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ENERGY

CONSERVATION

DEMONSTRATION OF A NEW ENERGY-SAVING PULP AND PAPER INDUSTRY SLUDGE-DISPOSAL SYSTEM

Final Report

By
Hans H. Peters

MASTER

August 1978
Date Published

Work Performed Under Contract No. ET-78-C-07-1687

Resources Conservation Company
Renton, Washington

U. S. DEPARTMENT OF ENERGY

Division of Industrial Energy Conservation

NOTICE

Available from the National Technical Information Service, U. S. Department of Commerce, Springfield, Virginia 22161.

Price: Paper Copy $7.25
Microfiche $3.00

Printed in the United States of America
USDOE Technical Information Center, Oak Ridge, Ten

IDO-1687-1
Distribution Category UC-95f

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SUMMARY

This demonstration program's primary purpose was to determine the net energy cost of drying pulp and paper mill sludge with use of Basic Extractive Sludge Treatment (B.E.S.T.), a solvent dewatering process. Additionally, this cost was compared with the cost of conventional sludge incineration to demonstrate obtainable energy savings through the use of the B.E.S.T. process.

Processing a blend of mill primary and secondary sludges normally pre-thickened by a coil filter to about 18% total solids, the system produced about ten tons of dry solids during the first quarter 1978, at the Crown Zellerbach Company's mill in Wauna, Oregon. During the first month, B.E.S.T. process parameters were established for the final production run. Pilot plant thermal, electrical and chemical processing costs were derived from operating data logged during the final phase: twenty-three days of sustained operation. During this period, 7.9 tons of 90% dry solids were produced from 45 tons of screened input feed (17.7% total solids).

The bulk of the dry sludge produced was burner/boiler tested as a fuel by the John Zink Co., at Tulsa, Oklahoma. The remainder, 250 lbs., was incinerated by Energy, Inc., of Idaho Falls, Idaho, with the use of a six-inch fluidized bed system. Laboratory standard heating value and fuel analyses tests were conducted by John Zink Co., Energy, Inc., Laucks Testing Labs/RCC, and Combustion Engineering, Inc.

Full scale (50 TPD) B.E.S.T. plant energy costs were calculated through the use of a computer program model and were compared to the estimated energy costs of operating a theoretical hog-fuel incinerator/boiler system.

Significant process parameters averaged as follows:

Centrifuge cake composition	26% solids, 35% TEA, 39% water
Input feed total solids	17.7%
Input feed suspended solids	16.6%
Input feed oil content (D.B.)	1.1%
Dry solids T.S.	89.6%
Dry Solids residual TEA	0.24%
Product water residual TEA	922 ppm
Product water total solids	5200 ppm
Product water suspended solids	3300 ppm

Some potentially toxic or carcinogenic compounds which may result from processing pulp mill sludge with triethylamine were periodically analyzed for in the B.E.S.T. flow streams.

After 21 days of operation, analyses of pilot plant outputs (solids, water and vent gas) and recycle solvent for specific amine compounds and polychlorinated biphenyl (PCB) showed concentrations of ethylamine and N -chlorodiethyl amine of less than 400 ppm and 200 ppm, respectively, in the recycled TEA. Diethylamine was detected at 690 ppm. PCB amounted to .05 ppm in the product water. Diethylabietamide levels of less than 50 ppm were found in the dry solids and recycled TEA and less than 5 ppm in the product water. These results are not considered to be significant because of the dilution effect of the fresh TEA added during the course of testing.

Pilot plant actual energy costs averaged 4175 btu (steam) and 0.71 Kwh per pound of input solids, with an overall thermal efficiency of 70%. Full scale (50 TPD) plant efficiency was projected as 83% because heat losses become a lesser fraction of the heat transferred as plant size increases. Energy costs for a 50 TPD plant were calculated as 2850 btu and 0.35 Kwh per pound of input solids.

Heat balances comparing a B.E.S.T. drying-burner-boiler, 50 TPD solids combustion system to a V-press-hog fuel incinerator-boiler, 50 TPD system show the B.E.S.T. drying model to save 4.9 million btu's/hour. This is equivalent to 8000 barrels of fuel oil yearly. Potential savings if the B.E.S.T. drying were applied to the entire industry would be in the order of 1,300,000 barrels per year.

Fuel analyses and heating value tests on composite dry solids samples produced the following average fuel composition (dry basis) and values:

- Composition:

Volatile solids*	69.5%
Ash	18.1%
Fixed carbon	12.4%
	100.0%

*(The solids averaged about 0.2% sulphur, 1.5% nitrogen, and 5.4% hydrogen.)

o High[1] heating value, Btu/lb. solids	6853
o Low heating value, Btu/lb. solids	6335
o High heating value, Btu/lb. volatiles	9860
o Low heating value, Btu/lb. volatiles	9115

The combustion tests projected no major problems with slagging, corrosion or gas emissions in a commercial size burner/boiler unit. Solids bridging at the burner feed inlet with resulting difficulty in maintaining a constant stoking rate is the only problem anticipated in using the solids as a fuel.

(1) "High" heating value includes latent heat of condensation of water vapor formed during combustion, "low" value does not.

CONCLUSIONS AND RECOMMENDATIONS

The dry solids produced by the B.E.S.T. pilot plant from pulp mill sludge were shown to have a useful heating value (6853 Btu/lb, D.B.) and to burn well in both a John Zink Co. "Combusticlone" burner, and in an Energy, Inc., fluidized bed reactor. Because of a strong tendency of the material to bridge, a mill-size burner will require a fuel feeder that supplies solids at predictable rates.

The B.E.S.T. drying model was shown to need about 50 percent less energy to dispose of the pulp mill sludge than is needed by the current practice of mixing damp (30% solids) sludge with wood chips and burning the resulting mixture in a hog fuel incinerator. For a 50 TPD plant, this energy savings is equivalent to 8000 barrels of fuel oil yearly (about $100,000). With credit for the heating value (as steam) of the solids produced, the net processing cost of a B.E.S.T. drying system was shown to be about $10/ton of input solids.

A finalized B.E.S.T. plant design will require additional test work on water still designs to show that 50 ppm TEA in the product water can be maintained at a maximum antifoam dosage of 1000 ppm. Additional test work on the effect of dryer residence time on TEA residuals in the dry solids is also recommended to show that a level of 0.2% TEA or less can be maintained.

An economic analysis of the cost of incinerating various blends of dried sludge (90% solids) with the wet sludge (18% solids) in a fluidized-bed system would also be useful.

1.0 SCOPE

Title:	Demonstration of a New Energy Savings Pulp and Paper Industry Sludge Disposal System
Subject:	Drying and Incineration of Pulp and Paper Mill Sludge.
Purpose:	To develop the net energy cost of drying pulp and paper mill sludge using a solvent dewatering process, basic extractive sludge treatment (B.E.S.T.), and to compare this cost with the cost of conventional incineration of sludge.
	To determine the fuel value and combustion characteristics of the dried solids produced.

2.0 INTRODUCTION

The national effort to reduce water pollution by upgrading wastewater treatment facilities has resulted in the production of ever increasing quantities and varieties of sludge. Many of the sludge types are difficult to dewater economically, and all sludges pose massive disposal problems. While sludge volumes continue to expand, the number of acceptable disposal options continues to drop. Increased energy costs have made conventional techniques for sludge drying and incineration extremely expensive. Land disposal options, on the other hand, are affected not only by higher energy and transportation costs, but also by more stringent environmental legislation. Thus, in the early 1970's it had become obvious that new technology was needed to provide a dynamic response to the sludge handling and disposal challenge.

An integrated systems approach developed by RCC was proposed to the D.O.E. to meet the need for a low energy, economical sludge disposal system. The basis of the integrated system and the element that makes the concept feasible is the recently developed Basic Extractive Sludge Treatment (B.E.S.T.) process.

The objective of the program proposed was to demonstrate the energy savings that can be produced by the use of an integrated sludge disposal system in the pulp and paper industry. This industry was chosen for the demonstration because it treats large quantities of wastewater and generates large volumes of sludge--with resulting high demands for energy. In addition, pulp mills are well distributed throughout the nation, and on an industry-wide basis energy and operating costs savings projected for the total pulp and paper industry would be substantial. Further, there is a potential of savings being achieved while maintaining high safety standards and minimizing environmental concerns.

The concept is illustrated in Figure 2-1. Thickened solids generated by pulp mill wastewater treatment facility were dried using the B.E.S.T. system, described in Section 2.1, and they were burned in a burner/boiler system to determine their fuel value.

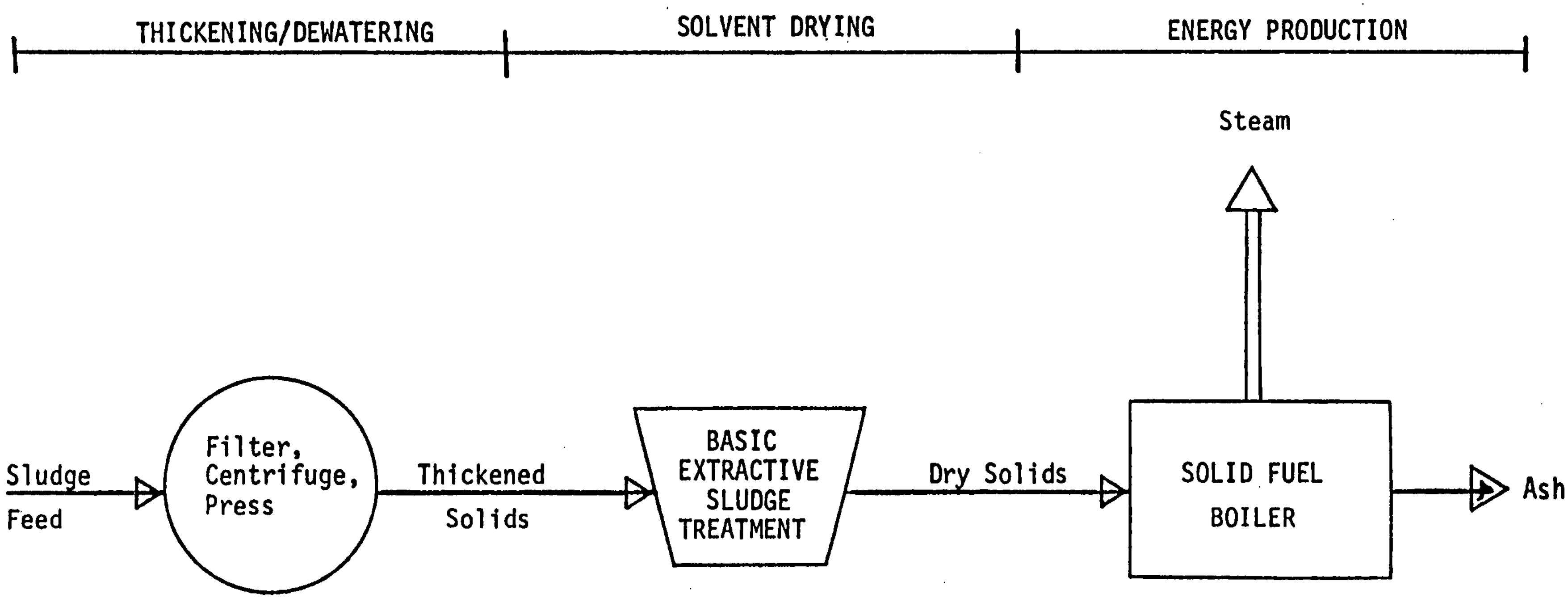


FIGURE 2-1. INTEGRATED SLUDGE DISPOSAL SYSTEM CONCEPT

2.1 B.E.S.T. PROCESS

A simplified process flow diagram of the B.E.S.T. system is shown in Figure 2-2. Sludge is continuously metered into the system, cooled (if necessary) to about 50°F and mixed with cold recycled solvent. In some cases, a pretreat chemical is added to the incoming sludge to raise pH and improve solids capture and product quality. Heat of solution is generated when the solvent and sludge are mixed resulting in a mix temperature of 60°F. This mixture is then cooled to 40°F before the solid liquid separation step.

This solvent-sludge mixture (approximately 6:1 volume ratio) enters a centrifuge where the solids are separated and delivered to a dryer. Liquid retained with these solids is vaporized in the dryer, condensed and returned to the decanter. The solids are discharged from the dryer. The solvent-water mixture (centrate) from the centrifuge is collected and heated with recuperative heat exchangers to 140°F. This stream is sent to a decanter where the warm mixture separates into a solvent phase (upper) and water phase (lower). The solvent phase is drawn off and cooled with recuperative heat exchangers, chilled and remixed with incoming sludge. Most oils and greases in the incoming sludge are dissolved by the solvent and remain in the solvent phase. These oils and greases are recovered by continuously distilling a small portion of the recycle solvent stream, returning fresh solvent to the stream. The water phase is pumped from the bottom of the decanter to a water still where any residual solvent is stream stripped and returned to the system.

In some cases, a third phase will form in the decanter at the solvent/water separation line. This phase, the "rag layer," is composed of solids which remain with the centrate due to incomplete separation in the centrifuge. When a rag layer starts to form, it is withdrawn from the decanter and reprocessed in the centrifuge, thereby increasing total system solids capture.

