4831	.91877+05	.91878+05
TEMP	F	90.0000	80.0000	2586.4429	167.7130	202.0019
PRES	PSI	14.7000	14.7000	1635.0000	1605.0000	1570.0000
ENTHALPY	BTU/LBMOL	-.12289+06	-.22735+06	.20301+05	-.12244+06	-.12165+06
VFRAC		0.0	0.0	1.0000	.81402-04	0.0
LFRAC		1.0000	1.0000	0.0	0.9999	1.0000
ENTROPY	BTU/LBMOL-R	-45.0802	-.80109+64	4.2312	-.27994+59	-.13997+57
DENSITY	LBMOL/CUFT	3.4018	0.4687	0.0489	3.2302	3.1674
AVG MW		18.1175	116.1600	31.9990	18.0172	18.0171

=====

S304 S305 VENT S307 DISCHARG

STREAM ID	S304	S305	VENT	S307	DISCHARG
FROM :	OXPREHT	OXREACT	GASSEP	GASSEP	COOL3
TO :	OXREACT	OXPREHT		COOL3	
CLASS:	CONVEN	CONVEN	CONVEN	CONVEN	CONVEN

SUBSTREAM: MIXED		STRUCTURE: CONVENTIONAL				
H2O	LBMOL/HR	.91869+05	.91870+05	6.3574	.91864+05	.91864+05
NBA	LBMOL/HR	0.3210	0.0016	.10648-05	0.0016	0.0016
C6H6O	LBMOL/HR	0.8944	0.0044	.59000-09	0.0044	0.0044
O2	LBMOL/HR	7.4831	1.7556	1.7553	.34873-03	.34873-03
C2H4O2	LBMOL/HR	0.0	1.5288	.11495-03	1.5287	1.5287
CO2	LBMOL/HR	0.0	4.1986	3.9707	0.2279	0.2279
TOTAL	LBMOL/HR	.91877+05	.91878+05	12.0836	.91865+05	.91865+05
TEMP	F	550.0000	580.0000	204.3269	204.3269	100.000
PRES	PSI	1595.0000	1580.0000	25.0000	25.0000	14.7000
ENTHALPY	BTU/LBMOL	-.11359+06	-.11280+06	-.10934+06	-.12165+06	-.12397+06
VFRAC		0.0	0.0	1.0000	0.0	.16014-05
LFRAC		1.0000	1.0000	0.0	1.0000	0.9999
ENTROPY	BTU/LBMOL-R	-.27994+59	-.13997+57	-.70594+57	-.13989+57	-.13989+57
DENSITY	LBMOL/CUFT	2.4153	2.3067	0.0035	3.1583	3.3564
AVG MW		18.0172	18.0171	28.5888	18.0157	18.0157

=====

OXSUP

STREAM ID	OXSUP
FROM :	
TO :	COMP
CLASS:	CONVEN

SUBSTREAM: MIXED		STRUCTURE: CONVENTIONAL
H2O	LBMOL/HR	0.0
NBA	LBMOL/HR	0.0
C6H6O	LBMOL/HR	0.0
O2	LBMOL/HR	7.4831
C2H4O2	LBMOL/HR	0.0
CO2	LBMOL/HR	0.0
TOTAL	LBMOL/HR	7.4831
TEMP	F	80.0000
PRES	PSI	14.7000
ENTHALPY	BTU/LBMOL	17.7653
VFRAC		1.0000
LFRAC		0.0
ENTROPY	BTU/LBMOL-R	0.0338
DENSITY	LBMOL/CUFT	0.0025
AVG MW		31.9990

ASPEN VERSION ONE

RELEASE 4

DATE: 11/08/84 PAGE I

R E P O R T

RUN CONTROL INFORMATION

TYPE OF RUN: NEW

INPUT FILE NAME: JCLINPUT

INPUT PROBLEM DATA FILE NAME: COST.DRH UPDATE NO. 0

MAIN CALLING PROGRAM NAME: COST.DRH

COST ONLY RUN

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COST BLOCK SECTION

COST-PUMP (CPC01): PUMP1

*** INPUT PARAMETERS ***

PUMP TYPE	ONE-STAGE, 1750 RPM, H.S.C.
NUMBER OF STANDBY PUMPS	1
MOTOR TYPE	TEFC
SEVERITY OF SERVICE	MODERATE SERVICE
MATERIAL OF CONSTRUCTION	CAST STEEL
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
TOTAL VOLUMETRIC FLOW, CUFT/HR	.27042+05
DENSITY OF FLUID, LB/CUFT	61.6300
PRESSURE RISE, PSI	15.3000

*** ASSOCIATED UTILITES ***

UTILITY IDENTIFICATION	ELECT
UTILITY TYPE	ELECTRICITY
POWER USAGE FOR UNIT, HP	40.0474

*** SIZING RESULTS ***

NUMBER OF PUMPS IN UNIT	1
REQUIRED PUMP HEAD, FT-LBF/LB	50.1828
VOLUMETRIC FLOW PER PUMP, CUFT/HR	.27042+05
PRESSURE RISE ACROSS PUMP, PSI	15.3000
MOTOR HORSEPOWER	40.0000
MATERIAL OF CONSTRUCTION FACTOR	1.2000
PUMP TYPE FACTOR	1.9644
PUMP EFFICIENCY	0.8435

*** COST RESULTS ***

BASE PUMP COST	\$ 2,646.25
BASE MOTOR COST	\$ 1,869.87
TOTAL BASE COST OF UNIT	\$ 9,032.24
ACTUAL COST OF PUMP	\$ 6,238.16
ACTUAL COST OF MOTOR	\$ 1,894.09
TOTAL COST OF UNIT	\$ 16,264.51

COST BLOCK SECTION

COST-PUMP (CPC01): PUMP2

*** INPUT PARAMETERS ***

PUMP TYPE	ONE-STAGE, 1750 RPM, H.S.C.
NUMBER OF STANDBY PUMPS	1
MOTOR TYPE	TEFC
SEVERITY OF SERVICE	MODERATE SERVICE
MATERIAL OF CONSTRUCTION	CAST STEEL
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
TOTAL VOLUMETRIC FLOW, CUFT/HR	2.2474
DENSITY OF FLUID, LB/CUFT	54.4400
PRESSURE RISE, PSI	25.3000

*** ASSOCIATED UTILITES ***

UTILITY IDENTIFICATION	ELECT
UTILITY TYPE	ELECTRICITY
POWER USAGE FOR UNIT, HP	6.7354

*** SIZING RESULTS ***

NUMBER OF PUMPS IN UNIT	1
REQUIRED PUMP HEAD, FT-LBF/LB	66.9213
VOLUMETRIC FLOW PER PUMP, CUFT/HR	2005.2021
PRESSURE RISE ACROSS PUMP, PSI	25.3000
MOTOR HORSEPOWER	7.5000
MATERIAL OF CONSTRUCTION FACTOR	1.2000
PUMP TYPE FACTOR	2.2573
PUMP EFFICIENCY	0.6443

*** COST RESULTS ***

BASE PUMP COST	\$ 1,214.35
BASE MOTOR COST	\$ 435.40
TOTAL BASE COST OF UNIT	\$ 3,299.51
ACTUAL COST OF PUMP	\$ 3,289.40
ACTUAL COST OF MOTOR	\$ 467.61
TOTAL COST OF UNIT	\$ 7,514.02

COST BLOCK SECTION

COST-PUMP (CPC01): PUMP3

*** INPUT PARAMETERS ***

PUMP TYPE	ONE-STAGE, 1750 RPM, H.S.C.
NUMBER OF STANDBY PUMPS	1
MOTOR TYPE	TEFC
SEVERITY OF SERVICE	MODERATE SERVICE
MATERIAL OF CONSTRUCTION	CAST STEEL
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
TOTAL VOLUMETRIC FLOW, CUFT/HR	.27043+05
DENSITY OF FLUID, LB/CUFT	61.5200
PRESSURE RISE, PSI	65.3000

*** ASSOCIATED UTILITES ***

UTILITY IDENTIFICATION	ELECT
UTILITY TYPE	ELECTRICITY
POWER USAGE FOR UNIT, HP	166.5151

*** SIZING RESULTS ***

NUMBER OF PUMPS IN UNIT	1
REQUIRED PUMP HEAD, FT-LBF/LB	152.8478
VOLUMETRIC FLOW PER PUMP, CUFT/HR	.27043+05
PRESSURE RISE ACROSS PUMP, PSI	65.3000
MOTOR HORSEPOWER	200.0000
MATERIAL OF CONSTRUCTION FACTOR	1.2000
PUMP TYPE FACTOR	1.9364
PUMP EFFICIENCY	0.8435

*** COST RESULTS ***

BASE PUMP COST	\$ 3,444.48
BASE MOTOR COST	\$ 10,112.04
TOTAL BASE COST OF UNIT	\$ 27,113.04
ACTUAL COST OF PUMP	\$ 8,004.10
ACTUAL COST OF MOTOR	\$ 9,243.17
TOTAL COST OF UNIT	\$ 34,494.55

COST BLOCK SECTION

COST-PUMP (CPC01): PUMP4

*** INPUT PARAMETERS ***

PUMP TYPE	ONE-STAGE, 1750 RPM, H.S.C.
NUMBER OF STANDBY PUMPS	1
MOTOR TYPE	TEFC
SEVERITY OF SERVICE	MODERATE SERVICE
MATERIAL OF CONSTRUCTION	CAST STEEL
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
TOTAL VOLUMETRIC FLOW, CUFT/HR	3159.4700
DENSITY OF FLUID, LB/CUFT	50.7200
PRESSURE RISE, PSI	45.3000

*** ASSOCIATED UTILITES ***

UTILITY IDENTIFICATION	ELECT
UTILITY TYPE	ELECTRICITY
POWER USAGE FOR UNIT, HP	17.2558

*** SIZING RESULTS ***

NUMBER OF PUMPS IN UNIT	1
REQUIRED PUMP HEAD, FT-LBF/LB	128.6119
VOLUMETRIC FLOW PER PUMP, CUFT/HR	3159.4700
PRESSURE RISE ACROSS PUMP, PSI	45.3000
MOTOR HORSEPOWER	20.0000
MATERIAL OF CONSTRUCTION FACTOR	1.2000
PUMP TYPE FACTOR	2.1311
PUMP EFFICIENCY	0.6908

*** COST RESULTS ***

BASE PUMP COST	\$ 1,453.32
BASE MOTOR COST	\$ 939.17
TOTAL BASE COST OF UNIT	\$ 4,784.98
ACTUAL COST OF PUMP	\$ 3,716.61
ACTUAL COST OF MOTOR	\$ 1,030.06
TOTAL COST OF UNIT	\$ 9,493.35

COST BLOCK SECTION

=====

COST-PUMP (CPC01): PUMP5A

*** INPUT PARAMETERS ***

PUMP TYPE	MULTI-STAGE, 3550 RPM, H.S.C
NUMBER OF STANDBY PUMPS	1
MOTOR TYPE	TEFC
SEVERITY OF SERVICE	MODERATE SERVICE
MATERIAL OF CONSTRUCTION	CAST STEEL
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
TOTAL VOLUMETRIC FLOW, CUFT/HR	.27188+05
DENSITY OF FLUID, LB/CUFT	60.8780
PRESSURE RISE, PSI	540.0000

*** ASSOCIATED UTILITES ***

UTILITY IDENTIFICATION	ELECT
UTILITY TYPE	ELECTRICITY
POWER USAGE FOR UNIT, HP	1477.2933

*** SIZING RESULTS ***

NUMBER OF PUMPS IN UNIT	3
REQUIRED PUMP HEAD, FT-LBF/LB	1277.3087
VOLUMETRIC FLOW PER PUMP, CUFT/HR	9062.5667
PRESSURE RISE ACROSS PUMP, PSI	540.0000
MOTOR HORSEPOWER	500.0000
MATERIAL OF CONSTRUCTION FACTOR	1.2000
PUMP TYPE FACTOR	8.3551
PUMP EFFICIENCY	0.7796

*** COST RESULTS ***

BASE PUMP COST	\$ 3,390.20
BASE MOTOR COST	\$ 25,316.91
TOTAL BASE COST OF UNIT	\$ 114,828.45
ACTUAL COST OF PUMP	\$ 33,990.72
ACTUAL COST OF MOTOR	\$ 25,316.91
TOTAL COST OF UNIT	\$ 237,230.52

COST BLOCK SECTION

COST-PUMP (CPC01): PUMP5B

*** INPUT PARAMETERS ***

PUMP TYPE	MULTI-STAGE, 3550 RPM, H.S.C
NUMBER OF STANDBY PUMPS	0
MOTOR TYPE	TEFC
SEVERITY OF SERVICE	MODERATE SERVICE
MATERIAL OF CONSTRUCTION	CAST STEEL
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
TOTAL VOLUMETRIC FLOW, CUFT/HR	.27188+05
DENSITY OF FLUID, LB/CUFT	60.8780
PRESSURE RISE, PSI	540.0000

*** ASSOCIATED UTILITIES ***

UTILITY IDENTIFICATION	ELECT
UTILITY TYPE	ELECTRICITY
POWER USAGE FOR UNIT, HP	1477.2933

*** SIZING RESULTS ***

NUMBER OF PUMPS IN UNIT	3
REQUIRED PUMP HEAD, FT-LBF/LB	1277.3087
VOLUMETRIC FLOW PER PUMP, CUFT/HR	9062.5667
PRESSURE RISE ACROSS PUMP, PSI	540.0000
MOTOR HORSEPOWER	500.0000
MATERIAL OF CONSTRUCTION FACTOR	1.2000
PUMP TYPE FACTOR	8.3551
PUMP EFFICIENCY	0.7796

*** COST RESULTS ***

BASE PUMP COST	\$ 3,390.20
BASE MOTOR COST	\$ 25,316.91
TOTAL BASE COST OF UNIT	\$ 86,121.34
ACTUAL COST OF PUMP	\$ 33,990.72
ACTUAL COST OF MOTOR	\$ 25,316.91
TOTAL COST OF UNIT	\$ 177,922.89

COST BLOCK SECTION

COST-PUMP (CPC01): PUMP5C

*** INPUT PARAMETERS ***

PUMP TYPE	MULTI-STAGE, 3550 RPM, H.S.C
NUMBER OF STANDBY PUMPS	0
MOTOR TYPE	TEFC
SEVERITY OF SERVICE	MODERATE SERVICE
MATERIAL OF CONSTRUCTION	CAST STEEL
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
TOTAL VOLUMETRIC FLOW, CUFT/HR	.27188+05
DENSITY OF FLUID, LB/CUFT	60.8780
PRESSURE RISE, PSI	535.0000

*** ASSOCIATED UTILITES ***

UTILITY IDENTIFICATION	ELECT
UTILITY TYPE	ELECTRICITY
POWER USAGE FOR UNIT, HP	1463.7561

*** SIZING RESULTS ***

NUMBER OF PUMPS IN UNIT	3
REQUIRED PUMP HEAD, FT-LBF/LB	1265.4818
VOLUMETRIC FLOW PER PUMP, CUFT/HR	9062.5667
PRESSURE RISE ACROSS PUMP, PSI	535.0000
MOTOR HORSEPOWER	500.0000
MATERIAL OF CONSTRUCTION FACTOR	1.2000
PUMP TYPE FACTOR	8.3492
PUMP EFFICIENCY	0.7796

*** COST RESULTS ***

BASE PUMP COST	\$ 3,382.35
BASE MOTOR COST	\$ 25,316.91
TOTAL BASE COST OF UNIT	\$ 86,097.77
ACTUAL COST OF PUMP	\$ 33,887.94
ACTUAL COST OF MOTOR	\$ 25,316.91
TOTAL COST OF UNIT	\$ 177,614.56

COST BLOCK SECTION

=====

COST-PUMP (CPC01): PUMP6

*** INPUT PARAMETERS ***

PUMP TYPE	ONE-STAGE, 1750 RPM, H.S.C.
NUMBER OF STANDBY PUMPS	1
MOTOR TYPE	TEFC
SEVERITY OF SERVICE	MODERATE SERVICE
MATERIAL OF CONSTRUCTION	CAST STEEL
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
TOTAL VOLUMETRIC FLOW, CUFT/HR	3114.7700
DENSITY OF FLUID, LB/CUFT	55.2000
PRESSURE RISE, PSI	30.3000