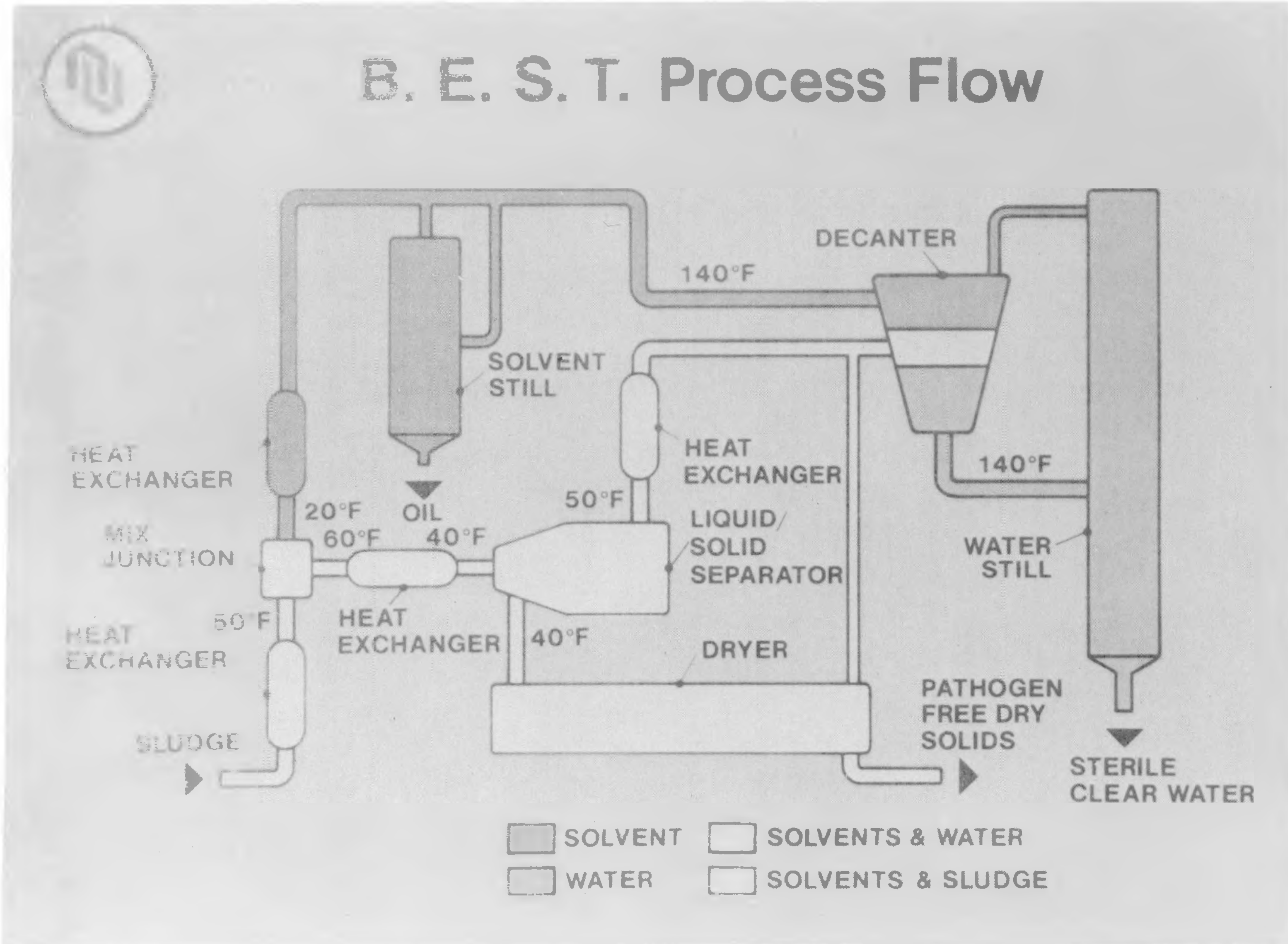


FIGURE 2-2 B.E.S.T. PROCESS FLOW

2.2 B.E.S.T. PILOT PLANT

The B.E.S.T. pilot plant is shown in Figure 2-3. Figure 2-4 shows the plant layout. It is mounted on an 8-foot wide, 40-foot long flatbed trailer which holds most of the support·utilities required for the B.E.S.T. Process. The external requirements are a sludge source, cooling water and 480V, 3-phase, 60-cycle power. Pilot plant steam (450 lbs/hr rated capacity) is supplied by a Williams and Davis 15 hp, 150 psig oil-fired steam-boiler mounted on the trailer. Refrigeration is provided by a Copeland 30 hp tandem unit with a water-cooled condenser.

The main level is surrounded by a field-erected catwalk for ease of access (23). An enclosed control room (2) houses gas analysis, temperature recording and process test apparatus, and motor control switches for unit startup and operation.

The B.E.S.T. pilot plant has a capacity of 1500 gpd at 20% solids input. Its purpose is to provide data which allow evaluation of the ability of the process to dry a specific sludge. In addition, design data, generalized operating costs, and equipment sizing can be derived which will allow the determination of economic feasibility of B.E.S.T. when operated as a full scale system.

The following major subsystems and components are shown in Figure 2-4:

- Chemical pretreat mix tanks and meter pumps (5).
- Water distillation column (8) consisting of five 6-inch diameter, 5-foot long sections packed with 5/8-inch stainless steel pall rings.
- Inverted cone decanter (11) with laminar flow sidearm feed and interface view ports.
- Main solvent chiller (14) where recycled solvent previously chilled to 40-45°F is chilled further to 15-20°F just prior to solvent/sludge mixing. The heat exchanger is a tube-in-shell unit with coolant in the tubes.

- Solid bowl centrifuge (16) mounted directly over the dryer (17). The centrifuge is a Bird 6-inch diameter solid bowl with adjustable scroll rpm. The dryer is a 16-inch diameter by 10-foot long steam jacketed Strong-Scott unit with internally rotating paddles.
- Two-stage condenser (18, 19). The first stage is tube-in-shell water cooled while the second stage is tube-in-shell cold solvent cooled.
- Solvent distillation column (21) consisting of two 6-inch diameter, 5-foot sections packed with 5/8-inch stainless steel pall rings.

One of the three sludge feed subsystems designed to handle the various sludges encountered is also shown in the pilot plant photograph. The pilot plant process flow diagram is shown in Figure 1-5. Letter locations refer to monitoring points.

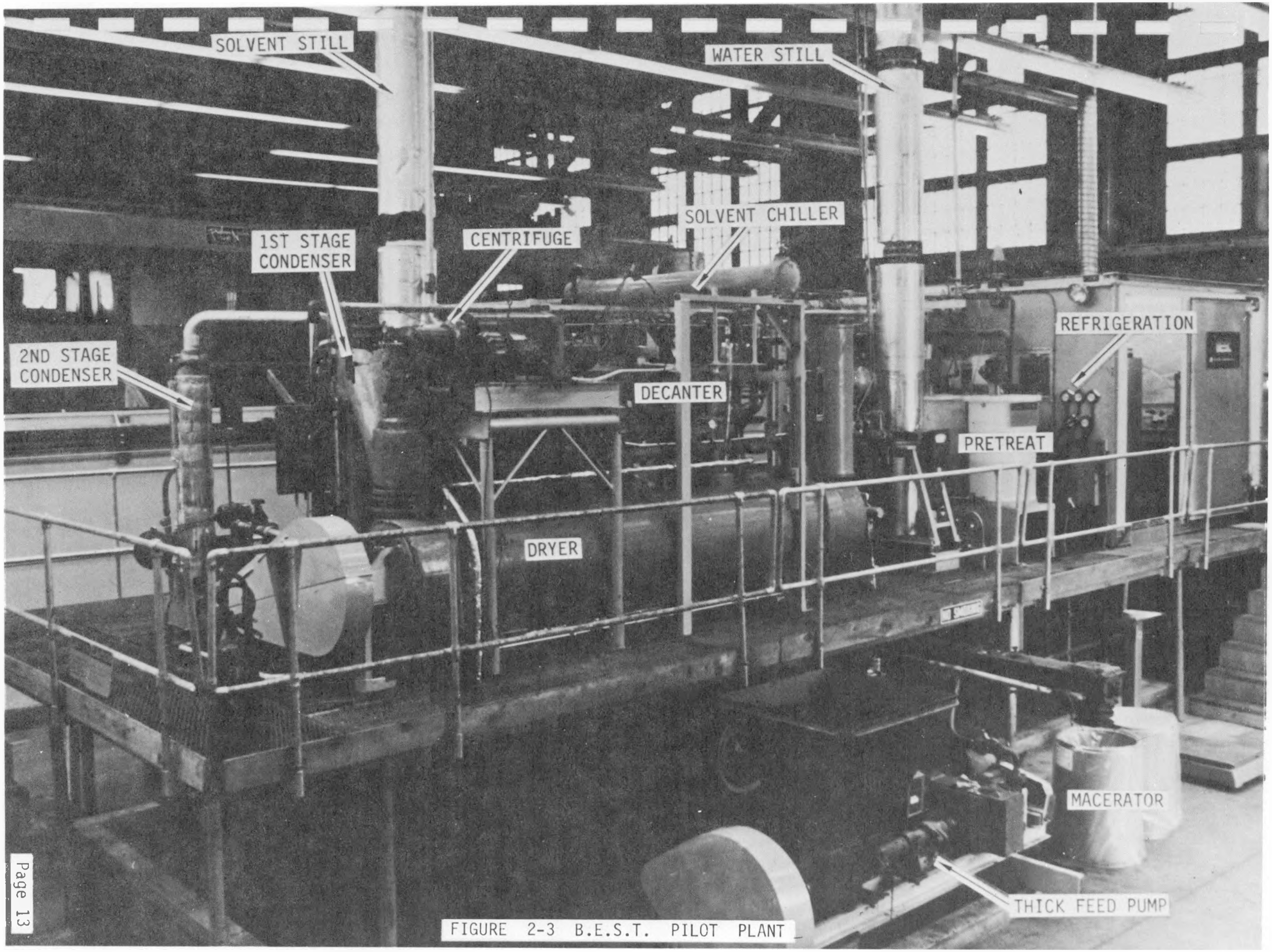


FIGURE 2-3 B.E.S.T. PILOT PLANT

1 STEAM BOILER
2 CONTROL ROOM
3 INSTRUMENT AIR SUPPLY
4 REFRIGERATION
5 CHEMICAL PRETREAT
6 ANTI-FOAM/CHEMICAL FEED
7 SLUDGE CHILLER
8 WATER STILL
9 INPUT SLUDGE PUMP
10 PRODUCT WATER COOLER
11 DECANTER
12 CENTRATE SUMP
13 SOLVENT INVENTORY TANK
14 SOLVENT CHILLER
15 POST-MIX CHILLER
16 CENTRIFUGE
17 DRYER
18 1ST STAGE CONDENSER
19 2ND STAGE CONDENSER
20 VENT GAS BLOWER & METER
21 SOLVENT STILL
22 SOLVENT COOLER
23 CATWALK