*** ASSOCIATED UTILITIES ***

UTILITY IDENTIFICATION	ELECT
UTILITY TYPE	ELECTRICITY
POWER USAGE FOR UNIT, HP	11.5261

*** SIZING RESULTS ***

NUMBER OF PUMPS IN UNIT	1
REQUIRED PUMP HEAD, FT-LBF/LB	79.0434
VOLUMETRIC FLOW PER PUMP, CUFT/HR	3114.7700
PRESSURE RISE ACROSS PUMP, PSI	30.3000
MOTOR HORSEPOWER	10.0000
MATERIAL OF CONSTRUCTION FACTOR	1.2000
PUMP TYPE FACTOR	2.1689
PUMP EFFICIENCY	0.6894

*** COST RESULTS ***

BASE PUMP COST	\$ 1,360.17
BASE MOTOR COST	\$ 482.71
TOTAL BASE COST OF UNIT	\$ 3,685.76
ACTUAL COST OF PUMP	\$ 3,540.14
ACTUAL COST OF MOTOR	\$ 585.55
TOTAL COST OF UNIT	\$ 8,251.38

COST BLOCK SECTION

=====

COST-COMPRES (CPC02): COMP

REPORT FOR COST BLOCK COMP

*** INPUT PARAMETERS ***

NUMBER OF STAND-BY COMPRESSORS	0
TYPE OF COMPRESSOR	MOTOR DRIVEN
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
TOTAL VOLUME FLOW, CUFT/HR	2993.2400
GAS HORSEPOWER, HP	59.6543
MECHANICAL EFFICIENCY	0.7500

*** USER SPECIFIED COST ***

USER SPECIFIED BASE COST	\$ 42,507.00
USER SPECIFIED ACTUAL COST	\$ 54,834.00
EQUIPMENT COST INDEX VALUE	340.4000
COST INDEX SOURCE	E.N.R. CONSTRUCTION INDEX

*** COST RESULTS ***

BASE COST OF COMPRESSOR	\$ 42,507.00
BASE COST OF UNIT	\$ 42,507.00
ACTUAL COST OF COMPRESSOR	\$ 54,834.00
ACTUAL COST OF UNIT	\$ 54,834.00

COST BLOCK SECTION

COST-HEATEX (CHE01): HEATX1
REPORT FOR COST BLOCK HEATX1

*** INPUT PARAMETERS ***

NUMBER OF STAND-BY HEAT EXCHANGERS	0
TYPE OF HEAT EXCHANGER	FLOATING HEAD
SHELL MATERIAL OF CONSTRUCTION	CARBON STEEL
TUBE MATERIAL OF CONSTRUCTION	STAINLESS STEEL 316
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
SHELL PRESSURE RATING, PSI	35.0000
TUBE PRESSURE RATING, PSI	50.0000
INLET TEMPERATURE OF FIRST FLUID, F	90.3140
OUTLET TEMPERATURE OF FIRST FLUID, F	183.3778
INLET TEMPERATURE OF SECOND FLUID, F	198.2786
OUTLET TEMPERATURE OF SECOND FLUID, F	100.000
HEAT TRANSFER COEFFICIENT, BTU/HR-SQFT-R	100.000
NUMBER OF SHELL PASSES	1
NUMBER OF TUBE PASSES	1

*** ASSOCIATED UTILITES ***

HEAT TRANSFER DUTY, BTU/HR	.78310+07
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*** SIZING RESULTS ***

NUMBER OF HEAT EXCHANGERS	1
HEAT TRANSFER AREA PER UNIT, SQFT	4312.1793
MATERIAL OF CONSTRUCTION FACTOR	2.3963
DESIGN TYPE FACTOR	1.0000
PRESSURE RATING FACTOR	1.0000
NUMBER OF PRESSURE RANGES	4

*** COST RESULTS ***

BASE COST OF HEAT EXCHANGER UNIT	\$ 60,960.82
ACTUAL COST OF HEAT EXCHANGER UNIT	\$ 146,084.61

COST BLOCK SECTION

COST-HEATEX (CHE01): HEATX2

REPORT FOR COST BLOCK HEATX2

*** INPUT PARAMETERS ***

NUMBER OF STAND-BY HEAT EXCHANGERS	0
TYPE OF HEAT EXCHANGER	FLOATING HEAD
SHELL MATERIAL OF CONSTRUCTION	CARBON STEEL
TUBE MATERIAL OF CONSTRUCTION	STAINLESS STEEL 316
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
SHELL PRESSURE RATING,PSI	1600.0000
TUBE PRESSURE RATING,PSI	1625.0000
INLET TEMPERATURE OF FIRST FLUID,F	580.0000
OUTLET TEMPERATURE OF FIRST FLUID,F	201.9900
INLET TEMPERATURE OF SECOND FLUID,F	167.7380
OUTLET TEMPERATURE OF SECOND FLUID,F	550.0000
HEAT TRANSFER COEFFICIENT, BTU/HR-SQFT-R	550.0000
NUMBER OF SHELL PASSES	1
NUMBER OF TUBE PASSES	1

*** ASSOCIATED UTILITIES ***

HEAT TRANSFER DUTY,BTU/HR	.81316+09
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*** SIZING RESULTS ***

NUMBER OF HEAT EXCHANGERS	11
HEAT TRANSFER AREA PER UNIT,SQFT	.11522+05
MATERIAL OF CONSTRUCTION FACTOR	2.6253
DESIGN TYPE FACTOR	1.0000
PRESSURE RATING FACTOR	2.2704
NUMBER OF PRESSURE RANGES	4

*** COST RESULTS ***

BASE COST OF HEAT EXCHANGER UNIT	\$ 1,621,426.59
ACTUAL COST OF HEAT EXCHANGER UNIT	\$ 9,664,887.31

COST BLOCK SECTION

COST-HEATEX (CHE01): HEATX3
REPORT FOR COST BLOCK HEATX3

*** INPUT PARAMETERS ***

NUMBER OF STAND-BY HEAT EXCHANGERS	0
TYPE OF HEAT EXCHANGER	FLOATING HEAD
SHELL MATERIAL OF CONSTRUCTION	CARBON STEEL
TUBE MATERIAL OF CONSTRUCTION	CARBON STEEL
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
SHELL PRESSURE RATING,PSI	30.0000
TUBE PRESSURE RATING,PSI	50.0000
INLET TEMPERATURE OF FIRST FLUID,F	80.0000
OUTLET TEMPERATURE OF FIRST FLUID,F	90.0000
INLET TEMPERATURE OF SECOND FLUID,F	100.000
OUTLET TEMPERATURE OF SECOND FLUID,F	90.0000
HEAT TRANSFER COEFFICIENT, BTU/HR-SQFT-R	90.0000
NUMBER OF SHELL PASSES	1
NUMBER OF TUBE PASSES	1
WATER RATE OF CONSUMPTION, LB/HR	.75767+06

*** ASSOCIATED UTILITES ***

HEAT TRANSFER DUTY,BTU/HR	.75767+06
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*** SIZING RESULTS ***

NUMBER OF HEAT EXCHANGERS	1
HEAT TRANSFER AREA PER UNIT,SQFT	757.6700
MATERIAL OF CONSTRUCTION FACTOR	1.0000
DESIGN TYPE FACTOR	1.0000
PRESSURE RATING FACTOR	1.0000
NUMBER OF PRESSURE RANGES	4

*** COST RESULTS ***

BASE COST OF HEAT EXCHANGER UNIT	\$ 17,643.77
ACTUAL COST OF HEAT EXCHANGER UNIT	\$ 17,643.77

COST BLOCK SECTION

COST-HEATEX (CHE01): HEATX4
REPORT FOR COST BLOCK HEATX4

*** INPUT PARAMETERS ***

NUMBER OF STAND-BY HEAT EXCHANGERS	0
TYPE OF HEAT EXCHANGER	FLOATING HEAD
SHELL MATERIAL OF CONSTRUCTION	CARBON STEEL
TUBE MATERIAL OF CONSTRUCTION	CARBON STEEL
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
SHELL PRESSURE RATING,PSI	30.0000
TUBE PRESSURE RATING,PSI	50.0000
INLET TEMPERATURE OF FIRST FLUID,F	80.0000
OUTLET TEMPERATURE OF FIRST FLUID,F	100.000
INLET TEMPERATURE OF SECOND FLUID,F	252.8084
OUTLET TEMPERATURE OF SECOND FLUID,F	90.0000
HEAT TRANSFER COEFFICIENT, BTU/HR-SQFT-R	90.0000
NUMBER OF SHELL PASSES	1
NUMBER OF TUBE PASSES	1
WATER RATE OF CONSUMPTION,LB/HR	.19473+07

*** ASSOCIATED UTILITES ***

HEAT TRANSFER DUTY,BTU/HR	.19473+07
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*** SIZING RESULTS ***

NUMBER OF HEAT EXCHANGERS	1
HEAT TRANSFER AREA PER UNIT,SQFT	371.7923
MATERIAL OF CONSTRUCTION FACTOR	1.0000
DESIGN TYPE FACTOR	1.0000
PRESSURE RATING FACTOR	1.0000
NUMBER OF PRESSURE RANGES	4

*** COST RESULTS ***

BASE COST OF HEAT EXCHANGER UNIT	\$ 11,960.83
ACTUAL COST OF HEAT EXCHANGER UNIT	\$ 11,960.83

COST BLOCK SECTION

=====

COST-HEATEX (CHE01): HEATX5
REPORT FOR COST BLOCK HEATX5

*** INPUT PARAMETERS ***

NUMBER OF STAND-BY HEAT EXCHANGERS	0
TYPE OF HEAT EXCHANGER	FLOATING HEAD
SHELL MATERIAL OF CONSTRUCTION	CARBON STEEL
TUBE MATERIAL OF CONSTRUCTION	CARBON STEEL
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
SHELL PRESSURE RATING,PSI	30.0000
TUBE PRESSURE RATING,PSI	25.0000
INLET TEMPERATURE OF FIRST FLUID,F	80.0000
OUTLET TEMPERATURE OF FIRST FLUID,F	100.000
INLET TEMPERATURE OF SECOND FLUID,F	204.3494
OUTLET TEMPERATURE OF SECOND FLUID,F	100.000
HEAT TRANSFER COEFFICIENT, BTU/HR-SQFT-R	100.000
NUMBER OF SHELL PASSES	1
NUMBER OF TUBE PASSES	1
WATER RATE OF CONSUMPTION,LB/HR	.21288+09

*** ASSOCIATED UTILITES ***

HEAT TRANSFER DUTY,BTU/HR	.21288+09
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*** SIZING RESULTS ***

NUMBER OF HEAT EXCHANGERS	3
HEAT TRANSFER AREA PER UNIT,SQFT	9265.1767
MATERIAL OF CONSTRUCTION FACTOR	1.0000
DESIGN TYPE FACTOR	1.0000
PRESSURE RATING FACTOR	1.0000
NUMBER OF PRESSURE RANGES	4

*** COST RESULTS ***

BASE COST OF HEAT EXCHANGER UNIT	\$ 359,450.20
ACTUAL COST OF HEAT EXCHANGER UNIT	\$ 359,450.20

COST BLOCK SECTION

=====

COST-HEATEX (CHE01): HEATX6
REPORT FOR COST BLOCK HEATX6

*** INPUT PARAMETERS ***

NUMBER OF STAND-BY HEAT EXCHANGERS	0
TYPE OF HEAT EXCHANGER	FLOATING HEAD
SHELL MATERIAL OF CONSTRUCTION	CARBON STEEL
TUBE MATERIAL OF CONSTRUCTION	STAINLESS STEEL 316
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
SHELL PRESSURE RATING, PSI	30.0000
TUBE PRESSURE RATING, PSI	80.0000
INLET TEMPERATURE OF FIRST FLUID, F	90.2019
OUTLET TEMPERATURE OF FIRST FLUID, F	283.7816
INLET TEMPERATURE OF SECOND FLUID, F	292.1773
OUTLET TEMPERATURE OF SECOND FLUID, F	110.0000
HEAT TRANSFER COEFFICIENT, BTU/HR-SQFT-R	110.0000
NUMBER OF SHELL PASSES	1
NUMBER OF TUBE PASSES	1

*** ASSOCIATED UTILITIES ***

HEAT TRANSFER DUTY, BTU/HR	.37032+09
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*** SIZING RESULTS ***

NUMBER OF HEAT EXCHANGERS	16
HEAT TRANSFER AREA PER UNIT, SQFT	.11609+05
MATERIAL OF CONSTRUCTION FACTOR	2.6271
DESIGN TYPE FACTOR	1.0000
PRESSURE RATING FACTOR	1.0000
NUMBER OF PRESSURE RANGES	4

*** COST RESULTS ***

BASE COST OF HEAT EXCHANGER UNIT	\$ 2,375,588.22
ACTUAL COST OF HEAT EXCHANGER UNIT	\$ 6,240,942.37

COST BLOCK SECTION

=====

COST-FURNACE (CHE03): FURNACE

*** INPUT PARAMETERS ***

NUMBER OF STAND-BY FURNACES	0
FURNACE TYPE	PROCESS HEATER
MATERIAL OF CONSTRUCTION	STAINLESS STEEL
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
TOTAL HEAT TRANSFER DUTY, BTU/HR	.71997+08
PRESSURE RATING, PSI	1610.0000
FURNACE THERMAL EFFICIENCY	0.7500

*** SIZING RESULTS ***

MATERIAL OF CONSTRUCTION FACTOR	1.3500
TYPE AND PRESSURE FACTOR	1.2600
HEAT DUTY PER FURNACE, BTU/HR	.71997+08
SCALED HEAT TRANSFER DUTY, BTU/HR	.71997+08
TOTAL ENERGY CONSUMPTION, BTU/HR	.95996+08

*** COST RESULTS ***

TOTAL BASE COST OF UNIT	\$ 727,982.29
TOTAL ACTUAL COST OF FURNACE	\$ 1,238,297.87

COST BLOCK SECTION

=====

COST-VERVSSL (CVS11): REACT

REPORT FOR COST BLOCK REACT

*** INPUT PARAMETERS ***

NUMBER OF VESSELS	6
NUMBER OF STAND-BY VESSELS	0
MATERIAL OF CONSTRUCTION	CARBON STEEL
MATERIAL OF CONSTRUCTION FACTOR	2.2500
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
INNER DIAMETER, FT	12.0000
TANGENT-TO-TANGENT LENGTH, FT	75.0000
VELOCITY RATIO	0.5000
PRESSURE, PSI	1600.0000
CORROSION ALLOWANCE, FT	0.0
PLATFORM ALLOWANCE, FT	0.0

*** SIZING RESULTS ***

CALCULATED NUMBER OF VESSELS	6
CALCULATED DIAMETER, FT	12.0000
CALCULATED TANGENT-TO-TANGENT LENGTH, FT	75.0000
TOP SHELL THICKNESS, FT	0.6250
BOTTOM SHELL THICKNESS, FT	0.6250
SHELL WEIGHT PER VESSEL, LB	.98196+06
VOLUME PER VESSEL, CUFT	8934.7103

*** COST RESULTS ***

BASE COST OF VESSEL	\$ 970,256.89
BASE COST OF UNIT	\$ 5,821,541.36
ACTUAL COST OF VESSEL	\$ 2,205,725.52
ACTUAL COST OF UNIT	\$ 13,234,353.13

COST BLOCK SECTION

COST-VERVSSL (CVS11): EXTRACT

REPORT FOR COST BLOCK EXTRACT

*** INPUT PARAMETERS ***

NUMBER OF VESSELS	3
NUMBER OF STAND-BY VESSELS	0
MATERIAL OF CONSTRUCTION	STAINLESS 316L
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	3.8236
INNER DIAMETER, FT	11.0000
TANGENT-TO-TANGENT LENGTH, FT	30.0000
VELOCITY RATIO	0.5000
PRESSURE, PSI	14.7000
CORROSION ALLOWANCE, FT	0.0
PLATFORM ALLOWANCE, FT	0.0

*** SIZING RESULTS ***

CALCULATED NUMBER OF VESSELS	3
CALCULATED DIAMETER, FT	11.0000
CALCULATED TANGENT-TO-TANGENT LENGTH, FT	30.0000
TOP SHELL THICKNESS, FT	0.0208
BOTTOM SHELL THICKNESS, FT	0.0208
SHELL WEIGHT PER VESSEL, LB	.14075+05
VOLUME PER VESSEL, CUFT	3199.4578

*** COST RESULTS ***

BASE COST OF VESSEL	\$ 134,247.66
BASE COST OF UNIT	\$ 402,742.97
ACTUAL COST OF VESSEL	\$ 338,028.03
ACTUAL COST OF UNIT	\$ 1,014,084.09

COST BLOCK SECTION

COST-VERVSSL (CVS11): SEP

REPORT FOR COST BLOCK SEP

*** INPUT PARAMETERS ***

NUMBER OF VESSELS	1
NUMBER OF STAND-BY VESSELS	0
MATERIAL OF CONSTRUCTION	CARBON STEEL
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
INNER DIAMETER, FT	2.0000
TANGENT-TO-TANGENT LENGTH, FT	7.0000
VELOCITY RATIO	0.5000
PRESSURE, PSI	40.0000
CORROSION ALLOWANCE, FT	0.0
PLATFORM ALLOWANCE, FT	0.0

*** SIZING RESULTS ***

CALCULATED NUMBER OF VESSELS	1
CALCULATED DIAMETER, FT	2.0000
CALCULATED TANGENT-TO-TANGENT LENGTH, FT	7.0000
TOP SHELL THICKNESS, FT	0.0208
BOTTOM SHELL THICKNESS, FT	0.0208
SHELL WEIGHT PER VESSEL, LB	555.1629
VOLUME PER VESSEL, CUFT	24.0856

*** COST RESULTS ***

BASE COST OF VESSEL	\$ 8,788.96
BASE COST OF UNIT	\$ 8,788.96
ACTUAL COST OF VESSEL	\$ 10,382.44
ACTUAL COST OF UNIT	\$ 10,382.44

COST BLOCK SECTION

=====

COST-VERVSSL (CVS11): GASSEP

REPORT FOR COST BLOCK GASSEP

*** INPUT PARAMETERS ***

NUMBER OF VESSELS	1
NUMBER OF STAND-BY VESSELS	0
MATERIAL OF CONSTRUCTION	CARBON STEEL
CAPACITY SCALE FACTOR	1.0000
COST ADJUSTMENT FACTOR	1.0000
INNER DIAMETER, FT	8.0000
TANGENT-TO-TANGENT LENGTH, FT	28.0000
VELOCITY RATIO	0.5000
PRESSURE, PSI	25.0000
CORROSION ALLOWANCE, FT	0.0
PLATFORM ALLOWANCE, FT	0.0

*** SIZING RESULTS ***

CALCULATED NUMBER OF VESSELS	1
CALCULATED DIAMETER, FT	8.0000
CALCULATED TANGENT-TO-TANGENT LENGTH, FT	28.0000
TOP SHELL THICKNESS, FT	0.0208
BOTTOM SHELL THICKNESS, FT	0.0208
SHELL WEIGHT PER VESSEL, LB	8882.6067
VOLUME PER VESSEL, CUFT	1541.4784

*** COST RESULTS ***

BASE COST OF VESSEL	\$ 26,479.94
BASE COST OF UNIT	\$ 26,479.94
ACTUAL COST OF VESSEL	\$ 38,315.52
ACTUAL COST OF UNIT	\$ 38,315.52

ECONOMIC EVALUATION SECTION

INSTALLATION COST SUMMARY

YEAR: 1984, QUARTER: 1

LOCATION: MISSING

SECTION:	PHSOLVAN	WETAIR	TOTAL
TOTAL EQUIPMENT COST	\$ 2,782,854	\$ 32,995,041	\$ 35,777,895
COMMODITY MATERIAL COST			
PIPING	\$ 802,905	\$ 5,732,714	\$ 6,535,619
CONCRETE	130,814	895,198	1,026,012
STEEL	102,080	600,766	702,846
INSTRUMENTS	154,398	1,163,470	1,317,867
ELECTRICAL	78,547	481,062	559,610
INSULATION	105,606	692,746	798,351
PAINTING	17,081	100,723	117,803
MISCELLANEOUS	0	0	0
TOTAL	\$ 1,391,431	\$ 9,666,678	\$ 11,058,109
COMMODITY LABOR COST			
PIPING	\$ 0	\$ 0	\$ 0
CONCRETE	0	0	0
STEEL	0	0	0
INSTRUMENTS	0	0	0
ELECTRICAL	0	0	0
INSULATION	0	0	0
PAINTING	0	0	0
MISCELLANEOUS	0	0	0
TOTAL	\$ 1,637,073	\$ 11,373,222	\$ 13,010,295
COMMODITY COST	\$ 3,028,504	\$ 21,039,900	\$ 24,068,403
BUILDING MATERIAL	\$ 0	\$ 0	\$ 0
TESTING MATERIAL	4,421	44,978	49,400
ADDL-INST. MATERIAL	0	0	0
CAPITALIZED SPARES	27,829	329,950	357,779
BUILDING LABOR	\$ 0	\$ 0	\$ 0
TESTING LABOR	134,496	1,368,265	1,502,761
ADDL-INST. LABOR	0	0	0
SETTING LABOR	0	0	0
TOTAL EQUIPMENT COST	\$ 2,782,854	\$ 32,995,041	\$ 35,777,895
TOTAL MATERIAL COST	1,423,681	10,041,606	11,465,287
TOTAL LABOR COST	1,771,568	12,741,488	14,513,056
SECTION INSTALLED COST	\$ 5,978,103	\$ 55,778,135	\$ 61,756,238

ECONOMIC EVALUATION SECTION

INSTALLATION LABOR HOUR SUMMARY

YEAR: 1984, QUARTER: 1

LOCATION: MISSING

SECTION	PHSOLVAN	WETAIR	TOTAL
COMMODITY LABOR HOUR			
PIPING	0	0	0
CONCRETE	0	0	0
STEEL	0	0	0
INSTRUMENTS	0	0	0
ELECTRICAL	0	0	0
INSULATION	0	0	0
PAINTING	0	0	0
MISCELLANEOUS	0	0	0
TOTAL	109138	758215	867353
BUILDING LABOR HOUR	0	0	0
TESTING LABOR HOUR	8966	91218	100184
ADDL-INST. LABOR HOUR	0	0	0
SETTING LABOR HOUR	0	0	0
GRAND TOTAL	118105	849433	967538

ECONOMIC EVALUATION SECTION

SECTION INSTALLED COST SUMMARY

YEAR: 1984, QUARTER: 1

LOCATION: MISSING

SECTION:PHSOLVAN

EQUIPMENT:

EQUIPMENT COST
\$ 2,782,854

COMMODITY:

	MATERIAL COST	LABOR COST	LABOR HOURS
PIPING	\$ 802,905	\$ 0	0
CONCRETE	130,814	0	0
STEEL	102,080	0	0
INSTRUMENTS	154,398	0	0
ELECTRICAL	78,547	0	0
INSULATION	105,606	0	0
PAINTING	17,081	0	0
MISCELLANEOUS	0	0	0
TOTAL	\$ 1,391,431	\$ 1,637,073	109,138

OTHERS:

	MATERIAL COST	LABOR COST	LABOR HOURS
BUILDING	\$ 0	\$ 0	0
TESTING	4,421	134,496	8,966
ADDITIONAL	0	0	0
SPARE	27,829	0	0
SETTING LABOR		0	0
TOTAL	\$ 32,250	\$ 134,496	8,967

SECTION INSTALLED COST

EQUIPMENT	\$ 2,782,854	
COMMODITY MATERIAL	1,391,431	
OTHERS MATERIAL	32,250	
SECTION EQUIP. & MATERIAL		\$ 4,206,535
COMMODITY LABOR	\$ 1,637,073	
SETTING LABOR	0	
OTHERS LABOR (EX. SET'G LAB)	134,496	
SECTION LABOR		1,771,568
SECTION INSTALLED		<u>\$ 5,978,103</u>

ECONOMIC EVALUATION SECTION

EQUIPMENT INSTALLATION REPORT

YEAR: 1984, QUARTER: 1

LOCATION: MISSING

SECTION:PHSOLVAN

COST BLOCK ID	PUMP1	PUMP2	PUMP3	PUMP4
QUANTITY	1	1	1	1
STANDBY	1	1	1	1
BASE EQUIP. COST \$	4,516 \$	1,650 \$	13,557 \$	2,392
ADJUSTED EQUIP. COST \$	8,132 \$	3,757 \$	17,247 \$	4,747
COMMODITY MATL. COST:				
PIPING \$	2,614 \$	955 \$	7,846 \$	1,385
CONCRETE	346	126	1,039	183
STEEL	0	0	0	0
INSTRUMENTS	260	95	779	138
ELECTRICAL	2,683	980	8,054	1,421
INSULATION	216	79	650	115
PAINTING	69	25	208	37
MISCELLANEOUS	0	0	0	0
TOTAL \$	6,189 \$	2,261 \$	18,577 \$	3,279
COMMODITY LABOR COST:				
PIPING \$	0 \$	0 \$	0 \$	0
CONCRETE	0	0	0	0
STEEL	0	0	0	0
INSTRUMENTS	0	0	0	0
ELECTRICAL	0	0	0	0
INSULATION	0	0	0	0
PAINTING	0	0	0	0
MISCELLANEOUS	0	0	0	0
TOTAL \$	7,281 \$	2,660 \$	21,857 \$	3,857
SETTING LABOR COST \$	0 \$	0 \$	0 \$	0
COMMODITY LABOR HOUR:				
PIPING	0	0	0	0
CONCRETE	0	0	0	0
STEEL	0	0	0	0
INSTRUMENTS	0	0	0	0
ELECTRICAL	0	0	0	0
INSULATION	0	0	0	0
PAINTING	0	0	0	0
MISCELLANEOUS	0	0	0	0
TOTAL	485	177	1,457	257
SETTING LABOR HOUR	0	0	0	0
EQUIP. & MATL. COST \$	22,453 \$	9,775 \$	53,072 \$	12,772
LABOR COST \$	7,281 \$	2,660 \$	21,857 \$	3,857

ECONOMIC EVALUATION SECTION

EQUIPMENT INSTALLATION REPORT (CONTINUED)

YEAR: 1984, QUARTER: 1

LOCATION: MISSING

SECTION:PHSOLVAN

COST BLOCK ID	PUMP6	HEATX1	HEATX3	HEATX4
QUANTITY	1	1	1	1
STANDBY	1	0	0	0
BASE EQUIP. COST \$	1,843 \$	60,961 \$	17,644 \$	11,961
ADJUSTED EQUIP. COST \$	4,126 \$	146,085 \$	17,644 \$	11,961
COMMODITY MATL. COST:				
PIPING \$	1,067 \$	26,638 \$	7,710 \$	5,227
CONCRETE	141	2,979	862	585
STEEL	0	1,811	524	355
INSTRUMENTS	106	5,959	1,725	1,169
ELECTRICAL	1,095	1,168	338	229
INSULATION	88	2,862	828	562
PAINTING	28	292	85	57
MISCELLANEOUS	0	0	0	0
TOTAL	\$ 2,525 \$	\$ 41,710 \$	\$ 12,072 \$	\$ 8,184
COMMODITY LABOR COST:				
PIPING \$	0 \$	0 \$	0 \$	0
CONCRETE	0	0	0	0
STEEL	0	0	0	0
INSTRUMENTS	0	0	0	0
ELECTRICAL	0	0	0	0
INSULATION	0	0	0	0
PAINTING	0	0	0	0
MISCELLANEOUS	0	0	0	0
TOTAL	\$ 2,971 \$	\$ 49,073 \$	\$ 14,203 \$	\$ 9,628
SETTING LABOR COST \$	0 \$	0 \$	0 \$	0
COMMODITY LABOR HOUR:				
PIPING	0	0	0	0
CONCRETE	0	0	0	0
STEEL	0	0	0	0
INSTRUMENTS	0	0	0	0
ELECTRICAL	0	0	0	0
INSULATION	0	0	0	0
PAINTING	0	0	0	0
MISCELLANEOUS	0	0	0	0
TOTAL	198	3,272	947	642
SETTING LABOR HOUR	0	0	0	0
EQUIP. & MATL. COST \$	10,777 \$	187,795 \$	29,716 \$	20,145
LABOR COST \$	2,971 \$	49,073 \$	14,203 \$	9,628

ECONOMIC EVALUATION SECTION

EQUIPMENT INSTALLATION REPORT (CONTINUED)

YEAR: 1984, QUARTER: 1

LOCATION: MISSING

SECTION:PHSOLVAN

COST BLOCK ID	TOWER1	TOWER2	SEP	EXTRACT
QUANTITY	1	1	1	3
STANDBY	0	0	0	0
BASE EQUIP. COST	\$ 676,171	\$ 153,802	\$ 8,789	\$ 134,248
ADJUSTED EQUIP. COST	\$ 1,139,560	\$ 234,604	\$ 10,382	\$ 338,028
COMMODITY MATL. COST:				
PIPING	\$ 386,183	\$ 88,431	\$ 5,053	\$ 231,563
CONCRETE	64,148	14,738	842	38,594
STEEL	51,189	11,791	674	30,875
INSTRUMENTS	74,515	16,949	969	44,383
ELECTRICAL	31,750	7,369	421	19,297
INSULATION	51,837	11,791	674	30,875
PAINTING	8,423	1,916	109	5,017
MISCELLANEOUS	0	0	0	0
TOTAL	\$ 668,045	\$ 152,985	\$ 8,742	\$ 400,604
COMMODITY LABOR COST:				
PIPING	\$ 0	\$ 0	\$ 0	\$ 0
CONCRETE	0	0	0	0
STEEL	0	0	0	0
INSTRUMENTS	0	0	0	0
ELECTRICAL	0	0	0	0
INSULATION	0	0	0	0
PAINTING	0	0	0	0
MISCELLANEOUS	0	0	0	0
TOTAL	\$ 785,981	\$ 179,993	\$ 10,286	\$ 471,327
SETTING LABOR COST	\$ 0	\$ 0	\$ 0	\$ 0
COMMODITY LABOR HOUR:				
PIPING	0	0	0	0
CONCRETE	0	0	0	0
STEEL	0	0	0	0
INSTRUMENTS	0	0	0	0
ELECTRICAL	0	0	0	0
INSULATION	0	0	0	0
PAINTING	0	0	0	0
MISCELLANEOUS	0	0	0	0
TOTAL	52,399	12,000	686	31,422
SETTING LABOR HOUR	0	0	0	0
EQUIP. & MATL. COST	\$ 1,807,605	\$ 387,588	\$ 19,125	\$ 1,414,688
LABOR COST	\$ 785,981	\$ 179,993	\$ 10,286	\$ 471,327

ECONOMIC EVALUATION SECTION

EQUIPMENT INSTALLATION REPORT (CONTINUED)

YEAR: 1984, QUARTER: 1

LOCATION: MISSING

SECTION: PHSOLVAN

COST BLOCK ID	TOTAL
QUANTITY	
STANDBY	
BASE EQUIP. COST	\$ 1,087,533
ADJUSTED EQUIP. COST	\$ 1,936,272

COMMODITY MATL. COST:	
PIPING	\$ 764,672
CONCRETE	124,585
STEEL	97,219
INSTRUMENTS	147,046
ELECTRICAL	74,807
INSULATION	100,577
PAINTING	16,267
MISCELLANEOUS	0

TOTAL	\$ 1,325,172

COMMODITY LABOR COST:	
PIPING	\$ 0
CONCRETE	0
STEEL	0
INSTRUMENTS	0
ELECTRICAL	0
INSULATION	0
PAINTING	0
MISCELLANEOUS	0

TOTAL	\$ 1,559,117

SETTING LABOR COST	\$ 0
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COMMODITY LABOR HOUR:	
PIPING	0
CONCRETE	0
STEEL	0
INSTRUMENTS	0
ELECTRICAL	0
INSULATION	0
PAINTING	0
MISCELLANEOUS	0

TOTAL	103,941

SETTING LABOR HOUR	0
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EQUIP. & MATL. COST	\$ 3,975,509
LABOR COST	\$ 1,559,117

ECONOMIC EVALUATION SECTION

SECTION INSTALLED COST SUMMARY

YEAR: 1984, QUARTER: 1

LOCATION: MISSING

SECTION:WETAIR

EQUIPMENT:

EQUIPMENT COST
\$ 32,995,041

COMMODITY:

	MATERIAL COST	LABOR COST	LABOR HOURS
PIPING	\$ 5,732,714 \$	0	0
CONCRETE	895,198	0	0
STEEL	600,766	0	0
INSTRUMENTS	1,163,470	0	0
ELECTRICAL	481,062	0	0
INSULATION	692,746	0	0
PAINTING	100,723	0	0
MISCELLANEOUS	0	0	0
TOTAL	\$ 9,666,678 \$	11,373,222	758,215

OTHERS:

	MATERIAL COST	LABOR COST	LABOR HOURS
BUILDING	\$ 0 \$	0	0
TESTING	44,978	1,368,265	91,218
ADDITIONAL	0	0	0
SPARE	329,950		
SETTING LABOR		0	0
TOTAL	\$ 374,929 \$	1,368,266	91,218

SECTION INSTALLED COST

EQUIPMENT	\$ 32,995,041	
COMMODITY MATERIAL	9,666,678	
OTHERS MATERIAL	374,929	
SECTION EQUIP. & MATERIAL		\$ 43,036,647
COMMODITY LABOR	\$ 11,373,222	
SETTING LABOR	0	
OTHERS LABOR (EX. SET'G LAB)	1,368,265	
SECTION LABOR		12,741,488
SECTION INSTALLED		<u>\$ 55,778,135</u>

ECONOMIC EVALUATION SECTION

EQUIPMENT INSTALLATION REPORT

YEAR: 1984, QUARTER: 1

LOCATION: MISSING

SECTION:WETAIR

COST BLOCK ID.	PUMP5A	PUMP5B	PUMP5C	COMP
QUANTITY	3	3	3	1
STANDBY	1	0	0	0
BASE EQUIP. COST	\$ 28,707	\$ 28,707	\$ 28,699	\$ 42,507
ADJUSTED EQUIP. COST	\$ 59,308	\$ 59,308	\$ 59,205	\$ 54,834
COMMODITY MATL. COST:				
PIPING	\$ 33,231	\$ 24,923	\$ 24,917	\$ 7,332
CONCRETE	4,401	3,301	3,300	4,073
STEEL	0	0	0	0
INSTRUMENTS	3,301	2,476	2,475	1,629
ELECTRICAL	34,112	25,584	25,577	815
INSULATION	2,751	2,063	2,063	0
PAINTING	880	660	660	0
MISCELLANEOUS	0	0	0	0
TOTAL	\$ 78,677	\$ 59,007	\$ 58,991	\$ 13,849
COMMODITY LABOR COST:				
PIPING	\$ 0	\$ 0	\$ 0	\$ 0
CONCRETE	0	0	0	0
STEEL	0	0	0	0
INSTRUMENTS	0	0	0	0
ELECTRICAL	0	0	0	0
INSULATION	0	0	0	0
PAINTING	0	0	0	0
MISCELLANEOUS	0	0	0	0
TOTAL	\$ 92,566	\$ 69,425	\$ 69,406	\$ 16,294
SETTING LABOR COST	\$ 0	\$ 0	\$ 0	\$ 0
COMMODITY LABOR HOUR:				
PIPING	0	0	0	0
CONCRETE	0	0	0	0
STEEL	0	0	0	0
INSTRUMENTS	0	0	0	0
ELECTRICAL	0	0	0	0
INSULATION	0	0	0	0
PAINTING	0	0	0	0
MISCELLANEOUS	0	0	0	0
TOTAL	6,171	4,628	4,627	1,086
SETTING LABOR HOUR	0	0	0	0
EQUIP. & MATL. COST	\$ 315,907	\$ 236,930	\$ 236,606	\$ 68,683
LABOR COST	\$ 92,566	\$ 69,425	\$ 69,406	\$ 16,294

ECONOMIC EVALUATION SECTION

EQUIPMENT INSTALLATION REPORT (CONTINUED)

YEAR: 1984, QUARTER: 1

LOCATION: MISSING

SECTION:WETAIR

COST BLOCK ID	HEATX2	HEATX5	HEATX6	REACT
QUANTITY	11	3	16	6
STANDBY	0	0	0	0
BASE EQUIP. COST	\$ 147,402	\$ 119,817	\$ 148,474	\$ 970,257
ADJUSTED EQUIP. COST	\$ 878,626	\$ 119,817	\$ 390,059	\$ 2,205,726

COMMODITY MATL. COST:

PIPING	\$ 708,521	\$ 157,070	\$ 1,038,069	\$ 3,324,869
CONCRETE	79,242	17,567	116,100	552,285
STEEL	48,167	10,678	70,571	440,713
INSTRUMENTS	158,485	35,134	232,200	641,544
ELECTRICAL	31,075	6,889	45,529	273,353
INSULATION	76,135	16,878	111,547	446,291
PAINTING	7,769	1,722	11,382	72,522
MISCELLANEOUS	0	0	0	0

TOTAL	\$ 1,109,394	\$ 245,939	\$ 1,625,398	\$ 5,751,578
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COMMODITY LABOR COST:

PIPING	\$ 0	\$ 0	\$ 0	\$ 0
CONCRETE	0	0	0	0
STEEL	0	0	0	0
INSTRUMENTS	0	0	0	0
ELECTRICAL	0	0	0	0
INSULATION	0	0	0	0
PAINTING	0	0	0	0
MISCELLANEOUS	0	0	0	0

TOTAL	\$ 1,305,245	\$ 289,357	\$ 1,912,344	\$ 6,766,955
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SETTING LABOR COST	\$ 0	\$ 0	\$ 0	\$ 0
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COMMODITY LABOR HOUR:

PIPING	0	0	0	0
CONCRETE	0	0	0	0
STEEL	0	0	0	0
INSTRUMENTS	0	0	0	0
ELECTRICAL	0	0	0	0
INSULATION	0	0	0	0
PAINTING	0	0	0	0
MISCELLANEOUS	0	0	0	0

TOTAL	87,016	19,290	127,490	451,130
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SETTING LABOR HOUR	0	0	0	0
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EQUIP. & MATL. COST	\$ 10,774,281	\$ 605,389	\$ 7,866,340	\$ 18,985,931
LABOR COST	\$ 1,305,245	\$ 289,357	\$ 1,912,344	\$ 6,766,955

ECONOMIC EVALUATION SECTION

EQUIPMENT INSTALLATION REPORT (CONTINUED)

YEAR: 1984, QUARTER: 1

LOCATION: MISSING

SECTION: WETAIR

COST BLOCK ID		GASSEP		FURNACE		TOTAL
QUANTITY		1		1		
STANDBY		0		0		
BASE EQUIP. COST	\$	26,480	\$	727,982	\$	2,269,033
ADJUSTED EQUIP. COST	\$	38,316	\$	1,238,298	\$	5,103,495

COMMODITY MATL. COST:

PIPING	\$	15,225	\$	125,569	\$	5,459,727
CONCRETE		2,538		69,761		852,569
STEEL		2,030		0		572,158
INSTRUMENTS		2,918		27,904		1,108,066
ELECTRICAL		1,269		13,952		458,155
INSULATION		2,030		0		659,758
PAINTING		330		0		95,926
MISCELLANEOUS		0		0		0

TOTAL	\$	26,339	\$	237,187	\$	9,206,360
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COMMODITY LABOR COST:

PIPING	\$	0	\$	0	\$	0
CONCRETE		0		0		0
STEEL		0		0		0
INSTRUMENTS		0		0		0
ELECTRICAL		0		0		0
INSULATION		0		0		0
PAINTING		0		0		0
MISCELLANEOUS		0		0		0

TOTAL	\$	30,989	\$	279,059	\$	10,831,640
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SETTING LABOR COST	\$	0	\$	0	\$	0
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COMMODITY LABOR HOUR:

PIPING		0		0		0
CONCRETE		0		0		0
STEEL		0		0		0
INSTRUMENTS		0		0		0
ELECTRICAL		0		0		0
INSULATION		0		0		0
PAINTING		0		0		0
MISCELLANEOUS		0		0		0

TOTAL		2,066		18,604		722,109
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SETTING LABOR HOUR		0		0		0
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EQUIP. & MATL. COST	\$	64,655	\$	1,475,484	\$	40,630,208
LABOR COST	\$	30,989	\$	279,059	\$	10,831,640

ECONOMIC EVALUATION SECTION

CAPITAL INVESTMENT REPORT
ASPEN METHOD

** PROJECT CONSTRUCTION COSTS FOR LOCATION, 1984 QUARTER : 1 **

	MATERIAL COST	LABOR COST	LABOR HOUR
PROCESS UNITS	\$ 47,243,182	\$ 14,513,056	967,537
UTILITY UNITS	0	0	0
RECVNG , SHIPPING & STOR	256,676	82,444	3,926
SERVICE BUILDING	3,756,781	1,714,251	114,283
SERVICE SYST & DISTRBUTN	2,965,539	982,837	65,522
ADDITIONAL DIRECT	0	0	0
SUBTOTAL	\$ 54,222,177	\$ 17,292,588	1,151,269
SITE DEVELOPMENT	570,780	17,363	1,158
FREIGHT	1,093,447		
TOTAL DIRECT COST	\$ 55,886,405	\$ 17,309,952	1,152,426
FIELD INDIRECT	4,673,687	8,654,976	480,832
FIELD CONSTR COST	\$ 60,560,092	\$ 25,964,927	
FIELD CONSTR LAB HRS			1,633,258

** PROJECT FIXED INVESTMENT FOR LOCATION, 1984 QUARTER : 1 **

TOTAL FIELD CONSTRUCTION COST		\$ 86,525,019
PROJECT MANAGEMENT COST	\$ 5,426,146	
ENGINEERING & HOME OFFICE COST	4,521,788	
FEES, PERMITS & INSURANCE	7,234,861	
ADDITIONAL DEPRECIABLE COST		0
SUBTOTAL - INDIRECTS		17,182,795
TOTAL DEPRECIABLE (EX. CONT.)		\$ 103,707,815
PROCESS CONTINGENCY	\$ 10,370,781	
PROJECT DEFINITION CONTINGENCY	10,370,781	
TOTAL CONTINGENCY		20,741,563
TOTAL DEPRECIABLE CAPITAL		\$ 124,449,378
LAND	\$ 1,269,679	
ROYALTY & EXPENSES	0	
ADDITIONAL NON-DEPRECIABLE	0	
SUBTOTAL (NON-DEPRECIABLE COSTS)		1,269,679
TOTAL FIXED INVESTMENT		\$ 125,719,057

APPENDIX C

PRPIEA PROFITABILITY ANALYSES RESULTS

2.694 2.470 2.245 2.021 1.796 1.572 1.347 1.123
 0.898 0.674 0.449 0.225
 SUM OF DEPRECIATION ALLOWANCES 34.355

PAYOUT TABULATION
 FEED PRICE: 0.50 CASE: PHENOSOLVIAN ALONE WITH NEGATIVE FEDERAL TAXES
 PRODUCT PRICE: 0.2900 RATE OF RETURN ON EQUITY 40.50%
 WORKING CAPITAL: 3.905 DEBT AT 12.00%
 START-UP COSTS: 0.033 D/E RATIO 0.70/0.30
 TRIAL 19

	1	2	3	4	5	6	7	8	9	10	11
	CPSTRT	EFFNCY	INCOME	PLTEXP	STLTAX	TXCRDT	FEDTAX	CASHFL	TOTRTN	AMORTZ	CAPEND
1	10.657	0.000	0.000	0.000	0.000	0.000	-0.412	0.412	2.190	-1.778	12.435
2	29.612	0.000	0.000	0.000	0.000	0.000	-1.144	1.144	6.085	-4.941	34.554
3	41.424	0.000	0.000	0.000	0.000	0.000	-1.601	1.601	8.513	-6.912	48.337
4	51.891	0.500	14.820	7.641	0.687	2.405	-3.179	9.671	10.664	-0.992	52.883
5	52.883	0.800	23.712	9.839	0.862	0.000	2.289	10.722	10.668	-0.146	53.029
6	53.029	0.900	26.675	10.583	0.935	0.000	3.374	11.783	10.898	0.886	52.143
7	52.143	0.950	28.157	10.955	0.977	0.000	4.003	12.223	10.716	1.507	50.636
8	50.636	0.950	28.157	10.955	0.988	0.000	4.159	12.055	10.406	1.649	48.987
9	48.987	0.950	28.157	10.955	0.999	0.000	4.321	11.882	10.067	1.816	47.171
10	47.171	0.950	28.157	10.955	1.010	0.000	4.490	11.703	9.694	2.009	45.162
11	45.162	0.950	28.157	10.955	1.022	0.000	4.665	11.516	9.281	2.235	42.928
12	42.928	0.950	28.157	10.955	1.034	0.000	4.849	11.319	8.822	2.498	40.430
13	40.430	0.950	28.157	10.955	1.047	0.000	5.043	11.112	8.308	2.804	37.626
14	37.626	0.950	28.157	10.955	1.061	0.000	5.248	10.893	7.732	3.161	34.465
15	34.465	0.950	28.157	10.955	1.076	0.000	5.467	10.660	7.083	3.577	30.888
16	30.888	0.950	28.157	10.955	1.092	0.000	5.701	10.410	6.347	4.062	26.825
17	26.825	0.950	28.157	10.955	1.109	0.000	5.954	10.141	5.513	4.628	22.198
18	22.198	0.950	28.157	10.955	1.127	0.000	6.227	9.848	4.562	5.287	16.911
19	16.911	0.950	28.157	10.955	1.147	0.000	6.526	9.530	3.475	6.055	10.856
20	10.856	0.950	28.157	10.955	1.169	0.000	6.853	9.181	2.231	6.950	3.906

THE LAST ENTRY IN THE FINAL YEAR SHOULD BE EQUAL TO THE WORKING CAPITAL INVESTMENT PLUS SALVAGE VALUE

WORKING CAPITAL RECOVERED INTACT AT END OF PROJECT LIFE = 3.905
 SALVAGE VALUE RECOVERED INTACT AT END OF PROJECT LIFE = 0.000
 TOTAL 3.905

KEY TO CALCULATIONS IS AS FOLLOWS.

COLUMN 1 = CAPITAL INVESTMENT OUTSTANDING AT START OF YEAR INCLUDING WORKING CAPITAL
 COLUMN 2 = ONSTREAM EFFICIENCY (PLANT OPERATING FACTOR)
 COLUMN 3 = GROSS INCOME FROM SALES
 COLUMN 4 = OPERATING EXPENSES EXCLUDING TAXES
 COLUMN 5 = STATE AND LOCAL TAXES
 COLUMN 6 = FEDERAL INVESTMENT TAX CREDIT
 COLUMN 7 = NET FEDERAL INCOME TAX PAID
 COLUMN 8 = CASH FLOW AFTER TAXES
 COLUMN 9 = INTEREST ON DEBT AND RETURN ON EQUITY BASED ON COLUMN 1

COLUMN 10 = REDUCTION OF OUTSTANDING CAPITAL INVESTMENT, COLUMN 8-9
 COLUMN 11 = OUTSTANDING CAPITAL INVESTMENT AT END OF YEAR INCL WORKING CAPITAL. COL 1-10
 COLUMN 11 PLUS NEW INVESTMENT = NEXT ENTRY OF COLUMN 1

ADDITIONAL DETAILS OF CALCULATION PHENOSOLVAN ALONE WITH NEGATIVE FEDERAL TAXES

	1	2	3	4	5	6	7	8	9	10	11
	CPSTRT	DBSTRT	EQSTRT	DBTADD	INTRST	TAXABL	STXABL	EQYRTN	PROPTY	DBTPAY	EQTPAY
1	10.657	7.460	3.197	7.460	0.895	-0.895	-0.895	1.295	0.000	-1.245	-0.533
2	29.612	20.729	8.884	12.024	2.487	-2.487	-2.487	3.598	0.000	-3.459	-1.482
3	41.424	28.997	12.427	4.810	3.480	-3.480	-3.480	5.033	0.000	-4.838	-2.074
4	51.891	36.324	15.567	2.488	4.359	-1.684	-0.997	6.305	0.687	-0.695	-0.298
5	52.883	37.018	15.865	0.000	4.442	4.975	5.838	6.425	0.687	-0.102	-0.044
6	53.029	37.120	15.909	0.000	4.454	7.335	8.270	6.443	0.687	-0.620	0.266
7	52.143	36.500	15.643	0.000	4.380	8.702	9.679	6.335	0.687	1.055	0.452
8	50.636	35.445	15.191	0.000	4.253	9.042	10.030	6.152	0.687	1.155	0.495
9	48.987	34.291	14.696	0.000	4.115	9.394	10.393	5.952	0.687	1.271	0.545
10	47.171	33.020	14.151	0.000	3.794	9.760	10.770	5.731	0.687	1.406	0.603
11	45.162	31.614	13.549	0.000	3.606	10.142	11.164	5.487	0.687	1.564	0.670
12	42.928	30.049	12.878	0.000	3.396	10.542	11.576	5.216	0.687	1.748	0.749
13	40.430	28.301	12.129	0.000	3.161	10.963	12.010	4.912	0.687	1.963	0.841
14	37.626	26.338	11.288	0.000	2.895	11.409	12.470	4.572	0.687	2.213	0.948
15	34.465	24.126	10.340	0.000	2.595	11.885	12.960	4.188	0.687	2.504	1.073
16	30.888	21.621	9.266	0.000	2.253	12.394	13.485	3.753	0.687	2.844	1.219
17	26.825	18.778	8.048	0.000	1.865	12.943	14.051	3.259	0.687	3.239	1.388
18	22.198	15.538	6.659	0.000	1.420	13.538	14.665	2.697	0.687	3.701	1.586
19	16.911	11.837	5.073	0.000	0.912	14.186	15.333	2.055	0.687	4.238	1.816
20	10.856	7.599	3.257	0.000		14.897	16.066	1.319	0.687	4.865	2.085

KEY TO CALCULATIONS IS AS FOLLOWS.