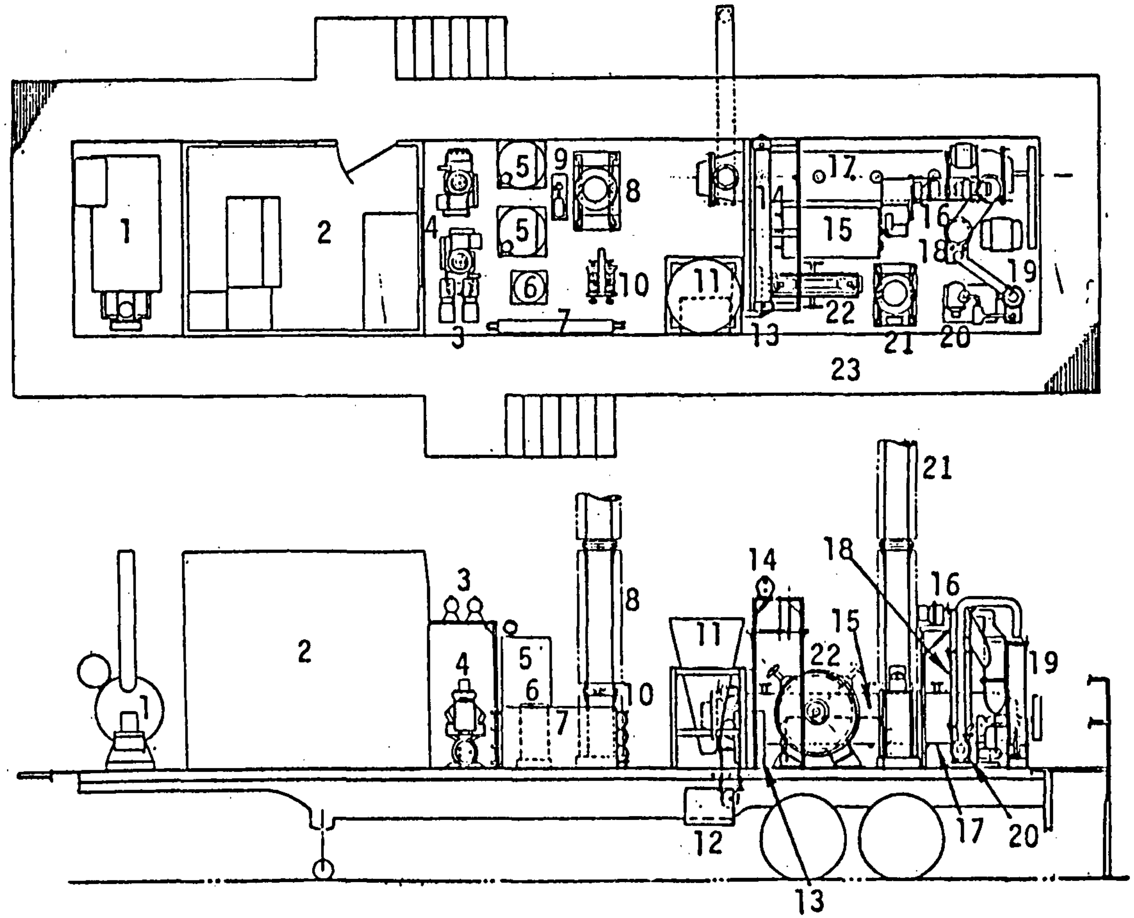


FIGURE 2-4 B.E.S.T. PILOT PLANT LAYOUT

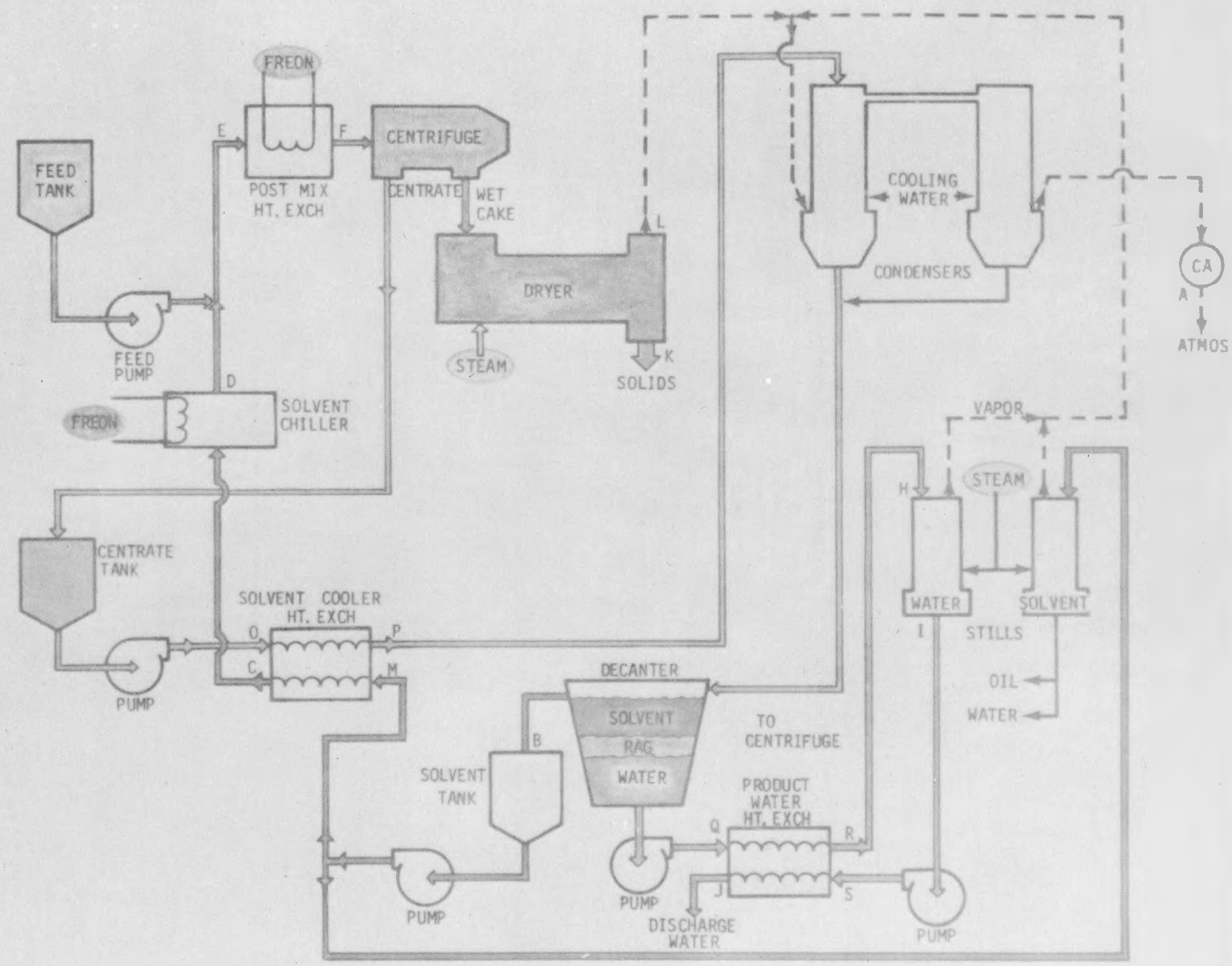


FIGURE 2-5 PILOT PLANT PROCESS FLOW

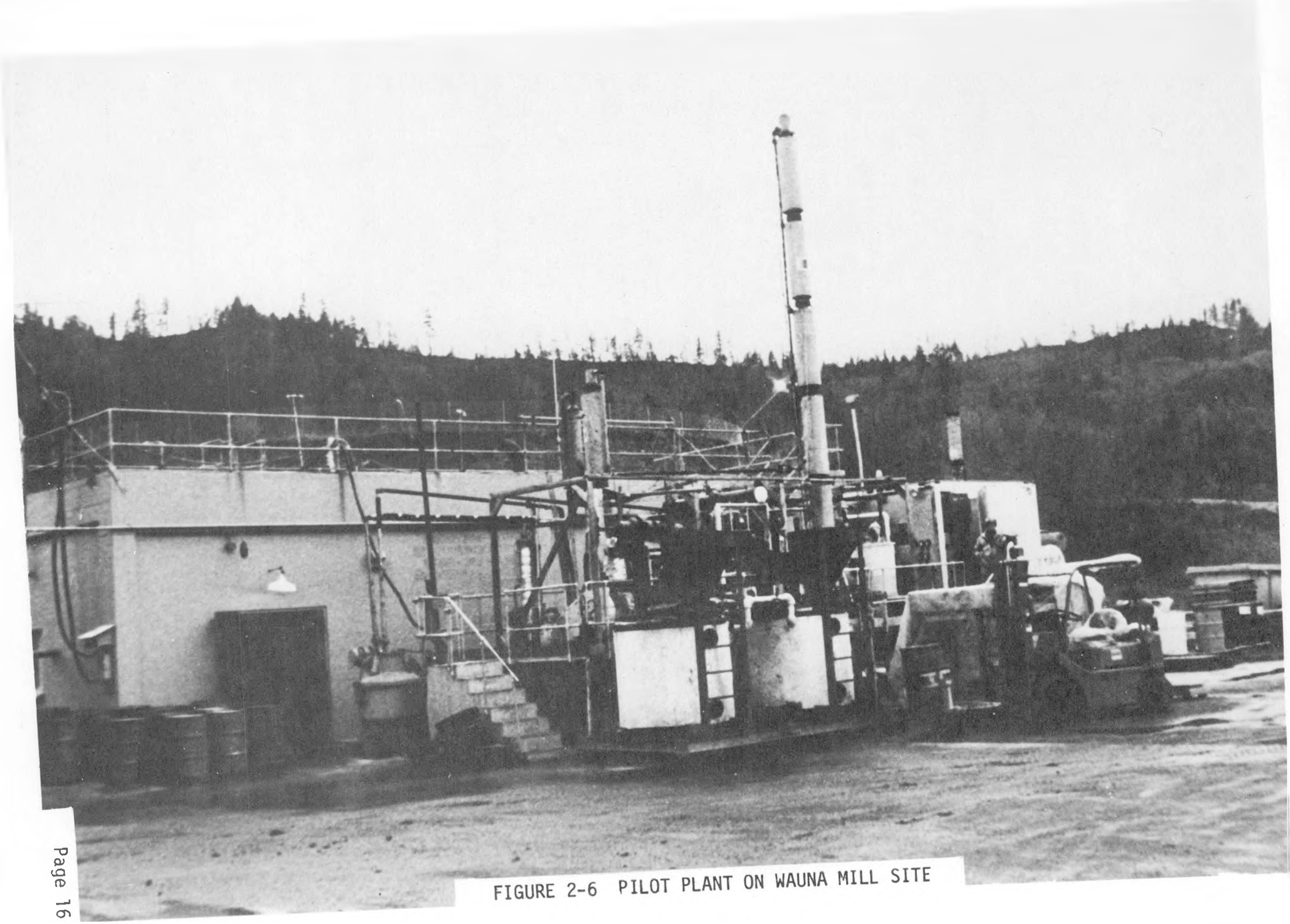


FIGURE 2-6 PILOT PLANT ON WAUNA MILL SITE

3.0 INITIAL TEST WORK

3.1 SLUDGE SOURCE

Testing was conducted on-site at the Crown Zellerbach mill in Wauna, Oregon, using the B.E.S.T. trailer-mounted pilot plant. The mill produces 360,000 tons of pulp and 280,000 tons of paper annually by use of two pulping techniques: a Kraft process and a groundwood process. The mill's water thruput is forty million gallons daily, and the wastewater system, as shown in Figure 3-1, incorporates both primary and secondary treatment systems. The primary and secondary sludges produced are blended and then thickened by coil filters. The resulting cake is landfilled at a site within a half mile of the sludge facility.

The mill averages about 40 TPD of primary sludge solids and 20 TPD secondary sludge solids. A portion of the mill's normal sludge output was B.E.S.T. processed periodically during the first quarter of 1978.

3.2 GLASSWARE LABORATORY TESTS AND RESULTS

On June 24, 1977, standard filter tests were run in the laboratory on a 21.1% total solids sludge supplied by Crown Zellerbach. The sludge was mixed with TEA in 4/1 and 6/1 TEA-Sludge volume to weight ratios, and filtered using a Modacrylic cloth. At a temperature of 40^{o}F, the filter time was one minute 28 seconds, and the composition of the wet cake was 19% solids, 59% TEA and 22% water.

Oven drying this cake produced an average of .13% TEA in the dry solids. From 0 to 200 ppm of Fleetcol 9181 antifoam was added to 200 ml samples at 95^{o}C, and foam heights were recorded. Foaming did not appear to be a significant problem.

CROWN ZELLERBACH
WAUNA MILL WASTE WATER SYSTEM

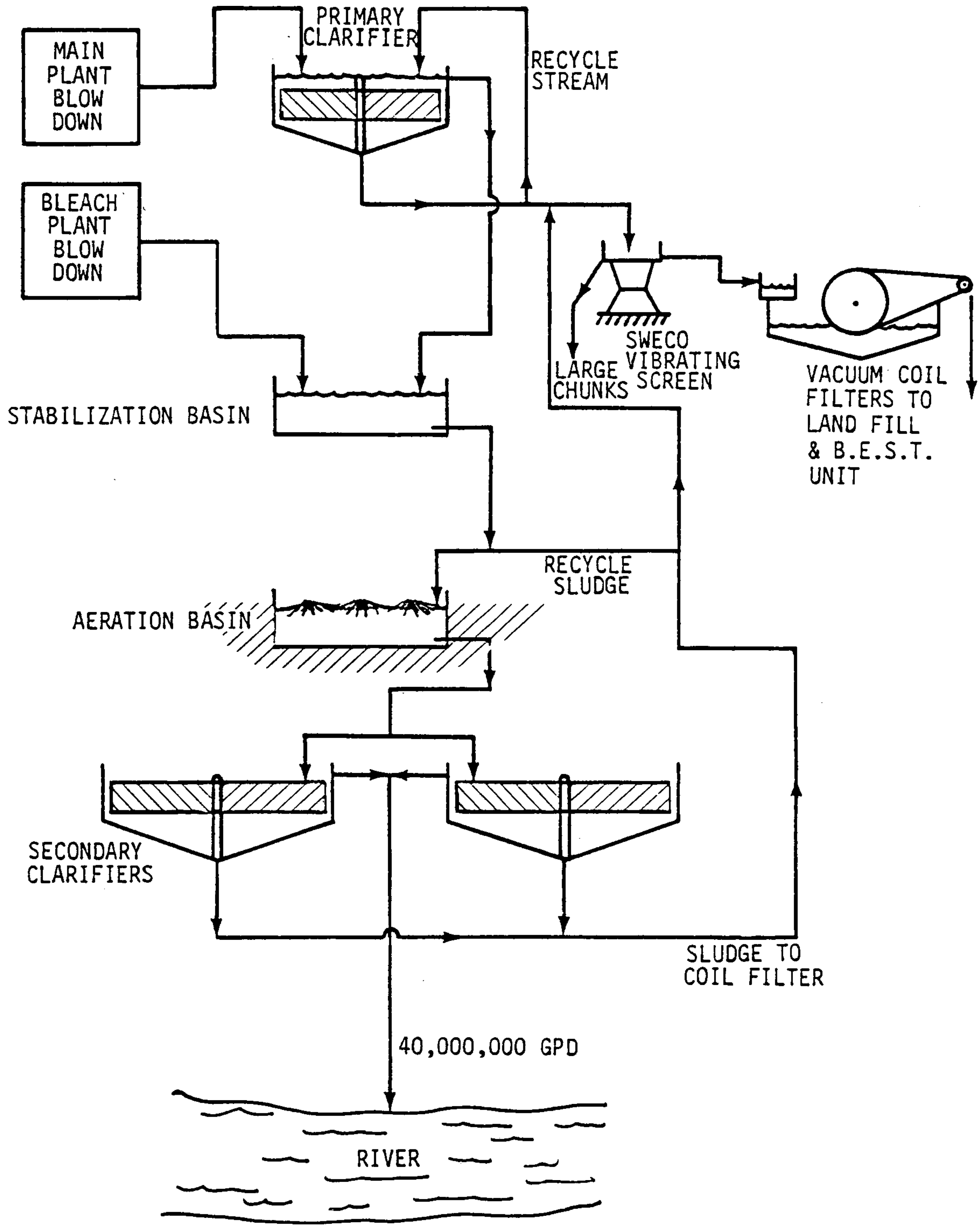


FIG. 3-1

3.2 GLASSWARE LABORATORY TESTS AND RESULTS (Continued)

On October 14, 1977, Crown Zellerbach provided samples of a mixture of primary and secondary sludge containing 16% total solids, of which 73% were volatile solids.

Filter tests were again run with a Modacrylic filter at 6 to 1 volume weight ratios (TEA-sludge) and pretreat levels of 0, 2, 4 and 6gms NaOH/l. Average filter time was 28 seconds, and the wet cakes produced averaged 24% solids, 54% TEA and 22% water.

Foaming tests again indicated that foaming in the process would not be a problem.

Oven drying the filter cakes resulted in an average of .24% TEA in the dried solids (maximum .62%, minimum .06%).

A pH titration test on the sludge produced the following:

NaOH gm/l	pH
0	6.7
1.0	9.2
2.0	11.0
3.0	12.1
4.0	12.5
5.0	12.9

Oil levels in the sludge were found to be less than .01% of total solids by the freon extraction method.

It was concluded that the Wauna sludge should process well with little or no pretreat and antifoam required. Centrifuge capture was predicted to be high (over 90%) on the basis of the relatively short filter times.

3.3 TEST PLAN 77-13

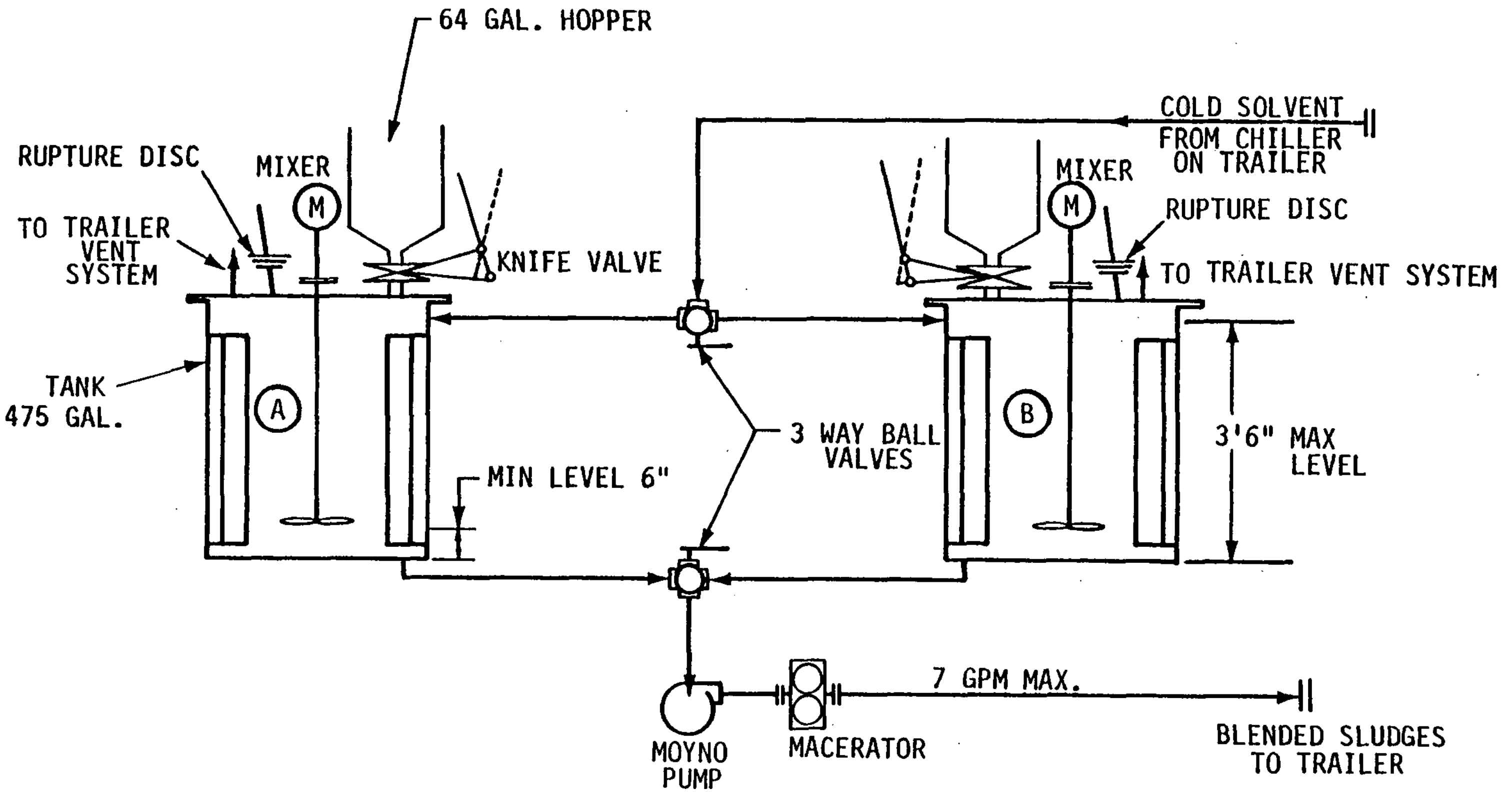
On December 5, 1977, the pilot plant was moved to the Wauna mill sludge facility. After assembly and utility hookups, checkout and startup commenced on December 12. Test plan 77-13 (RCC Document No. 105-ST-13015) outlined the methods of determining maximum sludge feed rate, minimum pretreat chemical feed rate, solvent still feed rate and water still antifoam feed rate. The test plan's objective was to establish optimum operating parameters for a continuous 21-day test.