COLUMN 1 = TOTAL CAPITAL AT START OF YEAR
 COLUMN 2 = DEBT OUTSTANDING AT START OF YEAR
 COLUMN 3 = EQUITY CAPITAL AT START OF YEAR
 COLUMN 4 = AMOUNT ADDED TO DEBT
 COLUMN 5 = INTEREST ON DEBT
 COLUMN 6 = FEDERAL TAXABLE INCOME
 COLUMN 7 = STATE TAXABLE INCOME
 COLUMN 8 = AFTER-TAX RETURN ON EQUITY
 COLUMN 9 = LOCAL PROPERTY TAX PAID
 COLUMN 10 = RETIREMENT OF DEBT
 COLUMN 11 = RETIREMENT OF EQUITY

2.694	2.470	2.245	2.021	1.796	1.572	1.347	1.123
0.898	0.674	0.449	0.225				
SUM OF DEPRECIATION ALLOWANCES 34.355							

PAYOUT TABULATION
 FEED PRICE: 0.50 CASE: PHENOSOLVIAN ALONE W/O NEGATIVE FEDERAL TAXES
 PRODUCT PRICE: 0.2900 RATE OF RETURN ON EQUITY 36.32%
 WORKING CAPITAL: 3.905 DEBT AT 12.00%
 START-UP COSTS: 0.033 D/E RATIO 0.70/0.30
 TRIAL 17

	1	2	3	4	5	6	7	8	9	10	11
	CPSTRT	EFFNCY	INCOME	PLTEXP	STLIAX	TXCRDT	FEDTAX	CASHFL	TOTRTN	AMORTZ	CAPEND
1	10.657	0.000	0.000	0.000	0.000	0.000	0.000	0.000	2.056	-2.056	12.713
2	29.890	0.000	0.000	0.000	0.000	0.000	0.000	0.000	5.768	-5.768	35.658
3	42.529	0.000	0.000	0.000	0.000	0.000	0.000	0.000	8.206	-8.206	50.735
4	54.290	0.500	14.820	7.641	0.687	0.000	0.000	6.492	10.476	-3.984	58.274
5	58.274	0.800	23.712	9.839	0.849	2.087	0.000	13.024	11.245	1.779	56.495
6	56.495	0.900	26.675	10.583	0.926	0.318	0.000	12.240	10.901	1.700	55.155
7	55.155	0.950	28.157	10.955	0.970	0.000	3.890	12.343	10.643	1.700	53.455
8	53.455	0.950	28.157	10.955	0.981	0.000	4.054	12.168	10.315	1.853	51.602
9	51.602	0.950	28.157	10.955	0.992	0.000	4.223	11.987	9.957	2.030	49.572
10	49.572	0.950	28.157	10.955	1.004	0.000	4.400	11.799	9.565	2.233	47.339
11	47.339	0.950	28.157	10.955	1.017	0.000	4.584	11.603	9.134	2.468	44.870
12	44.870	0.950	28.157	10.955	1.029	0.000	4.776	11.397	8.658	2.739	42.132
13	42.132	0.950	28.157	10.955	1.043	0.000	4.979	11.180	8.130	3.051	39.081
14	39.081	0.950	28.157	10.955	1.058	0.000	5.194	10.952	7.541	3.410	35.670
15	35.670	0.950	28.157	10.955	1.073	0.000	5.422	10.708	6.883	3.825	31.845
16	31.845	0.950	28.157	10.955	1.089	0.000	5.665	10.448	6.145	4.303	27.542
17	27.542	0.950	28.157	10.955	1.107	0.000	5.927	10.169	5.314	4.855	22.687
18	22.687	0.950	28.157	10.955	1.126	0.000	6.209	9.868	4.378	5.490	17.197
19	17.197	0.950	28.157	10.955	1.146	0.000	6.515	9.541	3.318	6.223	10.973
20	10.973	0.950	28.157	10.955	1.169	0.000	6.848	9.186	2.117	7.068	3.905

THE LAST ENTRY IN THE FINAL YEAR SHOULD BE EQUAL TO THE WORKING CAPITAL INVESTMENT PLUS SALVAGE VALUE

WORKING CAPITAL RECOVERED INTACT AT END OF PROJECT LIFE = 3.905
 SALVAGE VALUE RECOVERED INTACT AT END OF PROJECT LIFE = 0.000
 TOTAL 3.905

KEY TO CALCULATIONS IS AS FOLLOWS.

COLUMN 1 = CAPITAL INVESTMENT OUTSTANDING AT START OF YEAR INCLUDING WORKING CAPITAL
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 COLUMN 9 = INTEREST ON DEBT AND RETURN ON EQUITY BASED ON COLUMN 1

COLUMN 10 = REDUCTION OF OUTSTANDING CAPITAL INVESTMENT, COLUMN 8-9
 COLUMN 11 = OUTSTANDING CAPITAL INVESTMENT AT END OF YEAR INCL WORKING CAPITAL. COL 1-10
 COLUMN 11 PLUS NEW INVESTMENT = NEXT ENTRY OF COLUMN 1

ADDITIONAL DETAILS OF CALCULATION PHENOSOLVAN ALONE W/O NEGATIVE FEDERAL TAXES

	1	2	3	4	5	6	7	8	9	10	11
	CPSTRT	DBSTRT	EQSTRT	DBIADD	INTRST	TAXABL	STXABL	EQYRTN	PROPTY	DBTPAY	EQTPAY
1	10.657	7.460	3.197	7.460	0.895	-0.895	-0.895	1.161	0.000	-1.439	-0.617
2	29.890	20.923	8.967	12.024	2.511	-2.511	-2.511	3.257	0.000	-4.037	-1.730
3	42.529	29.770	12.759	4.810	3.572	-3.572	-3.572	4.634	0.000	-5.744	-2.462
4	54.290	38.003	16.287	2.488	4.560	-1.886	-1.198	5.915	0.687	-2.789	-1.195
5	58.274	40.792	17.482	0.000	4.395	4.536	5.385	6.350	0.687	1.246	0.534
6	56.495	39.546	16.948	0.000	4.746	7.053	7.979	6.156	0.687	0.937	0.402
7	55.155	38.609	16.547	0.000	4.633	8.456	9.426	6.010	0.687	1.190	0.510
8	53.455	37.419	16.037	0.000	4.490	8.813	9.793	5.824	0.687	1.297	0.556
9	51.602	36.121	15.481	0.000	4.335	9.181	10.174	5.623	0.687	1.421	0.609
10	49.572	34.700	14.872	0.000	4.164	9.565	10.569	5.401	0.687	1.563	0.670
11	47.339	33.137	14.202	0.000	3.976	9.964	10.981	5.158	0.687	1.728	0.740
12	44.870	31.409	13.461	0.000	3.769	10.383	11.413	4.889	0.687	1.917	0.822
13	42.132	29.492	12.639	0.000	3.539	10.824	11.867	4.591	0.687	2.136	0.915
14	39.081	27.357	11.724	0.000	3.283	11.291	12.348	4.258	0.687	2.387	1.023
15	35.670	24.969	10.701	0.000	2.996	11.786	12.859	3.887	0.687	2.678	1.148
16	31.845	22.292	9.554	0.000	2.675	12.316	13.405	3.470	0.687	3.012	1.291
17	27.542	19.279	8.263	0.000	2.314	12.884	13.991	3.001	0.687	3.398	1.456
18	22.687	15.881	6.806	0.000	1.906	13.498	14.623	2.472	0.687	3.843	1.647
19	17.197	12.038	5.159	0.000	1.445	14.163	15.309	1.874	0.687	4.356	1.867
20	10.973	7.681	3.292	0.000	0.922	14.888	16.056	1.196	0.687	4.948	2.120

KEY TO CALCULATIONS IS AS FOLLOWS.

COLUMN 1 = TOTAL CAPITAL AT START OF YEAR
 COLUMN 2 = DEBT OUTSTANDING AT START OF YEAR
 COLUMN 3 = EQUITY CAPITAL AT START OF YEAR
 COLUMN 4 = AMOUNT ADDED TO DEBT
 COLUMN 5 = INTEREST ON DEBT
 COLUMN 6 = FEDERAL TAXABLE INCOME
 COLUMN 7 = STATE TAXABLE INCOME
 COLUMN 8 = AFTER-TAX RETURN ON EQUITY
 COLUMN 9 = LOCAL PROPERTY TAX PAID
 COLUMN 10 = RETIREMENT OF DEBT
 COLUMN 11 = RETIREMENT OF EQUITY

9.761
3.254

8.947
2.440

8.134
1.627

7.321
0.813

6.507

5.694

4.880

4.067

SUM OF DEPRECIATION ALLOWANCES 124.449

PAYOUT TABULATION
FEED PRICE: 0.50
PRODUCT PRICE: 0.5860
WORKING CAPITAL: 5.958
START-UP COSTS: 0.197
CASE: PHENOSOLVAN--WAO WITH NEGATIVE FEDERAL TAXES
RATE OF RETURN ON EQUITY 15.00%
DEBT AT 12.00%
D/E RATIO 0.70/0.30
TRIAL 19

1	2	3	4	5	6	7	8	9	10	11
CPSTRT	EFFNCY	INCOME	PLTEXP	STLTAX	TACRDT	FEDTAX	CASHFL	TOTRTN	AMORTZ	CAPEND
1	38.605	0.000	0.000	0.000	0.000	-1.492	1.492	4.980	-3.488	42.093
2	104.318	0.000	0.000	0.000	0.000	-4.031	4.031	13.457	-9.426	113.754
3	138.634	0.000	0.000	0.000	0.000	-5.357	5.357	17.884	-12.527	151.161
4	155.848	0.500	29.947	2.489	8.711	-17.732	25.041	20.104	4.937	150.912
5	150.912	0.800	47.916	2.489	0.000	-2.215	23.090	19.468	3.623	147.289
6	147.289	0.900	53.905	2.586	0.000	0.304	24.930	19.000	5.930	141.359
7	141.359	0.950	56.900	2.693	0.000	1.884	25.473	18.235	7.238	134.121
8	134.121	0.950	56.900	2.735	0.000	2.518	24.796	17.302	7.494	126.627
9	126.627	0.950	56.900	2.779	0.000	3.162	24.109	16.335	7.774	118.853
10	118.853	0.950	56.900	2.823	0.000	3.816	23.411	15.332	8.079	110.775
11	110.775	0.950	56.900	2.867	0.000	4.482	22.700	14.290	8.410	102.364
12	102.364	0.950	56.900	2.913	0.000	5.160	21.976	13.205	8.771	93.593
13	93.593	0.950	56.900	2.959	0.000	5.852	21.238	12.074	9.165	84.428
14	84.428	0.950	56.900	3.007	0.000	6.558	20.484	10.891	9.593	74.836
15	74.836	0.950	56.900	3.055	0.000	7.280	19.713	9.654	10.059	64.776
16	64.776	0.950	56.900	3.105	0.000	8.020	18.923	8.356	10.567	54.209
17	54.209	0.950	56.900	3.156	0.000	8.779	18.113	6.993	11.120	43.088
18	43.088	0.950	56.900	3.209	0.000	9.559	17.281	5.558	11.723	31.366
19	31.366	0.950	56.900	3.263	0.000	10.361	16.425	4.046	12.379	18.987
20	18.987	0.950	56.900	3.318	0.000	11.188	15.542	2.449	13.093	5.894

THE LAST ENTRY IN THE FINAL YEAR SHOULD BE EQUAL TO THE WORKING CAPITAL INVESTMENT PLUS SALVAGE VALUE

WORKING CAPITAL RECOVERED INTACT AT END OF PROJECT LIFE = 5.958
SALVAGE VALUE RECOVERED INTACT AT END OF PROJECT LIFE = 0.000
TOTAL 5.958

KEY TO CALCULATIONS IS AS FOLLOWS.

COLUMN 1 = CAPITAL INVESTMENT OUTSTANDING AT START OF YEAR INCLUDING WORKING CAPITAL
COLUMN 2 = ONSTREAM EFFICIENCY (PLANT OPERATING FACTOR)
COLUMN 3 = GROSS INCOME FROM SALES
COLUMN 4 = OPERATING EXPENSES EXCLUDING TAXES
COLUMN 5 = STATE AND LOCAL TAXES
COLUMN 6 = FEDERAL INVESTMENT TAX CREDIT
COLUMN 7 = NET FEDERAL INCOME TAX PAID
COLUMN 8 = CASH FLOW AFTER TAXES
COLUMN 9 = INTEREST ON DEBT AND RETURN ON EQUITY BASED ON COLUMN 1

COLUMN 10 = REDUCTION OF OUTSTANDING CAPITAL INVESTMENT, COLUMN 8-9
 COLUMN 11 = OUTSTANDING CAPITAL INVESTMENT AT END OF YEAR INCL WORKING CAPITAL. COL 1-10
 COLUMN 11 PLUS NEW INVESTMENT = NEXT ENTRY OF COLUMN 1

ADDITIONAL DETAILS OF CALCULATION PHENOSOLVAN--HAO WITH NEGATIVE FEDERAL TAXES

	1	2	3	4	5	6	7	8	9	10	11
	CPSTRT	DBSTRT	EQSTRT	DBTADD	INTRST	TAXABL	STXABL	EQYRTN	PROPTY	DBTPAY	EQTAY
1	38.605	27.023	11.581	27.023	3.243	-3.243	-3.243	1.737	0.000	-2.442	-1.046
2	104.318	73.022	31.295	43.557	8.763	-8.763	-8.763	4.694	0.000	-6.598	-2.828
3	138.634	97.044	41.590	17.423	11.645	-11.645	-11.645	6.239	0.000	-8.769	-3.758
4	155.848	109.094	46.755	3.281	13.091	-19.610	-17.121	7.013	2.489	3.456	1.481
5	150.912	105.638	45.274	0.000	12.677	-4.816	-2.327	6.791	2.489	2.536	1.087
6	147.289	103.102	44.187	0.000	12.372	0.661	3.248	6.628	2.489	4.151	1.779
7	141.359	98.951	42.408	0.000	11.874	4.095	6.787	6.361	2.489	5.066	2.171
8	134.121	93.885	40.236	0.000	11.266	5.473	8.209	6.035	2.489	5.246	2.248
9	126.627	88.639	37.988	0.000	10.637	6.873	9.552	5.698	2.489	5.442	2.332
10	118.853	83.197	35.656	0.000	9.984	8.296	11.118	5.348	2.489	5.655	2.424
11	110.775	77.542	33.232	0.000	9.305	9.743	12.610	4.985	2.489	5.887	2.523
12	102.364	71.655	30.709	0.000	8.599	11.217	14.130	4.606	2.489	6.140	2.631
13	93.593	65.515	28.078	0.000	7.862	12.721	15.680	4.212	2.489	6.415	2.749
14	84.428	59.100	25.329	0.000	7.092	14.256	17.263	3.799	2.489	6.715	2.878
15	74.836	52.385	22.451	0.000	6.286	15.827	18.883	3.368	2.489	7.042	3.018
16	64.776	45.343	19.433	0.000	5.441	17.436	20.541	2.915	2.489	7.397	3.170
17	54.209	37.946	16.263	0.000	4.554	19.086	22.242	2.439	2.489	7.784	3.336
18	43.088	30.162	12.927	0.000	3.619	20.781	23.989	1.939	2.489	8.206	3.517
19	31.366	21.956	9.410	0.000	2.635	22.525	25.788	1.411	2.489	8.665	3.714
20	18.987	13.291	5.696	0.000	1.595	24.323	27.641	0.854	2.489	9.165	3.928

KEY TO CALCULATIONS IS AS FOLLOWS.

COLUMN 1 = TOTAL CAPITAL AT START OF YEAR
 COLUMN 2 = DEBT OUTSTANDING AT START OF YEAR
 COLUMN 3 = EQUITY CAPITAL AT START OF YEAR
 COLUMN 4 = AMOUNT ADDED TO DEBT
 COLUMN 5 = INTEREST ON DEBT
 COLUMN 6 = FEDERAL TAXABLE INCOME
 COLUMN 7 = STATE TAXABLE INCOME
 COLUMN 8 = AFTER-TAX RETURN ON EQUITY
 COLUMN 9 = LOCAL PROPERTY TAX PAID
 COLUMN 10 = RETIREMENT OF DEBT
 COLUMN 11 = RETIREMENT OF EQUITY

DATE: 11/ 4/84

[illegible]

OBJECTIVE IS TO CALCULATE PRODUCT PRICE

LEVEL OF INVESTMENT: 100.0%

CONSTANT OPERATING EXPENSES AT 100% LOAD-FACTOR: 12,288

TOTAL DEPRECIABLE INVESTMENT	124.449
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ANNUAL DEPRECIATION ALLOWANCES		BY SYD METHOD
1	100	100
2	100	100
3	100	100
4	100	100
5	100	100
6	100	100
7	100	100
8	100	100
9	100	100
10	100	100
11	100	100
12	100	100
13	100	100
14	100	100
15	100	100
16	100	100
17	100	100
18	100	100
19	100	100
20	100	100
21	100	100
22	100	100
23	100	100
24	100	100
25	100	100
26	100	100
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30	100	100
31	100	100
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42	100	100
43	100	100
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85	100	100
86	100	100
87	100	100
88	100	100
89	100	100
90	100	100
91	100	100
92	100	100
93	100	100
94	100	100
95	100	100
96	100	100
97	100	100
98	100	100
99	100	100
100	100	100

0,000

13-828

13.014

12.201

11.388

10.574

9.761 8.947 8.134 7.321 6.507 5.694 4.880 4.067
3.254 2.440 1.627 0.813

SUM OF DEPRECIATION ALLOWANCES 124.449

PAYOUT TABULATION
FEED PRICE: 0.50 CASE: PHENOSOLVAN--HAO W/O NEGATIVE FEDERAL TAXES
PRODUCT PRICE: 0.6409 RATE OF RETURN ON EQUITY 15.00%
WORKING CAPITAL: 5.958 DEBT AT 12.00%
START-UP COSTS: 0.197 D/E RATIO 0.70/0.30
TRIAL 18

	1	2	3	4	5	6	7	8	9	10	11
CPSTRT	EFFNCY	INCOME	PLTEXP	STLTAX	TXCRDT	FEDTAX	CASHFL	TOTRTN	AMORTZ	CAPEND	
1	38.605	0.000	0.000	0.000	0.000	0.000	0.000	4.980	-4.980	43.585	
2	105.809	0.000	0.000	0.000	0.000	0.000	0.000	13.649	-13.649	119.459	
3	144.349	0.000	0.000	0.000	0.000	0.000	0.000	18.621	-18.621	162.970	
4	167.657	0.500	20.149	2.489	0.000	0.000	10.112	21.628	-11.516	179.173	
5	179.173	0.800	24.551	2.489	0.000	0.000	25.360	23.113	2.246	176.927	
6	176.927	0.900	26.084	2.663	1.444	0.000	30.203	22.872	7.379	169.548	
7	169.548	0.950	26.851	2.781	3.203	0.000	32.593	21.872	10.721	158.826	
8	158.826	0.950	26.851	2.833	3.968	0.000	32.542	20.489	12.053	146.773	
9	146.773	0.950	26.851	2.888	0.096	4.687	27.800	18.934	8.866	137.907	
10	137.907	0.950	26.851	2.934	0.000	5.478	26.962	17.790	9.172	128.735	
11	128.735	0.950	26.851	2.982	0.000	6.185	26.208	16.607	9.601	119.134	
12	119.134	0.950	26.851	3.030	0.000	6.907	25.437	15.368	10.068	109.066	
13	109.066	0.950	26.851	3.080	0.000	7.648	24.646	14.069	10.577	98.489	
14	98.489	0.950	26.851	3.131	0.000	8.407	23.836	12.705	11.131	87.358	
15	87.358	0.950	26.851	3.184	0.000	9.187	23.003	11.269	11.734	75.623	
16	75.623	0.950	26.851	3.238	0.000	9.990	22.147	9.755	12.391	63.232	
17	63.232	0.950	26.851	3.293	0.000	10.817	21.264	8.157	13.107	50.125	
18	50.125	0.950	26.851	3.351	0.000	11.672	20.352	6.466	13.886	36.239	
19	36.239	0.950	26.851	3.410	0.000	12.555	19.409	4.675	14.734	21.505	
20	21.505	0.950	26.851	3.472	0.000	13.470	18.433	2.774	15.658	5.846	

THE LAST ENTRY IN THE FINAL YEAR SHOULD BE EQUAL TO THE WORKING CAPITAL INVESTMENT PLUS SALVAGE VALUE

WORKING CAPITAL RECOVERED INTACT AT END OF PROJECT LIFE = 5.958
SALVAGE VALUE RECOVERED INTACT AT END OF PROJECT LIFE = 0.000
TOTAL 5.958

KEY TO CALCULATIONS IS AS FOLLOWS.

COLUMN 1 = CAPITAL INVESTMENT OUTSTANDING AT START OF YEAR INCLUDING WORKING CAPITAL
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COLUMN 10 = REDUCTION OF OUTSTANDING CAPITAL INVESTMENT, COLUMN 8-9
 COLUMN 11 = OUTSTANDING CAPITAL INVESTMENT AT END OF YEAR INCL WORKING CAPITAL. COL 1-10
 COLUMN 11 PLUS NEW INVESTMENT = NEXT ENTRY OF COLUMN 1

ADDITIONAL DETAILS OF CALCULATION PHENOSOLVAN--WAO W/O NEGATIVE FEDERAL TAXES

	1	2	3	4	5	6	7	8	9	10	11
	CPSTRT	DBSTRT	EQSTRT	DBTADD	INTRST	TAXABL	STXABL	EQVRTN	PROPTY	DBTPAY	EQTPAY
1	38.605	27.023	11.581	27.023	3.243	-3.243	-3.243	1.737	0.000	-3.486	-1.494
2	105.809	74.067	31.743	43.557	8.888	-8.888	-8.888	4.761	0.000	-9.555	-4.095
3	144.349	101.044	43.305	17.423	12.125	-12.125	-12.125	6.496	0.000	-13.035	-5.586
4	167.657	117.360	50.297	3.281	14.083	-17.799	-15.310	7.545	2.489	-8.061	-3.455
5	179.173	125.421	53.752	0.000	15.051	-2.705	-0.216	8.063	2.489	1.572	0.674
6	176.927	123.849	53.078	0.000	14.862	3.140	5.803	7.630	2.489	5.165	2.214
7	169.548	118.683	50.864	0.000	14.242	6.964	9.745	7.147	2.489	7.505	3.216
8	158.826	111.178	47.648	0.000	13.341	8.626	11.459	7.147	2.489	8.437	3.616
9	146.773	102.741	44.032	0.000	12.329	10.397	13.285	6.605	2.489	6.206	2.660
10	137.907	96.535	41.372	0.000	11.584	11.909	14.843	6.206	2.489	6.420	2.752
11	128.735	90.114	38.621	0.000	10.814	13.445	16.427	5.793	2.489	6.721	2.880
12	119.134	83.394	35.740	0.000	10.007	15.016	18.047	5.361	2.489	7.048	3.020
13	109.066	76.346	32.720	0.000	9.162	16.626	19.706	4.908	2.489	7.404	3.173
14	98.489	68.942	29.547	0.000	8.273	18.276	21.408	4.432	2.489	7.792	3.339
15	87.358	61.150	26.207	0.000	7.338	19.972	23.156	3.931	2.489	8.214	3.520
16	75.623	52.936	22.687	0.000	6.352	21.717	24.955	3.403	2.489	8.674	3.717
17	63.232	44.262	18.970	0.000	5.311	23.516	26.809	2.845	2.489	9.175	3.932
18	50.125	35.088	15.038	0.000	4.211	25.373	28.724	2.256	2.489	9.720	4.166
19	36.239	25.367	10.872	0.000	3.044	27.293	30.703	1.631	2.489	10.314	4.420
20	21.505	15.053	6.452	0.000	1.806	29.283	32.755	0.968	2.489	10.961	4.698

KEY TO CALCULATIONS IS AS FOLLOWS.

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 COLUMN 4 = AMOUNT ADDED TO DEBT
 COLUMN 5 = INTEREST ON DEBT
 COLUMN 6 = FEDERAL TAXABLE INCOME
 COLUMN 7 = STATE TAXABLE INCOME
 COLUMN 8 = AFTER-TAX RETURN ON EQUITY
 COLUMN 9 = LOCAL PROPERTY TAX PAID
 COLUMN 10 = RETIREMENT OF DEBT
 COLUMN 11 = RETIREMENT OF EQUITY

9.761 8.947 8.134 7.321 5.694 4.880 4.067
 3.254 2.440 1.627 0.813
 SUM OF DEPRECIATION ALLOWANCES 124.449

PAYOUT TABULATION
 FEED PRICE: 0.50 CASE: PHENOSOLVAN--WAO WITH NEGATIVE FEDERAL TAXES
 PRODUCT PRICE: 17.5823 RATE OF RETURN ON EQUITY 15.00%
 WORKING CAPITAL: 7.555 DEBT AT 12.00%
 START-UP COSTS: 0.197 D/E RATIO 0.70/0.30
 TRIAL 23

	1	2	3	4	5	6	7	8	9	10	11
	CPSTRT	EFFNCY	INCOME	PLTEXP	STLTAX	TXCRDT	FEDTAX	CASHFL	TOTRTN	AMORTZ	CAPEND
1	38.605	0.000	0.000	0.000	0.000	0.000	-1.492	1.492	4.980	-3.488	42.093
2	104.318	0.000	0.000	0.000	0.000	0.000	-4.031	4.031	13.457	-9.426	113.744
3	138.634	0.000	0.000	0.000	0.000	0.000	-5.357	5.357	17.884	-12.527	151.161
4	157.445	0.500	30.099	20.149	2.489	8.711	-17.724	25.185	20.310	4.874	152.571
5	152.571	0.800	48.158	24.551	2.489	0.000	-2.168	23.285	19.682	3.604	148.967
6	148.967	0.900	54.178	26.084	2.590	0.000	0.363	25.140	19.217	5.923	143.044
7	143.044	0.950	57.188	26.851	2.697	0.000	1.949	25.691	18.453	7.238	135.806
8	135.806	0.950	57.188	26.851	2.740	0.000	3.227	25.014	17.519	7.495	128.310
9	128.310	0.950	57.188	26.851	2.783	0.000	3.227	24.327	16.552	7.775	120.535
10	120.535	0.950	57.188	26.851	2.827	0.000	3.881	23.629	15.549	8.080	112.456
11	112.456	0.950	57.188	26.851	2.872	0.000	4.547	22.918	14.507	8.411	104.044
12	104.044	0.950	57.188	26.851	2.917	0.000	5.225	22.194	13.422	8.773	95.271
13	95.271	0.950	57.188	26.851	2.964	0.000	5.917	21.456	12.290	9.166	86.105
14	86.105	0.950	57.188	26.851	3.011	0.000	6.624	20.702	11.108	9.595	76.511
15	76.511	0.950	57.188	26.851	3.060	0.000	7.346	19.931	9.870	10.061	66.450
16	66.450	0.950	57.188	26.851	3.110	0.000	8.086	19.141	8.572	10.569	55.881
17	55.881	0.950	57.188	26.851	3.161	0.000	8.845	18.331	7.209	11.123	44.758
18	44.758	0.950	57.188	26.851	3.213	0.000	9.625	17.499	5.774	11.725	33.033
19	33.033	0.950	57.188	26.851	3.267	0.000	10.427	16.643	4.261	12.381	20.652
20	20.652	0.950	57.188	26.851	3.323	0.000	11.254	15.760	2.664	13.096	7.556

THE LAST ENTRY IN THE FINAL YEAR SHOULD BE EQUAL TO THE WORKING CAPITAL INVESTMENT PLUS SALVAGE VALUE

WORKING CAPITAL RECOVERED INTACT AT END OF PROJECT LIFE = 7.555
 SALVAGE VALUE RECOVERED INTACT AT END OF PROJECT LIFE = 0.000
 TOTAL 7.555

KEY TO CALCULATIONS IS AS FOLLOWS.

- COLUMN 1 = CAPITAL INVESTMENT OUTSTANDING AT START OF YEAR INCLUDING WORKING CAPITAL
- COLUMN 2 = ONSTREAM EFFICIENCY (PLANT OPERATING FACTOR)
- COLUMN 3 = GROSS INCOME FROM SALES
- COLUMN 4 = OPERATING EXPENSES EXCLUDING TAXES
- COLUMN 5 = STATE AND LOCAL TAXES
- COLUMN 6 = FEDERAL INVESTMENT TAX CREDIT
- COLUMN 7 = NET FEDERAL INCOME TAX PAID
- COLUMN 8 = CASH FLOW AFTER TAXES
- COLUMN 9 = INTEREST ON DEBT AND RETURN ON EQUITY BASED ON COLUMN 1

COLUMN 10 = REDUCTION OF OUTSTANDING CAPITAL INVESTMENT, COLUMN 8-9
 COLUMN 11 = OUTSTANDING CAPITAL INVESTMENT AT END OF YEAR INCL WORKING CAPITAL. COL 1-10
 COLUMN 11 PLUS NEW INVESTMENT = NEXT ENTRY OF COLUMN 1

ADDITIONAL DETAILS OF CALCULATION PHENOSOLVAN--WAO WITH NEGATIVE FEDERAL TAXES

	1	2	3	4	5	6	7	8	9	10	11
	CPSTRT	DBSTRT	EQSTRT	DBTADD	INTRST	TAXABL	STXABL	EQYRTM	PROPTY	DBTPAY	EQTPAY
1	38.605	27.023	11.581	27.023	3.243	-3.243	-3.243	1.737	0.000	-2.442	-1.046
2	104.318	73.022	31.295	43.557	8.763	-8.763	-8.763	4.694	0.000	-6.598	-2.828
3	138.634	97.044	41.590	17.423	11.645	-11.645	-11.645	6.239	0.000	-8.769	-3.758
4	157.445	110.212	47.234	4.399	13.225	-19.593	-17.104	7.085	2.489	3.412	1.462
5	152.571	106.800	45.771	0.000	12.816	-4.713	-2.224	6.866	2.489	2.523	1.081
6	148.967	104.277	44.690	0.000	12.513	0.789	3.379	6.704	2.489	4.146	1.777
7	143.044	100.131	42.913	0.000	12.016	4.237	6.934	6.437	2.489	5.067	2.172
8	135.806	95.064	40.742	0.000	11.408	5.616	8.355	6.111	2.489	5.247	2.249
9	128.310	89.817	38.493	0.000	10.778	7.015	9.798	5.774	2.489	5.443	2.333
10	120.535	84.375	36.161	0.000	10.125	8.438	11.265	5.424	2.489	5.656	2.424
11	112.456	78.719	33.737	0.000	9.446	9.885	12.757	5.061	2.489	5.888	2.523
12	104.044	72.831	31.213	0.000	8.740	11.359	14.277	4.682	2.489	6.141	2.632
13	95.271	66.690	28.581	0.000	8.003	12.863	15.827	4.287	2.489	6.416	2.750
14	86.105	60.274	25.832	0.000	7.233	14.399	17.410	3.875	2.489	6.716	2.878
15	76.511	53.558	22.953	0.000	6.427	15.970	19.030	3.443	2.489	7.043	3.018
16	66.450	46.515	19.935	0.000	5.582	17.579	20.688	2.990	2.489	7.398	3.171
17	55.881	39.116	16.764	0.000	4.694	19.229	22.389	2.515	2.489	7.786	3.337
18	44.758	31.331	13.427	0.000	3.760	20.924	24.137	2.014	2.489	8.208	3.518
19	33.033	23.123	9.910	0.000	2.775	22.668	25.935	1.486	2.489	8.667	3.714
20	20.652	14.456	6.196	0.000	1.735	24.466	27.789	0.929	2.489	9.167	3.929

KEY TO CALCULATIONS IS AS FOLLOWS.