3.4 OPERATIONAL PROBLEMS

A 5-hour water solvent baseline was established on December 16, using 6/1 solvent-to-water volume ratio. The thick sludge feed skid, illustrated in Figure 3.4, was used for inputing sludge to the pilot plant.

Initial sludge-TEA ratio tests were not completed due to centrifuge and line plugging. Large wood chips to 3/4" x 1/2" x 1/8" at various places in the inlet piping caused plugs which prevented any sustained runs of more than a few hours duration. After reviewing several techniques, it was decided to install a Sweco vibrating screen (1/4" mesh) on the inlet to the vacuum coil filters to remove chips and splinters. Following procurement and installation of the 60-inch diameter screen, testing was resumed on January 20, 1978. Testing continued per plan 77-13 until the start of the 21 continuous day test on 2/21/78. During this 12-day test period, it became apparent that plugging would continue to be a major problem. The Sweco screened the large chips (about 2-3% by weight) effectively, but 100% capture was not possible. The 3/4" to one-inch piping of the pilot plant continued to plug but at a slower rate.

Prior to the start of the final test phase, the pilot plant operated 24 hours/day processing the sixteen barrels of filter cake into 1120 pounds of dry solids. Of the thirty-seven shut downs experienced during this 12-day period, 28 were directly attributed to fiber/chip plugging.



BEST THICK SLUDGE
FEED SKID

FIGURE 3-4

4.0 FINAL TEST WORK

4.1 TEST CONDITIONS

From the initial test results, the final test plan 78-1 was prepared defining process parameters, e.g., 6/1 solvent-sludge ratio, .5 gpm sludge feed rate, no pretreatment, 110-115^{o}F decanter temperature, 230-260^{o}F dryer solids temperature and 200 ml/hour antifoam rate.

4.2 PRODUCTION

The final continuous run was started on February 21, 1978. The plant was operated on a 24-hour/day basis for 23 days; 7.9 tons of dry solids were produced from 45 tons of input feed. The feed, averaging 17.7% total solids, was composed of primary and secondary sludges mixed by the mill operators in varying proportions during the test period (an estimated range of 10 to 40% secondary). Including solids from preliminary testing, 10.5 tons of solids were accumulated during the program. This material was boxed and shipped to the John Zink Co., Tulsa, Oklahoma, for testing as boiler fuel.

Of the 542 hours of continuous run time available during the continuous test period, the pilot plant was on-line, processing sludge, for 350 hours. The 35% down time was caused primarily by fiber/chip blockages as described in the Significant Events Log, (Appendix A), and as shown graphically in Figure 4.2.

The relatively small (1") diameter of the B.E.S.T. pilot plant plumbing, along with tight bends, caused the frequent plugs. A full scale (50 TPD) plant would use 4 to 6-inch diameter piping, with large radius bends in conjunction with sludge screening as described in 3-4. Plugging events would then be reduced to a minimum.

4.2 PRODUCTION (Continued)

Causes of shutdown and frequency distribution are tabulated as follows:

Item	No. of Shutdowns
Dryer Plug	13
Feed Line Plug	28
Centrifuge Plug	11
Condenser Plug	7
Trim Heater Plug	1
Emulsion/Still Problem	8
Water Still Plug	9
Operator Error	1
Other	5

4.3 OPERATING PARAMETERS

Average, maximum and minimum values obtained for significant pilot operating parameters during the final test period are summarized in Figure 4.3. Letter designations shown correspond to monitoring locations given in Figure 2.5.

CROWN ZELLERBACH WAUNA MILL
BEST PILOT PLANT

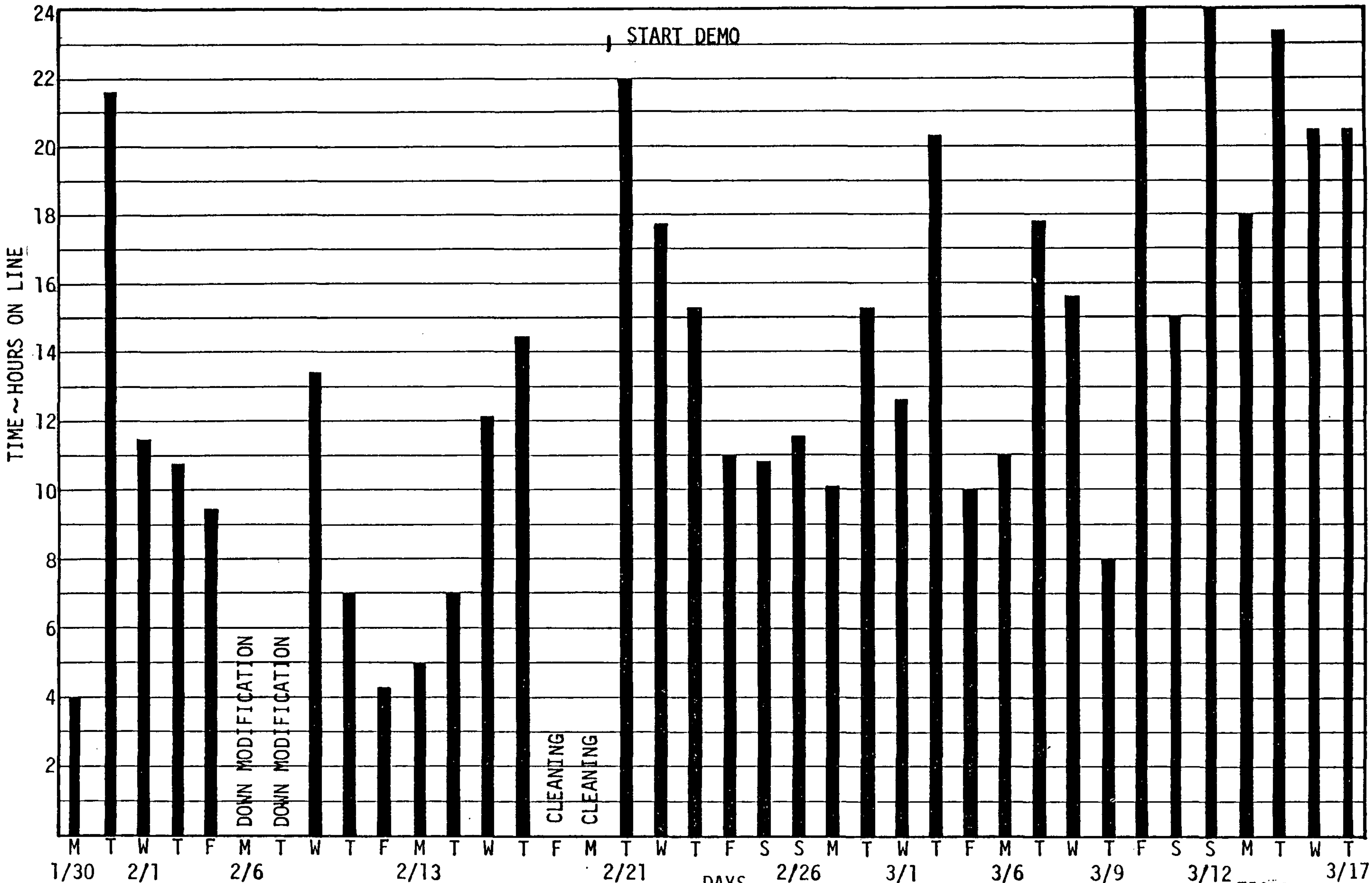


FIGURE 4.2

		Temperatures, °F		
		Maximum	Minimum	Average
A	Carbon Adsorption Unit, Carbon	99	41	65
B	Recycle Solvent	136	89	111
C	Cooled Solvent	64	36	49
D	Chilled Solvent	36	10	22.4
E	Mix Feed	64	41	57
F	Centrifuge Feed	58	34	44.3
G	Feed to Water Still	187	127	176
H	Water Still Top Vapor	180	145	166
I	Water Still Bottoms	220	208	210
J	Product Water Discharge	123	88	109
K	Dryer Solids Out	270	127	229
L	Dryer Gas Out	200	174	193
M	Solvent Cooler, Solvent In	136	89	111
N	Solvent Cooler, Solvent Out	74	60	65
O	Solvent Cooler, Centrate In	72	48	58
P	Solvent Cooler, Centrate Out	104	73	83
Q	Water Cooler, Decant Water In	116	80	101
R	Water Cooler, Decant Water Out	187	127	176
S	Water Cooler, Still Bottoms In	220	208	210
T	Water Cooler, Still Bottoms Out	123	88	109
	Dryer Steam Pressure, psig	141	114	137
	Water Cooler, Hot Side ΔP, psig	2	0	.9
	Solvent Cooler, Cool side ΔP, psig	4	0	2

FIGURE 4.3 OPERATIONAL PARAMETERS

5.0 TEST RESULTS

5.1 MATERIAL BALANCE

For the 350 hours of processing time, sludge cake input averaged 265 lbs/hr. The input/hr contained 46 lbs. of solids (including fat and oil). The dryer cake was produced at a rate of 46.8 lbs/hour at 89.6% total solids, equivalent to 41.2 lbs. dry basis solids recovered/hour. With suspended solids being 94% of total solids, (Appendix G), the overall system capture efficiency calculates as 96%.

System input/outputs are tabulated in Figure 5-1. Figure 5-1A shows a material balance in schematic form.

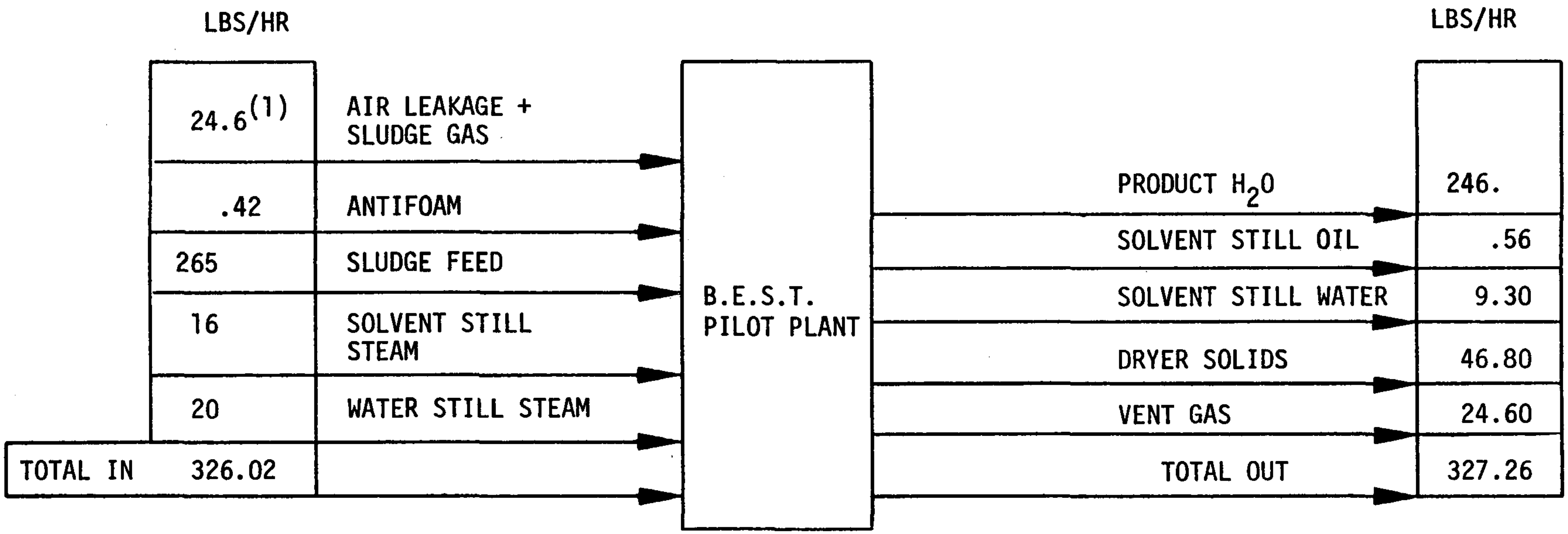
5.2 ANALYTICAL DATA SUMMARY

The average and range of the process stream analyses that were made twice daily during the 23-day final test period are tabulated in Figure 5.2. The percent of oil in the input sludge was obtained indirectly from solvent still steam usage in conjunction with the percent of oil in the recycled TEA. A typical calculation, given in Appendix J, results in a value of 1.53% of input solids as oil for a blend high in secondary sludge. A sludge mixture low in secondary sludge is calculated as 0.83% oil.

The percentage of suspended solids in the input sludge was determined by calculation to average 94%, i.e., 6% dissolved solids (see Appendix G). An average of 0.24% TEA in the dry solids (equivalent to $3.54/ton input solids) was obtained with a fixed residence time of solids in the dryer of about 12 minutes. Adjustment of the dryer rotor paddles would produce a longer residence time and may significantly reduce the residual TEA in the dried solids. The product water average residual TEA of 922 ppm (equivalent to $7.38/ton input solids) is attributed to poor feed distribution in the tower packing and between sections. A design goal of 20 ppm is considered attainable by a redesign of tower internal parts in a manner that will maintain uniform liquid distribution.

FIGURE 5.1 PILOT PLANT INPUT/OUTPUT
350 HOUR AVERAGES

INPUTS/OUTPUTS	Range Maximum	Minimum	Average
Sludge Processed, lbs/hr (W.B.)	315	165	265
Product Water, gal/hr	33.3	18.7	29
Dryer Solids Produced, lbs/hr	77.6	31.3	46.8
Vent Gas, CFH	431	143	333
Antifoam, mg/l still feed	2,841	0	1,624
Solvent Still Steam, lbs/hr	29.6	7.1	15.7
Solvent Still Oil, lbs/hr	.71	.32	.52
Solvent Still Water, lbs/hr	21.8	6.9	12.6
Water Still Steam, lbs/hr	32.3	19.0	19.8
Input Solids, lbs/hr	70.0	30.0	46.9
Process Time, hours/day	23.4	8	15.2



(1) Not Measured   Error = .38%

FIGURE 5.1A PILOT PLANT 350 HOUR MEASURED AVERAGE RATES OF INPUTS AND OUTPUTS

FIGURE 5.2 PROCESS STREAM ANALYSES - TWENTY-THREE DAY AVERAGES

	Range		Average
	Max.	Min.	
INPUT SLUDGE			
Total Solids, %	21.4	15.1	17.7
Suspended Solids, % (calc.)	20.1	14.2	16.6
Oil, % (calc.)	1.53	0.83	1.2
pH	8.3	6.5	7.4
Volatile Solids, %	81	70	75
DRY SOLIDS			
TEA, %	0.50	0.004	0.24
Total Solids, %	98.2	68.7	89.6
PRODUCT WATER			
TEA, ppm	4000	0	922
Total Solids, %	1.51	0.10	0.52
Suspended Solids, %	1.14	0.02	0.33
pH	9.6	7.4	8.7
WET CAKE			
Solids, %	29.1	20.8	25.7
TEA, %	49.5	23.0	35.3
Water, %	51.6	25.5	39.0
CENTRIFUGE CAPTURE, %	99	84	94
RECYCLE TEA, % OIL	2.23	0.61	1.07

5.3 ENERGY DATA SUMMARY

The energy delivered to the B.E.S.T. Pilot Plant averaged 4,178 Btu's and .71 kWh per pound of input solids processed. Energy costs were shown to be an inverse function of the solids feed rate (Figures 5-3A and 5-3B). Minimum energy costs of 2,770 Btu's and 0.45 kWh per pound of input solids were obtained for a solids input rate of 70.0 pounds/hr (3-17 and 3-18). Thermal (steam) energy usages by the dryer, water and solvent stills are summarized in Figure 5-3. Steam-use efficiency was about 70% by the above unit operations. As heat loss is proportional to the surface area/volume, it becomes less of a factor as system capacity increases, and steam-use efficiency in a full-scale plant is projected as 83% at design load operation.

FIGURE 5.3 ENERGY PARAMETERS
350 HOUR AVERAGES

	Btu/lb Input Solids		
	Maximum	Minimum	Average
Fuel Oil	13,936	3,541	6,555
Process Steam (delivered)	6,481	2,100	4,178
Net Process Steam (1)	4,779	1,413	2,908
Dryer Steam	4,166	1,793	2,611
Net Dryer Steam (1)	2,186	1,320	1,977
Water Still Steam	711	378	396
Net Water Still Steam (1)	513	161	216
Solvent Still Steam	582	90	314
Net Solvent Still Steam (1)	450	5	180
	Efficiency, %		
Dryer	83	66	75.7
Boiler	84	40	61.8
Process (thermal)	74	67	69.6
	Kilowatt Hours/lb Solids		
Pilot Plant	1.05	.42	.71

(1) Delivered steam minus steady state no load heat loss = net process steam.

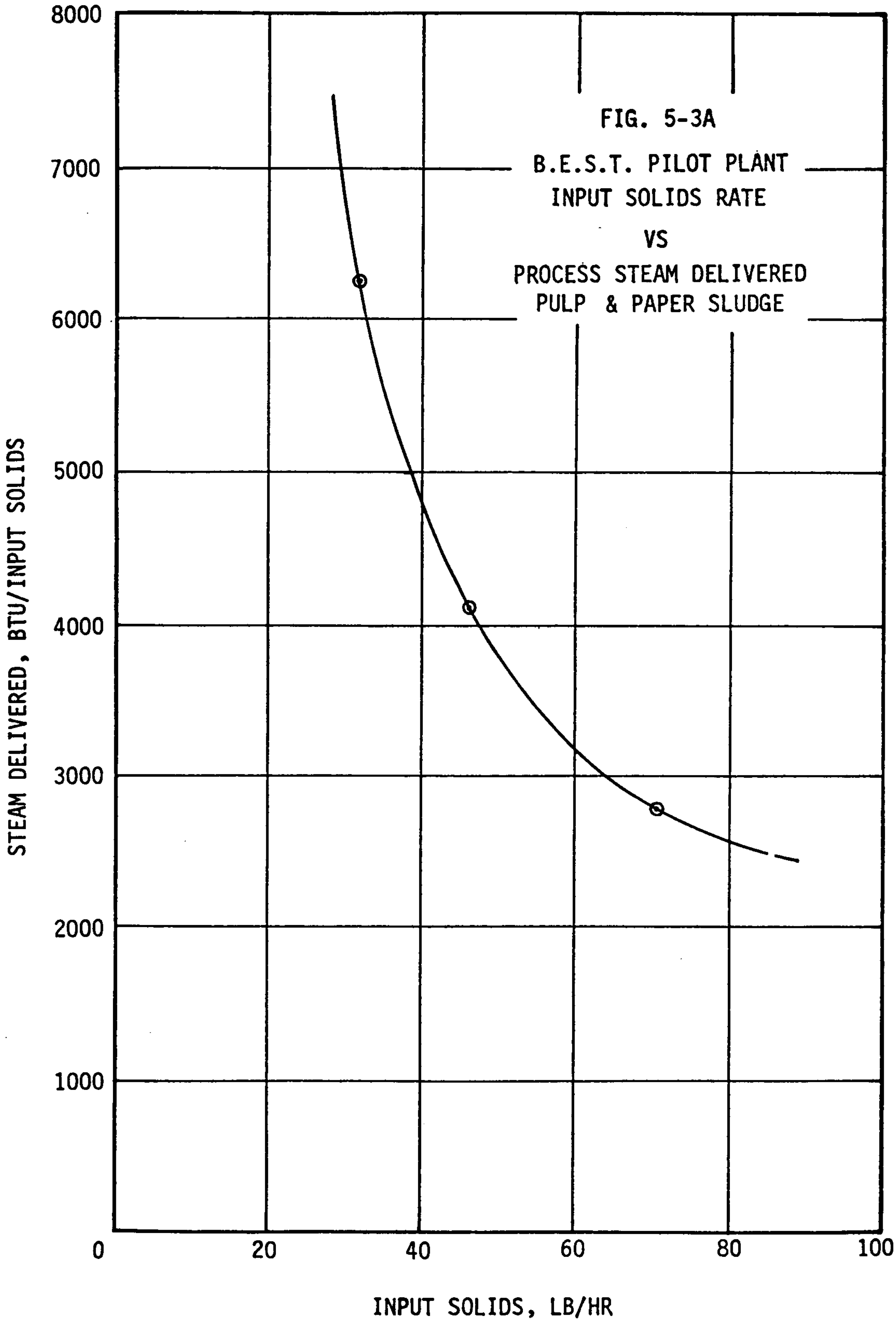


FIG. 5-3A
B.E.S.T. PILOT PLANT
INPUT SOLIDS RATE
VS
PROCESS STEAM DELIVERED
PULP & PAPER SLUDGE
8000
7000
6000
5000
4000
3000
2000
1000
0
20
40
60
80
100
STEAM DELIVERED, BTU/INPUT SOLIDS
INPUT SOLIDS, LB/HR

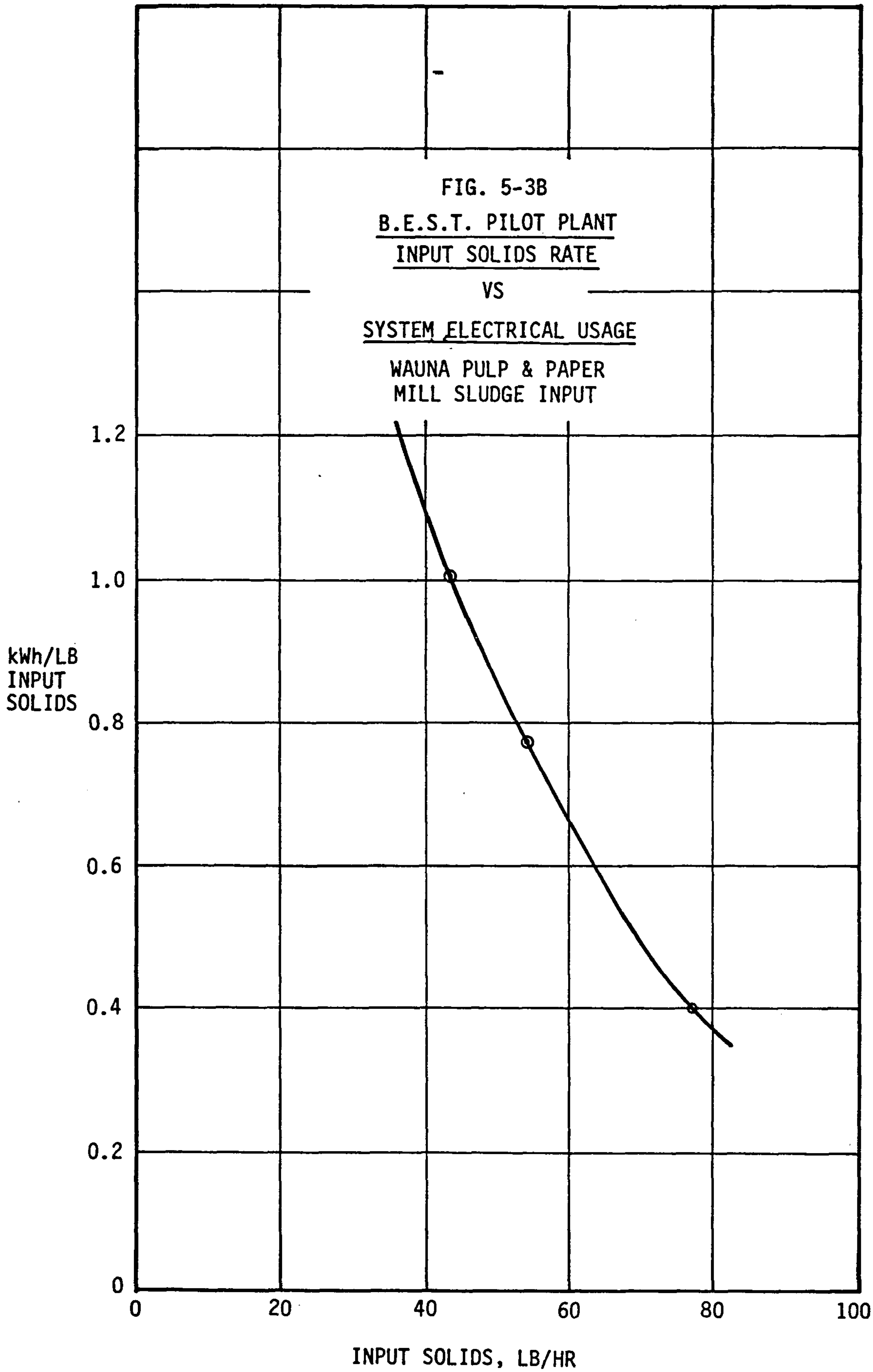


FIG. 5-3B
B.E.S.T. PILOT PLANT
INPUT SOLIDS RATE
VS
SYSTEM ELECTRICAL USAGE
WAUNA PULP & PAPER
MILL SLUDGE INPUT
kWh/LB
INPUT
SOLIDS
1.2
1.0
0.8
0.6
0.4
0.2
0
0
20
40
60
80
100
INPUT SOLIDS, LB/HR

6.0 HAZARDOUS CHEMICALS

6.1 DEMONSTRATION ANALYSES

The following process stream samples were taken on the 11th, 13th and 21st days of processing and were analyzed for the chemicals tabulated:

Process Stream Samples	Analyses for All Samples
Product Water	Diethylamine (DEA)
Dry Solids	Ethylamine (EA)
Vent Gas	N-Chlorodiethylamine
Recycle Solvent	Diethyl Abietamide
Solvent Still Oil*	Polychlorinated Biphenyl (PCB)

*One Sample

The levels of diethylamine detected in the above samples are plotted in Figure 6.1, versus process time. Maximum concentration found was 870 ppm in the recycle TEA after thirteen days of operation. After 21 days of operation, the level reduced to 690 ppm.

Ethylamine and n-chlorodiethylamine were detected at levels of 400 and 200 ppm in the recycle TEA after 11 days of operation. No increase in level was discerned after 13 and 21 days of operation.

Diethyl Abietamide levels were analyzed at less than 50 ppm in the recycle solvent and dry solids, and less than 5 ppm in the product water and vent gas absorbent.

PCB was detected at a level of less than 0.34 ppm in the input sludge and less than 10 ppm in the oil recovered from the recycle solvent by the solvent still. Dry solids contained less than .05 ppm PCB. Product water and vent gas absorbent contained less than .009 ppm PCB.

Complete analytical results and techniques used are given in Appendix F.