COLUMN 1 = TOTAL CAPITAL AT START OF YEAR
 COLUMN 2 = DEBT OUTSTANDING AT START OF YEAR
 COLUMN 3 = EQUITY CAPITAL AT START OF YEAR
 COLUMN 4 = AMOUNT ADDED TO DEBT
 COLUMN 5 = INTEREST ON DEBT
 COLUMN 6 = FEDERAL TAXABLE INCOME
 COLUMN 7 = STATE TAXABLE INCOME
 COLUMN 8 = AFTER-TAX RETURN ON EQUITY
 COLUMN 9 = LOCAL PROPERTY TAX PAID
 COLUMN 10 = RETIREMENT OF DEBT
 COLUMN 11 = RETIREMENT OF EQUITY

9.761 8.947 8.134 7.321 6.507 5.694 4.880 4.067
 3.254 2.440 1.627 0.813
 SUM OF DEPRECIATION ALLOWANCES 124.449

PAYOUT TABULATION
 FEED PRICE: 0.50
 PRODUCT PRICE: 20.7983
 WORKING CAPITAL: 7.555
 START-UP COSTS: 0.197
 CASE: PHENOSOLVAN--WAO W/O NEGATIVE FEDERAL TAXES
 RATE OF RETURN ON EQUITY 15.00%
 DEBT AT 12.00%
 D/E RATIO 0.70/0.30
 TRIAL 23

	1	2	3	4	5	6	7	8	9	10	11
	CPSTRT	EFFNCY	INCOME	PLTEXP	STLTAX	TXCRDT	FEDTAX	CASHFL	TOTRTN	AMORTZ	CAPEXND
1	38.605	0.000	0.000	0.000	0.000	0.000	0.000	0.000	4.980	-4.980	43.585
2	105.809	0.000	0.000	0.000	0.000	0.000	0.000	0.000	13.649	-13.649	119.459
3	144.349	0.000	0.000	0.000	0.000	0.000	0.000	0.000	18.621	-18.621	162.970
4	169.254	0.500	32.894	20.149	2.489	0.000	0.000	10.255	21.834	-11.579	180.833
5	180.833	0.800	52.630	24.551	2.489	0.000	0.000	25.589	23.327	2.262	178.571
6	178.571	0.900	59.208	26.084	2.667	1.498	0.000	30.457	23.036	7.422	171.149
7	171.149	0.950	62.498	26.851	2.785	3.265	0.000	32.727	22.078	10.783	160.366
8	160.366	0.950	62.498	26.851	2.837	3.949	0.000	32.727	20.687	12.039	148.327
9	148.327	0.950	62.498	26.851	2.892	0.000	0.083	27.909	19.134	8.775	139.552
10	139.552	0.950	62.498	26.851	2.938	0.000	5.538	27.171	18.002	9.169	130.383
11	130.383	0.950	62.498	26.851	2.986	0.000	6.245	26.417	16.819	9.597	120.786
12	120.786	0.950	62.498	26.851	3.034	0.000	6.967	25.645	15.581	10.064	110.722
13	110.722	0.950	62.498	26.851	3.084	0.000	7.707	24.855	14.283	10.572	100.149
14	100.149	0.950	62.498	26.851	3.135	0.000	8.467	24.045	12.919	11.126	89.023
15	89.023	0.950	62.498	26.851	3.188	0.000	9.246	23.213	11.484	11.729	77.294
16	77.294	0.950	62.498	26.851	3.242	0.000	10.049	22.356	9.971	12.385	64.909
17	64.909	0.950	62.498	26.851	3.297	0.000	10.876	21.474	8.373	13.100	51.809
18	51.809	0.950	62.498	26.851	3.355	0.000	11.730	20.562	6.683	13.879	37.930
19	37.930	0.950	62.498	26.851	3.414	0.000	12.613	19.620	4.893	14.727	23.203
20	23.203	0.950	62.498	26.851	3.476	0.000	13.528	18.643	2.993	15.650	7.553

THE LAST ENTRY IN THE FINAL YEAR SHOULD BE EQUAL TO THE WORKING CAPITAL INVESTMENT PLUS SALVAGE VALUE

WORKING CAPITAL RECOVERED INTACT AT END OF PROJECT LIFE = 7.555
 SALVAGE VALUE RECOVERED INTACT AT END OF PROJECT LIFE = 0.000
 TOTAL 7.555

KEY TO CALCULATIONS IS AS FOLLOWS.

COLUMN 1 = CAPITAL INVESTMENT OUTSTANDING AT START OF YEAR INCLUDING WORKING CAPITAL
 COLUMN 2 = ONSREAM EFFICIENCY (PLANT OPERATING FACTOR)
 COLUMN 3 = GROSS INCOME FROM SALES
 COLUMN 4 = OPERATING EXPENSES EXCLUDING TAXES
 COLUMN 5 = STATE AND LOCAL TAXES
 COLUMN 6 = FEDERAL INVESTMENT TAX CREDIT
 COLUMN 7 = NET FEDERAL INCOME TAX PAID
 COLUMN 8 = CASH FLOW AFTER TAXES
 COLUMN 9 = INTEREST ON DEBT AND RETURN ON EQUITY BASED ON COLUMN 1

COLUMN 10 = REDUCTION OF OUTSTANDING CAPITAL INVESTMENT, COLUMN 8-9
 COLUMN 11 = OUTSTANDING CAPITAL INVESTMENT AT END OF YEAR INCL WORKING CAPITAL, COL 1-10
 COLUMN 11 PLUS NEW INVESTMENT = NEXT ENTRY OF COLUMN 1

ADDITIONAL DETAILS OF CALCULATION PHENOSOLVAN--HAO W/O NEGATIVE FEDERAL TAXES

	1	2	3	4	5	6	7	8	9	10	11
CPSTRT	DBSTRT	EQSTRT	DBTADD	INTRST	TAXABL	STXABL	EQYRTN	PROPTY	DBIPAY	EQIPAY	
1 38.605	27.023	11.581	27.023	3.243	-3.243	-3.243	1.737	0.000	-3.486	-1.494	
2 105.809	74.067	31.743	43.557	8.888	-8.888	-8.888	4.761	0.000	-9.555	-4.095	
3 144.349	101.044	43.305	17.423	12.125	-12.125	-12.125	6.496	0.000	-13.035	-5.586	
4 169.254	118.478	50.776	4.399	14.217	-17.790	-15.301	7.616	2.489	-8.105	-3.474	
5 180.833	126.583	54.250	0.000	15.190	-2.615	-0.126	8.137	2.489	1.583	0.679	
6 178.571	125.000	53.571	0.000	15.000	3.257	5.923	8.036	2.489	5.195	2.227	
7 171.149	119.804	51.345	0.000	14.377	7.097	9.883	7.702	2.489	7.548	3.235	
8 160.366	112.256	48.110	0.000	13.471	8.765	11.602	7.216	2.489	8.428	3.612	
9 148.327	103.829	44.498	0.000	12.459	10.535	13.427	6.675	2.489	6.142	2.751	
10 139.552	97.686	41.866	0.000	11.722	12.039	14.977	6.280	2.489	6.418	2.879	
11 130.383	91.268	39.115	0.000	10.952	13.575	16.561	5.867	2.489	6.718	3.019	
12 120.786	84.550	36.236	0.000	10.146	15.146	18.180	5.435	2.489	7.045	3.172	
13 110.722	77.505	33.217	0.000	9.301	16.755	19.839	4.982	2.489	7.401	3.338	
14 100.149	70.104	30.045	0.000	8.413	18.405	21.541	4.507	2.489	7.788	3.519	
15 89.023	62.316	26.707	0.000	7.478	20.101	23.289	4.006	2.489	8.210	3.716	
16 77.294	54.106	23.188	0.000	6.493	21.846	25.087	3.478	2.489	8.670	3.930	
17 64.909	45.436	19.473	0.000	5.452	23.644	26.941	2.921	2.489	9.170	4.164	
18 51.809	36.266	15.543	0.000	4.352	25.500	28.855	2.331	2.489	9.715	4.418	
19 37.930	26.551	11.379	0.000	3.186	27.420	30.834	1.707	2.489	10.309	4.695	
20 23.203	16.242	6.961	0.000	1.949	29.409	32.885	1.044	2.489	10.955		

KEY TO CALCULATIONS IS AS FOLLOWS.

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 COLUMN 7 = STATE TAXABLE INCOME
 COLUMN 8 = AFTER-TAX RETURN ON EQUITY
 COLUMN 9 = LOCAL PROPERTY TAX PAID
 COLUMN 10 = RETIREMENT OF DEBT
 COLUMN 11 = RETIREMENT OF EQUITY

210.226 192.707 175.188 157.669 140.151 122.632 105.113 87.594
70.075 52.556 35.038 17.519

SUM OF DEPRECIATION ALLOWANCES 2680.377

PAYOUT TABULATION
FEED PRICE: 45.00 CASE: INDEPENDENT COAL GASIFICATION COST ANALYSIS
PRODUCT PRICE: 8.2362 RATE OF RETURN ON EQUITY 15.00%
WORKING CAPITAL: 98.225 DEBT AT 12.00%
START-UP COSTS: 0.694 D/E RATIO 0.70/0.30
TRIAL 22

	1	2	3	4	5	6	7	8	9	10	11
	CPSTRT	EFFNCY	INCOME	PLTEXP	SILTAX	TXCRDT	FEDTAX	CASHFL	TOTRTN	AMORTZ	CAPEND
1	812.424	0.000	0.000	0.000	0.000	0.000	0.000	0.000	104.803	-104.803	917.226
2	2257.416	0.000	0.000	0.000	0.000	0.000	0.000	0.000	291.207	-291.207	2548.622
3	3084.698	0.000	0.000	0.000	0.000	0.000	0.000	0.000	397.926	-397.926	3482.624
4	3572.539	0.500	534.943	219.390	53.608	0.000	0.000	261.946	460.857	-198.911	3771.450
5	3771.450	0.800	855.909	256.969	53.663	0.000	0.000	545.277	486.517	58.760	3712.690
6	3712.690	0.900	962.898	269.727	57.163	28.225	0.000	636.007	478.937	157.070	3555.620
7	3555.620	0.950	1016.392	276.106	59.498	62.952	0.000	680.787	458.675	222.113	3333.507
8	3333.507	0.950	1016.392	276.106	60.583	79.094	0.000	679.702	430.022	249.680	3083.827
9	3083.827	0.950	1016.392	276.106	61.738	17.355	78.914	599.634	397.813	201.820	2882.007
10	2882.007	0.950	1016.392	276.106	62.772	0.000	111.650	565.863	371.779	194.084	2687.923
11	2687.923	0.950	1016.392	276.106	63.787	0.000	126.741	549.757	346.742	203.015	2484.908
12	2484.908	0.950	1016.392	276.106	64.824	0.000	142.167	533.294	320.553	212.741	2272.167
13	2272.167	0.950	1016.392	276.106	65.886	0.000	157.958	516.441	293.109	223.332	2048.835
14	2048.835	0.950	1016.392	276.106	66.974	0.000	174.146	499.166	264.300	234.866	1813.969
15	1813.969	0.950	1016.392	276.106	68.092	0.000	190.766	481.428	234.002	247.426	1566.542
16	1566.542	0.950	1016.392	276.106	69.241	0.000	207.856	463.189	202.084	261.105	1305.438
17	1305.438	0.950	1016.392	276.106	70.424	0.000	225.460	444.402	168.401	276.000	1029.438
18	1029.438	0.950	1016.392	276.106	71.645	0.000	243.621	425.019	132.797	292.222	737.216
19	737.216	0.950	1016.392	276.106	72.907	0.000	262.391	404.988	95.101	309.887	427.329
20	427.329	0.950	1016.392	276.106	74.214	0.000	281.822	384.250	55.126	329.124	98.205

THE LAST ENTRY IN THE FINAL YEAR SHOULD BE EQUAL TO THE WORKING CAPITAL INVESTMENT PLUS SALVAGE VALUE

WORKING CAPITAL RECOVERED INTACT AT END OF PROJECT LIFE = 98.225
SALVAGE VALUE RECOVERED INTACT AT END OF PROJECT LIFE = 0.000
TOTAL 98.225

KEY TO CALCULATIONS IS AS FOLLOWS.

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COLUMN 9 = INTEREST ON DEBT AND RETURN ON EQUITY BASED ON COLUMN 1

COLUMN 10 = REDUCTION OF OUTSTANDING CAPITAL INVESTMENT, COLUMN 8-9
 COLUMN 11 = OUTSTANDING CAPITAL INVESTMENT AT END OF YEAR INCL WORKING CAPITAL, COL 1-10
 COLUMN 11 PLUS NEW INVESTMENT = NEXT ENTRY OF COLUMN 1

ADDITIONAL DETAILS OF CALCULATION INDEPENDENT COAL GASIFICATION COST ANALYSIS

	1	2	3	4	5	6	7	8	9	10	11
	CPSTRT	DBSTRT	EQSTRT	DBTADD	INTRST	TAXABL	STXABL	EQYRTN	PROPTY	DBTPAY	EQTPAY
1	812.424	568.696	243.727	568.696	68.244	-68.244	-68.244	36.559	0.000	-73.362	-31.401
2	2257.416	1580.191	677.225	938.133	189.623	-189.623	-189.623	101.584	0.000	-203.845	-87.362
3	3084.698	2159.288	925.410	375.253	259.115	-259.115	-259.115	138.811	0.000	-278.548	-119.378
4	3572.539	2500.776	1071.762	62.941	300.093	-335.967	-282.359	160.764	53.608	-139.238	-59.673
5	3771.450	2640.014	1131.436	0.000	316.802	-51.826	1.837	169.715	53.608	41.132	17.628
6	3712.690	2598.882	1113.808	0.000	311.866	61.359	118.522	167.071	53.608	109.949	47.121
7	3555.620	2488.933	1066.687	0.000	298.672	136.852	196.350	160.003	53.608	155.479	66.634
8	3333.507	2333.454	1000.053	0.000	280.014	171.943	232.527	150.008	53.608	174.776	74.904
9	3083.827	2158.678	925.149	0.000	259.041	209.280	271.019	138.772	53.608	141.274	60.546
10	2882.007	2017.404	864.604	0.000	242.088	242.718	305.490	129.691	53.608	135.859	58.225
11	2687.923	1881.544	806.378	0.000	225.785	275.525	339.312	120.957	53.608	142.111	60.905
12	2484.908	1739.434	745.474	0.000	208.732	309.060	373.884	111.821	53.608	148.919	63.822
13	2272.167	1590.515	681.652	0.000	190.862	343.387	409.273	102.248	53.608	156.332	67.000
14	2048.835	1434.183	614.652	0.000	172.102	378.278	445.552	92.198	53.608	164.406	70.460
15	1813.969	1269.776	544.953	0.000	152.373	414.708	482.799	81.629	53.608	173.199	74.228
16	1566.542	1096.577	469.965	0.000	131.589	451.861	521.102	70.495	53.608	182.773	78.331
17	1305.438	913.804	391.634	0.000	109.656	490.129	560.554	58.745	53.608	193.200	82.800
18	1029.438	720.604	308.834	0.000	86.472	529.611	601.256	46.325	53.608	204.555	87.667
19	737.216	516.049	221.167	0.000	61.926	570.415	643.322	33.175	53.608	216.921	92.966
20	427.329	299.128	128.201	0.000	35.895	612.657	686.871	19.230	53.608	230.387	98.737

KEY TO CALCULATIONS IS AS FOLLOWS.