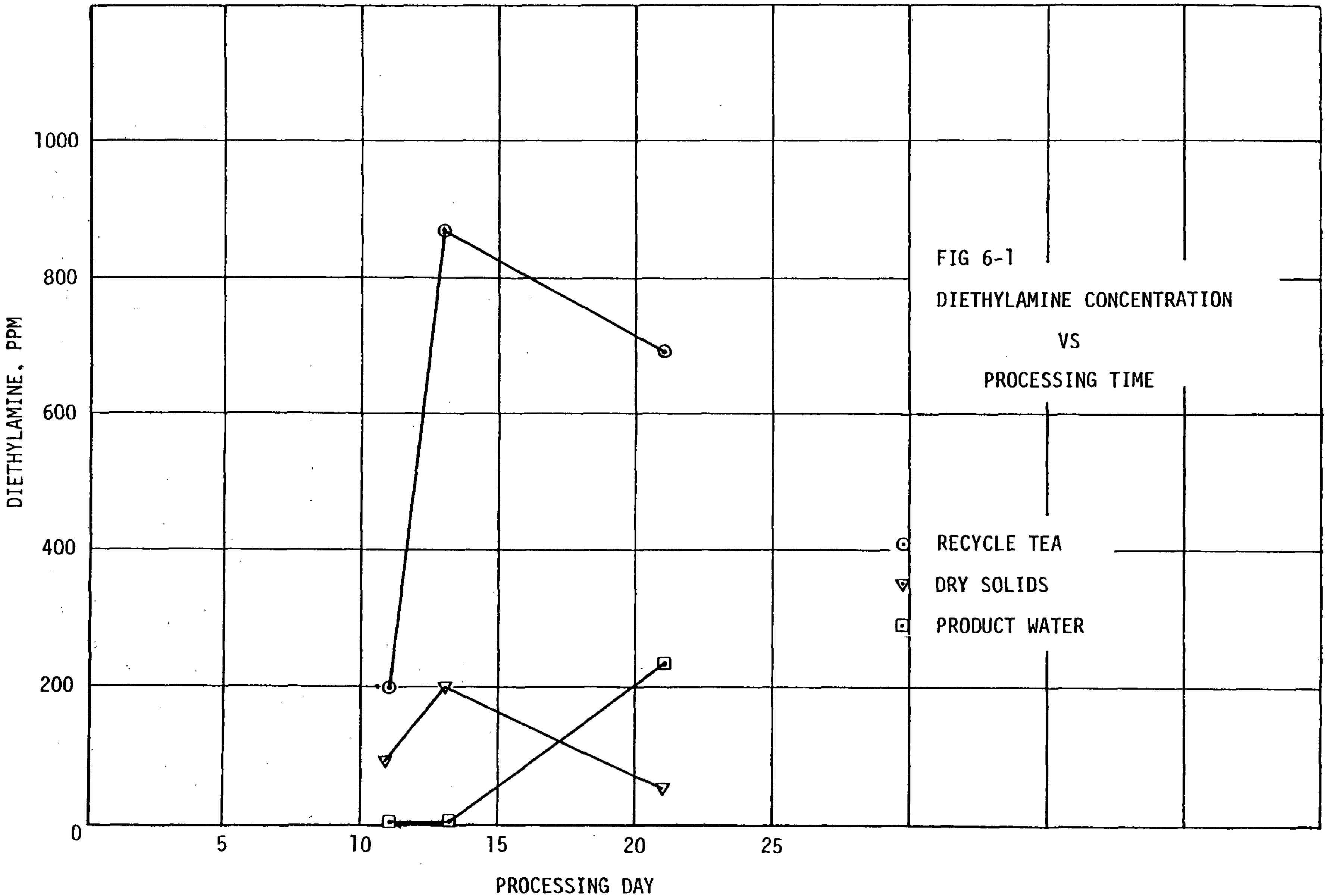


FIG 6-1
DIETHYLAMINE CONCENTRATION
VS
PROCESSING TIME

6.2 TRIETHYLAMINE, DIETHYLAMINE, ETHYLAMINE

Triethylamine (TEA) is the solvent used by the B.E.S.T. system. The TEA used in the system is basically self-contained and is continually recycled by the process. The process is subject to malfunctions as is any system. However, with adequate safeguards built into the system, leaks of TEA into the atmosphere due to a ruptured or leaking line can be quickly detected and appropriate shutdown procedures implemented. Both the product water and the solids can be sampled for high levels of TEA caused by process upsets. The design levels of TEA in the output streams of a full-scale plant (50 TPD - see Section 8.0) are as follows:

Product Water	-	50 PPM maximum
Product Solids	-	2,000 PPM maximum
Vent Gases	-	2,800 PPM maximum

Diethylamine (DEA) and Ethylamine (EA) are present in the system only because they exist as impurities in commercial grade TEA as used in the B.E.S.T. system. Commercial grade TEA contains, as a maximum, 0.2 percent DEA and 0.2 percent EA. Because of other allowable impurities in TEA (such as water and ethanol) and with a guarantee of 99.5 percent purity, the probable level of each DEA and EA in TEA is only 0.1 percent. A major difference between TEA, DEA, and EA relative to the B.E.S.T. process is their boiling points (TEA: 90^{o}C; DEA: 55^{o}C; EA: 39^{o}C). Thus, the dryer will more efficiently remove DEA and EA from the product solids than it will remove TEA. Steam distillation will do the same for the product water. The condenser will first condense TEA and then DEA and finally EA from the dryer vapor. Thus, it could be expected that the ratios of DEA and EA to TEA found in the various product streams would be altered from the original input ratios. However, during the demonstration, there was no evidence to indicate any buildup of the DEA or EA concentration levels in the recycled TEA.

The product water discharge (assuming a 17 percent solids input sludge feed at 50 TPD) will contain 25 pounds/day of TEA. The DEA and EA will each be less than 0.05 pounds/day. This compares to the following hazardous spill levels (into an aquatic environment) as defined by the EPA Water Programs:

TEA - 5,000 pounds/day
DEA - 1,000 pounds/day
EA - limits not defined

The product solids will contain 200 pounds/day of TEA and less than 0.4 pounds/day of each DEA and EA. These amounts, on a weight basis, are equivalent to 2,000 PPM (TEA) and 4 PPM (DEA and EA).

The vent will discharge 120 pounds/day of TEA and a maximum of 0.7 pounds/day of each DEA and EA. As a percentage (by volume), these discharge levels are 2,800 PPM for TEA, 34 PPM for DEA and 50 PPM for EA. The respective eight hour OSHA maximum allowable exposure levels for breathing atmosphere are 25 PPM, 25 PPM and 10 PPM. The vent gas will be naturally dispersed in the air to reduce exposure to less than OSHA limits.

The B.E.S.T. system processes only a small percentage of the wastewater leaving a treatment plant. At Wauna this amounts to 1.4% of the wastewater leaving the mill, or a dilution ratio of 73/1 when mixed with the total outflow. With a maximum TEA content of 50 PPM projected for a full scale B.E.S.T. plant, the wastewater entering the river would have a concentration of 0.7 PPM. The amine concentrations would immediately be reduced to undetectable levels by dilution with river water.

Triethylamine in an aqueous media (simulated lake or stream) has shown about a 50% biodegradibility in 100 days. Eventually it is assimilated by the natural environment.

6.3 N-CHLORODIETHYLAMINE

Very little information is available about this chemical. The level of its presence in the output streams of the B.E.S.T. process appears to be a function of the chlorine levels in the input sludge. No input chlorine data was taken during the running of this test, but there is no reason to believe that the sludge from the Wauna Mill contained lesser amounts of chlorine than would be found at other pulp and paper mills. The input chlorine level can also be reduced by oxidation processes should it cause an unacceptably high level of N-Chlorodiethylamine in any of the output streams. The laboratory analysis (Appendix "F") demonstrates that the output streams contained less than the following quantities of N-Chlorodiethylamine:

Product Water	-	1 PPM
Product Solids	-	5 PPM
Vent Gases	-	10 PPM

It is important to note that these levels are not true levels of concentration but rather reflect the limits of detectability of the laboratory analysis. Thus, the actual levels of concentration are probably much lower than indicated.

These levels probably would not be significantly changed by the improved operation of the water still as assumed in the development of the 50 TPD plant data in Section 8.0. In terms of total stream discharge, and assuming levels of detectability as the levels of detection, only one pound of N-Chlorodiethylamine will be discharged daily (50 percent of which will be in the product solids). There are no EPA or OSHA discharge standards for this chemical.

6.4 DIETHYL ABIETAMIDE

Very little information is available about this chemical. The level of its presence in the output streams of the B.E.S.T. process is a

6.4 DIETHYL ABIETAMIDE (Continued)

function of the abietic acid levels in the input sludge. No input data was taken during the running of this test. The laboratory analysis (Appendix "F") demonstrates that the output streams contained less than the following quantities of Diethyl Abietamide:

Product Water	-	5 PPM
Product Solids	-	50 PPM
Vent Gases	-	5 PPM

It is important to note that these levels are not true levels of concentration but rather reflect the limits of detectability of the laboratory analysis. Thus, the actual levels of concentration are probably much lower than indicated.

These levels probably would not be significantly changed by the improved operation of the water still as assumed in the development of the 50 TPD plant data in Section 8.0. In terms of total stream discharge, and assuming levels of detectability as the levels of detection, 7.5 pounds of Diethyl Abietamide will be discharged daily (65 percent of which will be in the product solids). There are no EPA or OSHA discharge standards for this chemical.

6.5 PCB's

PCB's pass through the B.E.S.T. process unaltered. The PCB level in the output streams will be directly proportional to the PCB level in the input sludge. The laboratory analysis (Appendix "F") demonstrates that there is no PCB buildup in the recycled TEA. The product water and vent gas PCB levels were both less than 0.009 PPM. The product solids level was 0.05 PPM. At this level, less than 0.1 pound/day of PCB's would be discharged from a 50 TPD operation using the Wauna mill sludge. Of this, 70 percent will be in the product solids. The EPA has established 10 pounds/day of PCB spill into an aquatic environment as constituting a hazardous spill.

6.6 GENERAL

The B.E.S.T. system is operated at a slight negative pressure so that all leakage is inward. The vent to atmosphere is first scrubbed with activated carbon, leaving a maximum of 3000 PPM of triethylamine for dispersal in the atmosphere from an elevated stack.

The B.E.S.T. pilot plant has been monitored, while operating inside a building, for a four day period at several locations using an automatic sequential sampler. Analysis of the samples for TEA showed one out of twenty-four in excess of the threshold limit of 25 PPM (28 PPM, max.). The 28 PPM was immediately noted by the operators since levels of one PPM can be easily detected by the human nose. The pilot plant design has since been modified to eliminate the source of the above leakage (a dryer circulating fan bearing).

Subsequent pilot plant testing at Portland and Wauna has shown vent gas odor not to be a problem. No complaints were received in a total of three months of continuous operation at either location from plant personnel or local residents.

If an odor problem is predicted for a full-scale plant, the vent gas could be ducted to the gas burner of a steam boiler and the amines oxidized to nitrogen, water vapor and carbon dioxide. Hydrogen sulfide has never been detected in four years of processing a variety of municipal sewage sludges and pulp and paper sludges.

Adequate interlocks and cut-off mechanisms are incorporated into the system to prevent unacceptable operating conditions such as improper (too high or too low) flow rates or too high pressure.

6.6 GENERAL (Continued)

With regard to the overall operation of the B.E.S.T. process, the preventative measures required for safe operation would be taken during design. The only chemical present in significant quantities from an operational point of view is triethylamine. The other chemicals are so dilute (2,000 PPM maximum in TEA) that toxic concern for them during a system upset can be discounted. Their presence is too small to be of any concern from a contributory standpoint (explosion, corrosion, etc.). TEA, however, does require some safeguards due to its flammability and toxicity.

A fire and explosion analysis of the B.E.S.T. pilot plant was prepared by Hazard Research Corp., N.J. In this study, design recommendations for blow out disc installations in the dryer, decanter and dryer condenser were made. A vent flame arrestor and a safe start-up and shutdown procedure were recommended. A three variable explosibility diagram, air-water-TEA, was prepared using lab test results. From this analysis, a hazardous concentration can be predicted.

In four years of pilot plant operation, under less than ideal conditions, there has been no incidence of fire or explosion, i.e., triethylamine is a relatively safe solvent to use in a process system.

With respect to contamination levels of the six chemicals in the solids (TEA, DEA, EA, N-Chlorodiethylamine, Diethyl Abietamide and PCB's), the utlimate use of the solids will be determined by the chemical and biological makeup of the dry solids as well as economical considerations. In all cases, the levels of the above chemicals are too low to cause concern during the physical handling of the solid product.

7.0 FUEL CHARACTERISTICS OF DRY SOLIDS

7.1 LAB RESULTS

Composite samples of the dried cake were analyzed for high heating value (HHV), moisture, volatiles, fixed carbon and ash by the John Zink Co., Energy, Inc., Laucks Testing/RCC labs and Combusion Engineering, Inc.

Results are summarized in Figure 7-1. Converting the "as received" data to a dry basis gave the following typical composition and value for the dried Wauna Mill Sludge solids:

- o Composition:

Volatile Solids	69.5%
Ash	18.1%
Fixed Carbon	12.4%
	100.0%

- o High heating value 6853 Btu/lb

The variations in C.E. and Energy, Inc., ash analyses, shown in Appendices E and K, are probably due to a variable solids composition caused by day-to-day changes in proportioning the weight of primary to secondary sludge and changes in the composition of the waste streams leaving the mill. Major ash composition differences were in the amounts of iron, silicon and titanium oxides reported.

From the hydrogen analysis, [1] the low heating value used for calculating the heat balance of Figure 9-3 was determined as 9115 Btu/lb of volatiles (Figure 7-1).

1 J. J. Kurpen, Director, Kreisinger Development Laboratory, C-E Power Systems Combustion Engineering, Inc., Windsor, Connecticut, provided the Appendix K data.

ANALYSIS By	Solids	Water	V.S.	F.C.	Ash	Total	V.S. D.B.	HHV AS REC'D BTU/LB	ΔH H_2O, BTU/LB	HHV D.B. BTU/LB	BTU/LB VOL.
Zink(Avg.)	.867	.132	.508	.145	.213	.999	.586	5370	153	6370	10870
Zink(Wet)	.705	.295	.383	.126	.195	.999	.543	4430	339	6764	12457
Zink(Dry)	.972	.028	.658	.158	.156	1.0	.677	6360	32	6576	9713
Comb. Engr.	.856	.144	.639	.117	.10	1.0	.746	6000	166	7203	9655
Energy, Inc.	.960	.041	.706	.109	.144	1.0	.734	6392	47	6715	9111
"								6327	47	6647	9019
Laucks/RCC	.924	.076						6600	87	7237	9727
"	.927	.073	.688				.744	6505	84	7108	9554
"	.924	.076						6430	87	7053	9480
Weighted Avg.	.900	.100	.629	.110	.161	1.0	.695	6046	115	6853	9860

H_2O = $(470^o - 70^o) + 750 = 1150$ BTU/LB

V. S. = Volatile Solids

D. B. = Dry Basis

HHV = High Heating Value

F. C. = Fixed Carbon

Low Heating Value (LHV) calc:

H_2 in solids (% D.B) = 5.37, resulting water = $[.695/.746](.0537) \times 9 = .45$ lb,
.45 lb H_2O (1150 BTU/lb) = 518 BTU/lb solids, LHV = 6853-518 = 6335 BTU/lb
LHV of volatiles = 6335/6853 X 9860 = 9115 BTU/lb volatiles

Fig. 7-1 LAB FUEL ANALYSES OF PULP & PAPER MILL DRIED SLUDGE

7.2 BOILER/BURNER TESTS

John Zink Co. Test

Fuel evaluation tests were conducted on the Wauna sludge by the John Zink Co, Tulsa, Oklahoma, using a cyclonic burner design (Combusticlone). The primary test objective was to evaluate the solids as a fuel for energy production. To this end, the following tasks were accomplished.

1. Analysis of gaseous products of combustion.
2. Analysis of residual solids.
3. Heat balance of the burner/boiler system for heat recovery, thermal efficiencies and heating value.
4. Comparison of actual burner heating value with laboratory calorimeter data.
5. Stoking characteristics of the solid waste.
6. Suitable feeding methods.
7. Material balance of waste solids and combustion air.
8. Fuel/air ratio that could be used in commercial operation.
9. Commercial operating temperatures and pressures.
10. Boiler efficiency.
11. Quality of flue gases with respect to SO_x and ash contamination.
12. Methods of ash removal from the unit and gas stream.

The 5'6" I.D. x 16' long horizontal, refractory lined, cyclonic combustion vessel was operated at feed ratio from 200 to 1200 lbs/hr. Total burn time was 36.5 hours during which 14,810 lbs. of solids were incinerated.

Constant feed rates were difficult to establish because of bridging of the dried solids. Nine different methods were tried, none of which were completely successful.

Sulphur oxides ranged from 1.1 to 168 ppm in the flue gas while ash content averaged .25 grains/standard cubic foot. EPA Anderson classifier train tests indicated that 80% of the particulates were between 25 and .34 microns, with 10% above 25 microns and 10% below .34 microns. The ash concentration in the flue gas amounted to 1.7 grains/ft^3. This concentration would require particulate removal from the flue gas stream by bag filter, venturi scrubber or an electrostatic precipitator.

Steam production for a commercial system was projected at 4.4 pounds of steam per pound of dried solids incinerated.

The concentration of sulphur oxides (SO_x) in the flue gas was found to be 168 ppm which is considered low enough for venting to the atmosphere.

Combustion of the dried solids in the "Combusticlone" was 99.5% complete. The burner was estimated to have an incinerator capacity of 3.1 lbs/hr of dried solids per cubic foot of burner volume.

Additional conclusions are given in the test report by John Zink Process Systems test personnel (Appendix D).

Energy, Inc. Combustion Test

Laboratory scale, fluidized bed combustion tests of the Wauna Mill sludge solids were conducted by Energy, Inc., at Idaho Falls, Idaho.

The 6" diameter burner system (diagramed in Appendix E) was operated during three test periods at solids feed rates ranging from 11 to 26 lbs/hr. Total burn time was about 14 hours.

Constant feed rates were difficult to establish because of the bridging tendency of the dried solids.

Nitrogen oxides emissions in the exhaust gas were found to be less than one ppm. Cyclone collected ash particulates ranged from 32% larger than 50 mesh (297 microns) to 34% less than 400 mesh (36 microns).

These initial tests indicate there will be no major problems with slagging, corrosion or gas emissions in a commercial size unit. Maintaining a constant feed to the bed appears to be the chief problem, and a high pressure air transport system is recommended for stoking the burner.

8.0 FULL SCALE B.E.S.T. PLANT PROCESSING COSTS

Calculation of the thermal, electrical and chemical processing costs of a 50 TPD plant for drying pulp mill sludge is based on the computer program input factors given in Figure 8-1.

The program printout (Figure 8-2) provides composition, flow rates, energy and temperatures at 68 process points in the system. Results are given per pound of input suspended solids (I.S.S.). A processing costs summary is given in Figure 8-2D. Location of points calculated (station locations, "sta") are shown in Figure 8-3.

B.E.S.T. processing costs for Wauna Mill sludge were projected, using current Northwest energy rates, as follows:

	$/Ton of Input Solids
Chemical	$12.14
Electrical	8.87
Thermal	14.80
	$35.81
Value of steam produced from dry solids (see Section 9.4)	- 26.36
Net Cost	$ 9.45

For comparison purposes, test results from pilot plant operations, conducted for Seattle Metro on a blend (20% solids) of primary and waste activated sewage sludges, were projected as follows (no credit shown for fertilizer value of dried solids).

	$/Ton of Input Solids
Chemical (KOH pretreat)	$20.01
Electrical	6.17
Thermal	13.90
	$40.08

Capital costs for a Wauna plant (turnkey) were projected at 3.7 million. Amortization costs (15 years at 8%) were calculated as $23.72/ton of input solids.

FIGURE 8-1

B.E.S.T. BASIC INPUT DATA

PROJECT TITLE WAUNA BLEND

PROJECT I. D. 50 TPD

C(1)	Sludge Solids Fraction, Suspended	.166
C(2)	Sludge Temperature, ^{o}F	60
C(8)	Combined Condenser Temperature, ^{o}F	130
C(9)	Oil Fraction of Input Solids	.012
C(11)	Wet Cake Solids Fraction	.257
C(12)	Wet Cake TEA & Oil Fraction	.353
C(14)	Pretreat Solids/LB Input Sludge Solids	0
C(15)	Solids Fraction Pretreat Solution	.45
C(16)	Specific Heat of Pretreat Solution	.90
C(21)	Recycle Solvent Oil Fraction	.02
C(41)	Solvent to Water Still-Decant Bottoms	.05
C(53)	Total Solids Fraction of Input Sludge	.177
C(55)	Product H_2O Exit T, ^{o}F (from system)	90
C(82)	LBS Input Solids/Day, Suspended	93,785
C(83)	Sludge Density, LBS/GAL	7.7
C(84)	$/Kilowatt Hours	.0125
C(85)	%Boiler Eff./100	.85
C(86)	%Process Eff./100	.83
C(87)	$\$/10^6$ BTU/Steam (as fuel oil)	2.20
C(88)	$/LB Pretreat Chemical	.175
C(89)	PPM Antifoam	1000
C(90)	$/LB Antifoam	.68
C(91)	$/LB TEA	.785
C(92)	Refrigeration Condenser Cooling Water T	20

FIGURE 8-1

B. E. S. T.

DEFAULT PARAMETER CHANGES

C(19)	130	decanter Temp.
C(28)	.06	lbs non/cond./lb input S.S.
C(31)	16	molecular wt. sludge gas
C(44)	0	TEA in solvent still water
C(45)	0	oil in solvent still water
C(46)	0	TEA in solvent still oil
C(47)	.4	water fraction solvent still oil
C(49)	0	oil fraction wet cake
C(54)	169	Feed to water still
C(116)	.0003	Solvent still bottoms TEA fraction
C(64)	.002	TEA fraction dry cake
C(25)	.00005	TEA fraction product water
C(99)	.0028	TEA fraction vent gas

PROJECT: WAUNA BLEND

FIGURE 8-2

DATE RUN: 25-Jul-78, 04:19 PM

ACCOUNT NO: 78-BEST-3

STA	WT FLOW LB/LBISS	TEMP DEG F	SPHT BTU/LB	ENERGY BTU/LBISS	ENTHALPY BTU/LB	AMMONIA LB/LBISS
1	6.08410	60.00000	0.89994	328.51778	53.99615	0.06000
2	0.00000	60.00000	0.90000	0.00000	54.00000	0.00000
3	6.08410	60.04445	0.89994	328.76115	54.03615	0.06000
4	38.19622	60.00000	0.65580	2069.96813	54.19301	0.06000
5	38.19622	40.00000	0.65580	1001.96722	26.23210	0.06000
6	34.16276	40.43926	0.63080	871.45800	25.50900	0.00000
7	34.16276	40.51853	0.63080	873.16614	25.55900	0.00000
8	37.96028	130.00000	0.61374	3028.70099	79.78606	0.00000
9	28.71213	130.00000	0.53200	1985.73068	69.16000	0.00000
10	28.71213	130.09398	0.53200	1987.16629	69.21000	0.00000
11	28.11213	130.09398	0.53200	1945.64029	69.21000	0.00000
12	28.11213	60.00000	0.53200	897.33908	31.92000	0.00000
13	28.11213	24.10316	0.53200	360.47842	12.82288	0.00000
14	0.60000	130.09398	0.53200	41.52600	69.21000	0.00000
15	0.60000	152.87513	0.53200	48.79774	81.32957	0.00000
16	0.10718	207.96581	0.99893	22.26491	207.74316	0.00000
17	0.10718	208.01581	0.99893	22.27027	207.79311	0.00000
18	0.10718	140.09398	0.99893	14.99853	139.94400	0.00000
19	5.43428	130.00000	0.98256	694.13647	127.73295	0.00000
20	5.43428	130.05089	0.98256	694.40818	127.78295	0.00000
21	5.43428	169.00000	0.98256	902.37741	166.05283	0.00000
22	5.45325	207.96581	1.00000	1134.08863	207.96581	0.00000
23	5.45325	208.01581	1.00000	1134.36129	208.01581	0.00000
24	5.45325	169.87903	1.00000	926.39206	169.87903	0.00000
25	5.56042	169.30553	0.99998	941.39059	169.30204	0.00000
26	5.56042	90.00000	0.99998	500.42755	89.99814	0.00000
27	0.23488	169.76333	0.38000	69.85198	297.38954	0.00000
28	37.96028	129.81381	0.61374	3024.36312	79.67178	0.00000
29	0.63531	169.76333	0.38000	213.48499	336.03129	0.00000
30	62.14244	24.10316	0.53200	796.84503	12.82288	0.00000
31	0.15929	227.13491	0.36091	189.22725	1187.95654	0.00000
32	4.03346	40.43926	0.86759	141.51242	35.08460	0.06000
33	0.00000	338.08000	0.45000	0.00000	1221.80000	0.00000
34	1.03017	230.00000	0.44816	106.18650	103.07680	0.00000
35	3.02223	192.00000	0.39809	2161.01520	715.03946	0.06000
36	37.71489	130.00000	0.61639	3022.13587	80.13112	0.00000
37	0.34030	130.00000	0.42460	79.20719	232.75482	0.06000
38	0.24539	44.10316	0.20580	2.22725	9.07626	0.00000
39	0.09491	44.10316	0.43899	4.98707	52.54525	0.06000
40	0.18259	130.00000	0.96800	22.97680	125.84017	0.00000
41	4.00000	130.00000	0.83589	434.66358	108.66590	0.00000
42	4.00000	130.05982	0.83589	434.86358	108.71590	0.00000
43	4.00000	0.00000	0.83589	0.00000	0.00000	0.00000
44	0.01680	207.96581	0.62143	2.17116	129.23589	0.00000
45	29.90733	40.51853	0.63080	764.40148	25.55900	0.00000
46	29.90733	72.30629	0.63080	1364.09280	45.61066	0.00000
47	4.25543	40.51853	0.63080	108.76466	25.55900	0.00000
48	4.25543	178.74057	0.63080	479.79673	112.74919	0.00000
49	4.93228	75.00000	0.96800	369.92092	75.00000	0.00000
50	4.93228	110.00000	0.96800	542.55068	110.00000	0.00000
51	0.08089	94.10316	0.45716	4.96311	61.35950	0.06000
52	0.16856	239.38372	0.00000	195.58260	1160.29477	0.00000

PROJECT: WAUNA BLEND

FIGURE 8-2A

DATE RUN: 25-Jul-78, 04:19 PM

ACCOUNT NO: 78-BEST-3

STA	WT FLOW LB/LBISS	TEMP DEG F	SPHT BTU/LB	ENERGY BTU/LBISS	ENTHALPY BTU/LB	AMMONIA LB/LBISS
53	0.18259	239.38372	0.00000	195.60656	1071.30513	0.00000
54	0.00354	130.00000	0.58340	0.00207	75.84183	0.00000
55	0.25385	227.13491	0.00000	301.56320	1187.95654	0.00000
56	0.00444	227.13491	0.00000	5.28042	1187.95654	0.00000
57	0.00444	211.98169	0.00000	0.94255	212.04950	0.00000
58	46.62472	75.00000	0.00000	3498.26360	75.03024	0.00000
59	46.62472	110.00000	0.00000	5126.12528	109.94437	0.00000
60	34.16276	85.56412	0.63080	1843.88953	53.97367	0.00000
61	34.16276	106.02663	0.63080	2284.85257	66.88138	0.00000
62	1.87805	338.08000	0.00000	0.00000	1221.80000	0.00000
63	1.87805	211.98169	0.00000	0.00000	212.04950	0.00000
64	62.14244	42.54564	0.53200	1406.54933	22.63428	0.00000
65	0.01894	70.00000	0.24000	0.31816	16.80000	0.00000
66	34.03031	24.10316	0.53200	436.36661	12.82288	0.00000
67	34.03031	24.15015	0.53200	437.21737	12.84788	0.00000
68	34.03031	28.12675	0.53200	509.21024	14.96343	0.00000

STA	WATER LB/LBISS	OIL LB/LBISS	SOLIDS LB/LBISS	SOLVENT LB/LBISS	VOLUME CFM	PRESSURE PSIA
1	5.02410	0.01200	0.98800	0.00000	0.00000	0.00000
2	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
3	5.02410	0.01200	0.98800	0.00000	0.00000	0.00000
4	8.18312	0.57424	1.02800	28.35087	0.00000	0.00000
5	8.18312	0.57424	1.02800	28.35087	0.00000	0.00000
6	6.61006	0.57424	0.05140	26.92705	0.00000	0.00000
7	6.61006	0.57424	0.05140	26.92705	0.00000	0.00000
8	8.21707	0.57424	0.05140	29.11756	0.00000	0.00000
9	0.73437	0.57424	0.00000	27.40352	0.00000	0.00000
10	0.73437	0.57424	0.00000	27.40352	0.00000	0.00000
11	0.71902	0.56224	0.00000	26.83087	0.00000	0.00000
12	0.71902	0.56224	0.00000	26.83087	0.00000	0.00000
13	0.71902	0.56224	0.00000	26.83087	0.00000	0.00000
14	0.01535	0.01200	0.00000	0.57265	0.00000	0.00000
15	0.01535	0.01200	0.00000	0.57265	0.00000	0.00000
16	0.10688	0.00000	0.00000	0.00030	0.00000	14.98480
17	0.10688	0.00000	0.00000	0.00030	0.00000	0.00000
18	0.10688	0.00000	0.00000	0.00030	0.00000	0.00000
19	5.21127	0.00000	0.01140	0.21161	0.00000	0.00000
20	5.21127	0.00000	0.01140	0.21161	0.00000	0.00000
21	5.21127	0.00000	0.01140	0.21161	0.00000	0.00000
22	5.44185	0.00000	0.01140	0.00026	0.00000	14.98480
23	5.44185	0.00000	0.01140	0.00000	0.00000	0.00000
24	5.44185	0.00000	0.01140	0.00000	0.00000	0.00000
25	5.54872	0.00000	0.01140	0.00030	0.00000	0.00000
26	5.54872	0.00000	0.01140	0.00030	0.00000	0.00000
27	0.02328	0.00000	0.00000	0.21161	0.00000	14.84040
28	8.21707	0.57424	0.05140	29.11756	0.00000	0.00000
29	0.06296	0.00000	0.00000	0.57235	0.00000	14.84040
30	1.58941	1.24285	0.00000	59.31018	0.00000	0.00000
31	0.15929	0.00000	0.00000	0.00000	0.00000	19.69600