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 COLUMN 4 = AMOUNT ADDED TO DEBT
 COLUMN 5 = INTEREST ON DEBT
 COLUMN 6 = FEDERAL TAXABLE INCOME
 COLUMN 7 = STATE TAXABLE INCOME
 COLUMN 8 = AFTER-TAX RETURN ON EQUITY
 COLUMN 9 = LOCAL PROPERTY TAX PAID
 COLUMN 10 = RETIREMENT OF DEBT
 COLUMN 11 = RETIREMENT OF EQUITY

219.987 201.654 183.322 164.990 146.658 128.326 109.993 91.661
 73.329 54.997 36.664 18.332

SUM OF DEPRECIATION ALLOWANCES 2804.827

PAYOUT TABULATION CASE: INDEPENDENT COAL GASIFICATION COST ANALYSIS
 FEED PRICE: 45.00 RATE OF RETURN ON EQUITY 15.00%
 PRODUCT PRICE: 8.5192 DEBT AT 12.00%
 WORKING CAPITAL: 104.178 D/E RATIO 0.70/0.30
 START-UP COSTS: 0.836 TRIAL 22

	1	2	3	4	5	6	7	8	9	10	11
	CPSTRT	EFFNCY	INCOME	PLIEXP	STLTAX	TXCRDT	FEDTAX	CASHFL	TOTRIN	AMORTZ	CAPEND
1	851.024	0.000	0.000	0.000	0.000	0.000	0.000	0.000	109.782	-109.782	960.806
2	2363.220	0.000	0.000	0.000	0.000	0.000	0.000	0.000	304.855	-304.855	2668.076
3	3229.042	0.000	0.000	0.000	0.000	0.000	0.000	0.000	416.546	-416.546	3645.588
4	3740.190	0.500	567.587	239.144	56.097	0.000	0.000	272.346	482.484	-210.138	3950.328
5	3950.328	0.800	908.140	281.335	56.146	0.000	0.000	570.658	509.592	61.066	3889.262
6	3889.262	0.900	1021.657	295.677	59.826	29.658	0.000	666.155	501.715	164.440	3724.822
7	3724.822	0.950	1078.416	302.848	62.278	66.127	0.000	713.291	480.502	232.789	3492.033
8	3492.033	0.950	1078.416	302.848	63.414	83.032	0.000	712.154	450.472	261.682	3230.351
9	3230.351	0.950	1078.416	302.848	64.624	17.521	83.999	627.445	416.715	270.730	3019.622
10	3019.622	0.950	1078.416	302.848	65.705	0.000	117.098	592.765	389.531	203.234	2816.387
11	2816.387	0.950	1078.416	302.848	66.767	0.000	132.895	575.906	363.314	212.592	2603.795
12	2603.795	0.950	1078.416	302.848	67.852	0.000	149.043	558.672	335.889	222.783	2381.012
13	2381.012	0.950	1078.416	302.848	68.964	0.000	165.573	541.031	307.150	233.881	2147.132
14	2147.132	0.950	1078.416	302.848	70.103	0.000	182.519	522.946	276.980	245.966	1911.166
15	1911.166	0.950	1078.416	302.848	71.273	0.000	199.918	504.377	245.250	259.127	1642.039
16	1642.039	0.950	1078.416	302.848	72.476	0.000	217.810	485.282	211.823	273.459	1368.580
17	1368.580	0.950	1078.416	302.848	73.715	0.000	236.239	465.614	176.547	289.067	1079.514
18	1079.514	0.950	1078.416	302.848	74.993	0.000	255.253	445.321	139.257	306.064	773.450
19	773.450	0.950	1078.416	302.848	76.315	0.000	274.905	424.349	99.775	324.573	448.876
20	448.876	0.950	1078.416	302.848	77.682	0.000	295.250	402.635	57.905	344.730	104.146

THE LAST ENTRY IN THE FINAL YEAR SHOULD BE EQUAL TO THE WORKING CAPITAL INVESTMENT PLUS SALVAGE VALUE

WORKING CAPITAL RECOVERED INTACT AT END OF PROJECT LIFE = 104.178
 SALVAGE VALUE RECOVERED INTACT AT END OF PROJECT LIFE = 0.000
 TOTAL 104.178

KEY TO CALCULATIONS IS AS FOLLOWS.

- COLUMN 1 = CAPITAL INVESTMENT OUTSTANDING AT START OF YEAR INCLUDING WORKING CAPITAL
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- COLUMN 3 = GROSS INCOME FROM SALES
- COLUMN 4 = OPERATING EXPENSES EXCLUDING TAXES
- COLUMN 5 = STATE AND LOCAL TAXES
- COLUMN 6 = FEDERAL INVESTMENT TAX CREDIT
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- COLUMN 8 = CASH FLOW AFTER TAXES
- COLUMN 9 = INTEREST ON DEBT AND RETURN ON EQUITY BASED ON COLUMN 1

COLUMN 10 = REDUCTION OF OUTSTANDING CAPITAL INVESTMENT, COLUMN 8-9
 COLUMN 11 = OUTSTANDING CAPITAL INVESTMENT AT END OF YEAR INCL WORKING CAPITAL, COL 1-10
 COLUMN 11 PLUS NEW INVESTMENT = NEXT ENTRY OF COLUMN 1

ADDITIONAL DETAILS OF CALCULATION INDEPENDENT COAL GASIFICATION COST ANALYSIS

	1	2	3	4	5	6	7	8	9	10	11
	CPSIRT	DBSIRT	EQSTRT	DBTADD	INIRST	TAXABL	STXABL	EQVRTN	PROPTY	DBTPAY	EQIPAY
1	851.024	595.717	255.307	595.717	71.486	-71.486	-71.486	38.296	0.000	-76.847	-32.935
2	2363.220	1654.254	708.967	981.690	198.510	-198.510	-198.510	106.345	0.000	-213.399	-91.457
3	3229.042	2260.328	968.713	392.676	271.239	-271.239	-271.239	145.307	0.000	-291.582	-124.964
4	3740.190	2618.132	1122.058	66.272	314.176	-353.477	-297.380	168.309	56.097	-147.097	-63.041
5	3950.328	2765.229	1185.099	0.000	331.827	-54.485	1.662	177.765	56.097	42.746	18.320
6	3889.562	2722.482	1166.780	0.000	326.698	64.474	124.300	175.017	56.097	115.108	49.332
7	3724.822	2607.374	1117.448	0.000	312.885	143.755	206.032	167.617	56.097	162.952	69.837
8	3492.033	2444.422	1047.611	0.000	293.331	180.505	243.919	157.142	56.097	183.177	78.505
9	3230.351	2261.205	969.107	0.000	271.349	219.609	284.232	145.366	56.097	147.511	63.219
10	3019.622	2113.734	905.888	0.000	253.648	254.561	320.265	135.883	56.097	142.264	60.970
11	2816.387	1971.469	844.918	0.000	236.576	288.903	355.669	126.738	56.097	148.814	63.778
12	2603.795	1822.655	781.140	0.000	218.719	324.007	391.859	117.171	56.097	155.948	66.835
13	2381.012	1666.707	714.305	0.000	200.005	359.942	428.906	107.146	56.097	163.716	70.164
14	2147.132	1502.990	644.141	0.000	180.359	396.780	466.884	96.621	56.097	172.176	77.790
15	1901.166	1330.814	570.352	0.000	159.698	434.604	505.877	85.553	56.097	181.389	77.738
16	1642.039	1149.425	492.614	0.000	137.931	473.500	545.976	73.892	56.097	191.421	82.038
17	1368.580	958.004	410.576	0.000	114.960	513.563	587.279	61.586	56.097	202.347	86.720
18	1079.514	755.657	323.856	0.000	90.679	554.899	629.892	48.578	56.097	214.245	91.819
19	773.450	541.413	232.037	0.000	64.970	597.619	673.934	34.806	56.097	227.201	97.372
20	448.676	314.211	134.666	0.000	37.705	641.848	719.531	20.200	56.097	241.311	103.419

KEY TO CALCULATIONS IS AS FOLLOWS.

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212.920 195.177 177.434 159.690 141.947 124.204 106.460 88.717
 70.973 53.230 35.487 17.743
 SUM OF DEPRECIATION ALLOWANCES 2714.733

PAYOUT TABULATION
 FEED PRICE: 45.00 CASE: COAL GASIFICATION--PHENOSOLVAN ANALYSIS
 PRODUCT PRICE: 8.1769 RATE OF RETURN ON EQUITY 15.00%
 WORKING CAPITAL: 102.126 DEBT AT 12.00%
 START-UP COSTS: 0.719 TRIAL 22

	1	2	3	4	5	6	7	8	9	10	11
	CPSTRT	EFFNCY	INCOME	PLTEXP	STLTAX	TXCRDT	FEDTAX	CASHFL	TOTRTN	AMORTZ	CAPEND
1	823.077	0.000	0.000	0.000	0.000	0.000	0.000	0.000	106.177	-106.177	929.253
2	2286.621	0.000	0.000	0.000	0.000	0.000	0.000	0.000	294.974	-294.974	2581.595
3	3124.542	0.000	0.000	0.000	0.000	0.000	0.000	0.000	403.066	-403.066	3527.608
4	3621.077	0.500	546.025	226.886	54.295	0.000	0.000	264.844	467.119	-202.275	3823.352
5	3823.352	0.800	873.640	266.703	54.351	0.000	0.000	552.585	493.212	59.373	3763.979
6	3763.979	0.900	982.845	280.216	57.904	0.000	0.000	644.725	485.553	159.172	3604.807
7	3604.807	0.950	1037.447	286.972	60.273	63.936	0.000	690.203	465.020	225.183	3379.625
8	3379.625	0.950	1037.447	286.972	61.372	80.293	0.000	689.103	435.971	253.132	3126.493
9	3126.493	0.950	1037.447	286.972	62.543	17.098	80.601	607.332	403.317	204.015	2922.479
10	2922.479	0.950	1037.447	286.972	63.589	0.000	113.262	573.625	377.000	196.625	2725.854
11	2725.854	0.950	1037.447	286.972	64.617	0.000	128.548	557.310	351.635	205.675	2520.178
12	2520.178	0.950	1037.447	286.972	65.667	0.000	144.174	540.633	325.103	215.530	2304.648
13	2304.648	0.950	1037.447	286.972	66.743	0.000	160.170	523.563	297.300	226.263	2078.385
14	2078.385	0.950	1037.447	286.972	67.845	0.000	176.567	506.063	268.112	237.951	1840.434
15	1840.434	0.950	1037.447	286.972	68.977	0.000	193.403	488.095	237.416	250.679	1589.755
16	1589.755	0.950	1037.447	286.972	70.141	0.000	210.716	469.618	205.078	264.540	1325.216
17	1325.216	0.950	1037.447	286.972	71.340	0.000	228.548	450.587	170.953	279.634	1045.582
18	1045.582	0.950	1037.447	286.972	72.577	0.000	246.946	430.952	134.880	296.072	749.510
19	749.510	0.950	1037.447	286.972	73.856	0.000	265.960	410.660	96.687	313.973	435.537
20	435.537	0.950	1037.447	286.972	75.179	0.000	285.645	389.651	56.184	333.467	102.071

THE LAST ENTRY IN THE FINAL YEAR SHOULD BE EQUAL TO THE WORKING CAPITAL INVESTMENT PLUS SALVAGE VALUE

WORKING CAPITAL RECOVERED INTACT AT END OF PROJECT LIFE = 102.126
 SALVAGE VALUE RECOVERED INTACT AT END OF PROJECT LIFE = 0.000
 TOTAL 102.126

KEY TO CALCULATIONS IS AS FOLLOWS.

- COLUMN 1 = CAPITAL INVESTMENT OUTSTANDING AT START OF YEAR INCLUDING WORKING CAPITAL
- COLUMN 2 = ONSTREAM EFFICIENCY (PLANT OPERATING FACTOR)
- COLUMN 3 = GROSS INCOME FROM SALES
- COLUMN 4 = OPERATING EXPENSES EXCLUDING TAXES
- COLUMN 5 = STATE AND LOCAL TAXES
- COLUMN 6 = FEDERAL INVESTMENT TAX CREDIT
- COLUMN 7 = NET FEDERAL INCOME TAX PAID
- COLUMN 8 = CASH FLOW AFTER TAXES
- COLUMN 9 = INTEREST ON DEBT AND RETURN ON EQUITY BASED ON COLUMN 1

COLUMN 10 = REDUCTION OF OUTSTANDING CAPITAL INVESTMENT, COLUMN 8-9
 COLUMN 11 = OUTSTANDING CAPITAL INVESTMENT AT END OF YEAR INCL WORKING CAPITAL, COL 1-10
 COLUMN 11 PLUS NEW INVESTMENT = NEXT ENTRY OF COLUMN 1

ADDITIONAL DETAILS OF CALCULATION COAL GASIFICATION--PHENOSOLVAN ANALYSIS

	1	2	3	4	5	6	7	8	9	10	11
	CPSTRT	DBSTRT	EQSTRT	DBTADD	INTRST	TAXABL	STXABL	EQRTN	PROPTY	DBTPAY	EQTPAY
1	823.077	576.154	246.923	576.154	69.138	-69.138	-69.138	37.038	0.000	-74.324	-31.853
2	2286.621	1600.635	685.987	950.157	192.076	-192.076	-192.076	102.898	0.000	-206.482	-88.492
3	3124.542	2187.179	937.363	380.063	262.461	-262.461	-262.461	140.604	0.000	-282.146	-120.920
4	3621.077	2534.753	1086.324	65.429	304.170	-340.963	-286.668	162.949	54.295	-141.592	-60.682
5	3823.352	2676.345	1147.006	0.000	321.161	-52.470	1.881	172.051	54.295	41.561	17.812
6	3763.979	2634.784	1129.195	0.000	316.174	62.401	120.304	169.379	54.295	111.420	47.751
7	3604.807	2523.364	1081.443	0.000	302.804	138.992	199.264	162.216	54.295	157.628	67.555
8	3379.625	2365.736	1013.888	0.000	283.888	174.551	235.923	152.083	54.295	171.192	75.939
9	3126.493	2188.544	937.949	0.000	262.625	212.387	274.930	140.692	54.295	142.810	61.204
10	2922.479	2045.734	876.745	0.000	245.488	246.221	309.810	131.512	54.295	137.637	58.988
11	2725.854	1908.096	817.757	0.000	228.972	279.453	344.070	122.664	54.295	143.973	61.703
12	2520.178	1764.124	756.055	0.000	211.695	313.423	379.090	113.408	54.295	150.871	64.659
13	2304.648	1613.252	691.396	0.000	193.590	348.195	414.938	103.709	54.295	158.384	67.879
14	2078.385	1454.868	623.517	0.000	174.584	383.842	451.688	93.528	54.295	166.566	71.385
15	1840.434	1288.302	552.132	0.000	154.596	420.441	489.419	82.820	54.295	175.475	75.204
16	1589.755	1112.827	476.928	0.000	133.539	458.078	528.219	71.539	54.295	185.178	79.362
17	1325.216	927.649	397.567	0.000	111.318	496.843	568.184	59.635	54.295	195.744	83.890
18	1045.582	731.905	313.677	0.000	87.829	536.839	609.416	47.051	54.295	207.250	88.822
19	749.510	524.655	224.855	0.000	62.959	578.174	652.030	33.728	54.295	219.781	94.192
20	435.537	304.874	130.663	0.000	36.585	620.968	696.147	19.599	54.295	233.427	100.040

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VITA

Dale R. Heron was born in Bristol, Tennessee on August 28, 1958. He attended elementary and junior high schools in Sullivan county and graduated from Sullivan East High School in June, 1976. The following September he entered The University of Tennessee.

After attending this university for two years, he entered into the cooperative engineering program in June, 1978 and began working for Kendall Fibers in Athens, Georgia as a P-3 Technician making the "Stay Dry" lining for Pampers diapers. After completing one work period of 10 weeks at this facility, he changed work positions in January, 1979 to work for Westinghouse Electric Corporation in their nuclear fuel fabrication facility at Columbia, South Carolina. Four work periods were completed as an Engineering Assistant at this facility to fulfill the requirements of the Co-op program.

On December 11, 1981 he recieved a degree of Bachelor of Science in Chemical Engineering from The University of Tennessee. In the winter quarter of 1982 he was formally accepted as a Graduate Teaching Assistant at The University of Tennessee and began work toward a Master of Science Degree. During this work, the author was responsible for the installation and maintenance of the ASPEN program on the university's computer system.

The author is a member of the student chapter of the American Institute of Chemical Engineers and the ASPEN User's Group where he serves on the Custody and Corrections Committee. Mr. Heron is planning to continue his education toward a Doctorate Degree at The University of Tennessee upon graduation.