PROJECT: WAUNA BLEND

FIGURE 8-2B

DATE RUN: 25-Jul-78, 04:19 PM

ACCOUNT NO: 78-BEST-3

STA	WATER LB/LBISS	OIL LB/LBISS	SOLIDS LB/LBISS	SOLVENT LB/LBISS	VOLUME CFM	PRESSURE PSIA
32	1.57305	0.00000	0.97660	1.42381	0.00000	0.00000
33	0.00000	0.00000	0.00000	0.00000	0.00000	115.00000
34	0.05151	0.00000	0.97660	0.00206	0.00000	0.00000
35	1.52154	0.00000	0.00000	1.42175	0.00000	0.00000
36	8.19671	0.57424	0.05140	28.89253	0.00000	0.00000
37	0.02113	0.00000	0.00000	0.24024	3.33453	14.55160
38	0.02036	0.00000	0.00000	0.22503	0.00000	0.00000
39	0.00077	0.00000	0.00000	0.01520	1.59604	14.62380
40	0.16856	0.00000	0.00000	0.01402	0.00000	0.00000
41	2.44000	0.00000	0.04000	1.52000	0.00000	0.00000
42	2.44000	0.00000	0.04000	1.52000	0.00000	0.00000
43	2.44000	0.00000	0.04000	1.52000	0.00000	0.00000
44	0.00480	0.01200	0.00000	0.00000	0.00000	14.98480
45	5.78669	0.50271	0.04500	23.57292	0.00000	0.00000
46	5.78669	0.50271	0.04500	23.57292	0.00000	0.00000
47	0.82337	0.07153	0.00640	3.35413	0.00000	0.00000
48	0.82337	0.07153	0.00640	3.35413	0.00000	0.00000
49	4.93228	0.00000	0.00000	0.00000	0.00000	0.00000
50	4.93228	0.00000	0.00000	0.00000	0.00000	0.00000
51	0.00077	0.00000	0.00000	0.00118	1.68563	14.73210
52	0.16856	0.00000	0.00000	0.00000	0.00000	24.69600
53	0.16856	0.00000	0.00000	0.01402	0.00000	0.00000
54	0.00000	0.00000	0.00000	0.00354	0.00000	0.00000
55	0.25385	0.00000	0.00000	0.00000	0.00000	19.69600
56	0.00444	0.00000	0.00000	0.00000	0.00000	19.69600
57	0.00444	0.00000	0.00000	0.00000	0.00000	0.00000
58	46.62472	0.00000	0.00000	0.00000	0.00000	0.00000
59	46.62472	0.00000	0.00000	0.00000	0.00000	0.00000
60	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
61	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
62	1.87805	0.00000	0.00000	0.00000	0.00000	0.00000
63	1.87805	0.00000	0.00000	0.00000	0.00000	0.00000
64	1.58941	1.24285	0.00000	59.31018	0.00000	0.00000
65	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
66	0.87039	0.68061	0.00000	32.47932	0.00000	0.00000
67	0.87039	0.68061	0.00000	32.47932	0.00000	0.00000
68	0.87039	0.68061	0.00000	32.47932	0.00000	0.00000

Q(1) =	0.000	Q(12) =	0.272	Q(23) =	207.969
Q(2) =	0.243	Q(13) =	0.273	Q(24) =	434.864
Q(3) =	1380.729	Q(14) =	248.771	Q(25) =	0.200
Q(4) =	1068.001	Q(15) =	440.963	Q(26) =	172.630
Q(5) =	11.003	Q(16) =	286.453	Q(27) =	0.000
Q(6) =	1.708	Q(17) =	0.000	Q(28) =	0.000
Q(7) =	4.338	Q(18) =	1896.358	Q(29) =	0.000
Q(8) =	1.436	Q(19) =	1627.862	Q(30) =	0.000
Q(9) =	1048.301	Q(20) =	71.993	Q(31) =	0.000
Q(10) =	609.704	Q(21) =	7.272	Q(32) =	0.000
Q(11) =	0.005	Q(22) =	156.136	Q(33) =	0.000

PROJECT: WAUNA BLEND

FIGURE 8-2C

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ACCOUNT NO: 78-BEST-3

C(1) =	0.166000	C(46) =	0.000000	C(91) =	0.785000
C(2) =	60.000000	C(47) =	0.400000	C(92) =	20.000000
C(3) =	996.431362	C(48) =	0.990000	C(93) =	230.000000
C(4) =	6.000000	C(49) =	0.000000	C(94) =	115.000000
C(5) =	20.000000	C(50) =	0.000000	C(95) =	338.080000
C(6) =	4.000000	C(51) =	0.050000	C(96) =	1221.800000
C(7) =	0.532000	C(52) =	8.000000	C(97) =	-0.990000
C(8) =	130.000000	C(53) =	0.177000	C(98) =	192.000000
C(9) =	0.012000	C(54) =	169.000000	C(99) =	0.002800
C(10) =	0.865000	C(55) =	90.000000	C(100) =	10.000000
C(11) =	0.257000	C(56) =	0.610000	C(101) =	0.065000
C(12) =	0.353000	C(57) =	0.380000	C(102) =	50.000000
C(13) =	0.000000	C(58) =	0.010000	C(103) =	1.110000
C(14) =	0.000000	C(59) =	0.450000	C(104) =	0.000000
C(15) =	0.450000	C(60) =	0.000000	C(105) =	0.000000
C(16) =	0.900000	C(61) =	4.000000	C(106) =	0.000000
C(17) =	60.000000	C(62) =	5.000000	C(107) =	0.000000
C(18) =	40.000000	C(63) =	20.000000	C(108) =	10.000000
C(19) =	130.000000	C(64) =	0.002000	C(109) =	0.735000
C(20) =	0.000000	C(65) =	5.000000	C(110) =	1.055000
C(21) =	0.020000	C(66) =	0.000000	C(111) =	0.000000
C(22) =	0.420000	C(67) =	0.000000	C(112) =	1.600000
C(23) =	0.470000	C(68) =	0.000000	C(113) =	0.950000
C(24) =	0.000000	C(69) =	4.499878	C(114) =	0.000000
C(25) =	0.000050	C(70) =	2.225203	C(115) =	0.000000
C(26) =	0.000000	C(71) =	7.826519	C(116) =	0.000300
C(27) =	15.000000	C(72) =	970.660351	C(117) =	2.000000
C(28) =	0.060000	C(73) =	75.000000	C(118) =	0.000000
C(29) =	101.200000	C(74) =	0.025577	C(119) =	0.000000
C(30) =	18.000000	C(75) =	0.000000	C(120) =	0.000000
C(31) =	16.000000	C(76) =	0.950000	C(121) =	0.000000
C(32) =	-4.000000	C(77) =	4.000000	C(122) =	0.000000
C(33) =	14.696000	C(78) =	4.000000	C(123) =	0.000000
C(34) =	-2.000000	C(79) =	20.000000	C(124) =	0.000000
C(35) =	0.520000	C(80) =	20.000000	C(125) =	0.078938
C(36) =	60.000000	C(81) =	0.240000	C(126) =	19.118805
C(37) =	0.568000	C(82) =	93785.000000	C(127) =	0.452826
C(38) =	169.000000	C(83) =	7.700000	C(128) =	100.000000
C(39) =	70.000000	C(84) =	0.012500	C(129) =	0.100000
C(40) =	0.380000	C(85) =	0.850000	C(130) =	1.000000
C(41) =	0.050000	C(86) =	0.830000	C(131) =	0.000000
C(42) =	8.000000	C(87) =	2.200000	C(132) =	0.000000
C(43) =	10.000000	C(88) =	0.175000	C(133) =	0.000000
C(44) =	0.000000	C(89) =	1000.000000	C(134) =	0.000000
C(45) =	0.000000	C(90) =	0.680000	C(135) =	0.000000

CALCULATION TIME WAS 363 SECONDS (CPU TIME WAS 17.3 SECONDS)

PROJECT: WAUNA BLEND

FIGURE 8-2D

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ACCOUNT NO: 78-BEST-3

PROCESS DATA FOR 49.9998 TON/DAY TOTAL SOLIDS PLANT

THERMAL (STEAM AT $2.59/1000 LB)		$/TON	BTU/HR	ELECTRICAL (AT $0.0125/KWH)		$/TON	KW
WATER STILL	=	1.24	972125	REFRIGERATION	=	5.24	874.131
SOLVENT STILL	=	0.78	610135	PUMPS AND AUX EQUIP	=	3.62	603.584
DRYER	=	9.43	7410414				
TRIM HEATER	=	0.02	16951				
CARBON ABS.	=	0.82	674587				
TOTAL		12.28		TOTAL		8.87	1477.715

THERMAL TOTAL COST AT EFFICIENCY = 83% IS $14.80

STEAM REQUIRED = 9629.34 LB/HR

CHEMICAL DATA

	$/TON	LB/DAY
TEA LOSS	5.21	311
ANTIFOAM	6.93	478
PRETREAT CHEMICALS	0.00	0
TOTAL CHEMICAL COST	12.14	

PROCESS COST/TON TOTAL SOLIDS = $35.81

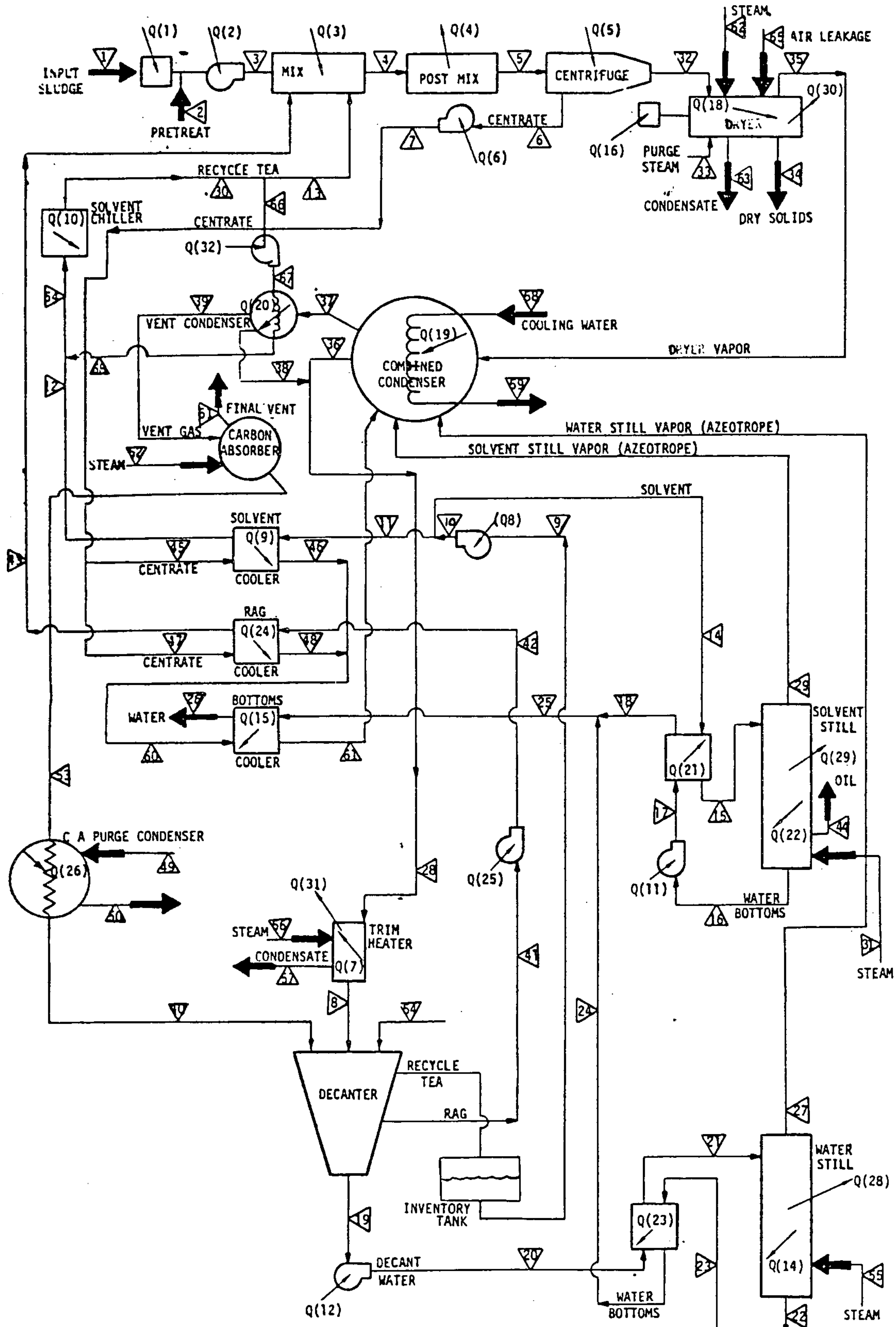

FIGURE 8-3 BEST PROCESS FLOW
MINI BEST WAUNA CONFIGURATION

3/17/78

9.0 COMPARISON OF B.E.S.T. WITH CONVENTIONAL DRYING

9.1 B.E.S.T. SYSTEM MODEL

A proposed system using B.E.S.T. pulp mill sludge drying is shown in Figure 9-1. Input to the B.E.S.T. Process consists of 18% solids cake from a coil filter operating at 98% solids recovery. The solids are B.E.S.T. dried to 90% solids at 96% solids recovery. The dried solids are incinerated by a stoker-burner-boiler system designed for these solids which is 85% efficient at rated load. A portion of the steam generated supplies the thermal energy to the B.E.S.T. process. The balance of the steam is credited against the electrical requirements of the system at a conversion of 10,500 Btu/kWh.

9.2 CONVENTIONAL DRYING MODEL

A typical current practice system for disposing of pulp mill sludge is shown in Figure 9-2.

The coil filter input is lime pretreated to give a 20% solids output with 98% solids recovery.

The filter cake is further dried in a V-press to 30% solids at 98% solids recovery. The 30% sludge cake is mixed with wood chips and burned in a hog fuel incinerator/boiler system.

The evaporation of the water in the sludge cake is shown to be at a thermal cost of 3000 Btu/lb (see Appendix M) and the heat evolved from the sludge solids is assumed converted to steam at 75% efficiency. Electrical power is converted to Btu's at 10,500 Btu's/kWh.

9.3 HEAT BALANCE RESULTS

The energy requirements for the two systems described above are shown in the Figure 9-3, 50 TPD heat balance. With electrical energy converted to heat energy at an efficiency of 32.5%, the B.E.S.T. model required 5.6 x 10^6 Btu/hr, compared to 10.5 x 10^6 Btu/hr for the conventional model. A projected net energy savings of 4.9 x 10^6 Btu/hr results if the B.E.S.T. drying process is used. This is equivalent to 8000 barrels of fuel oil yearly, with a minimum value of $100,000.

9.4 NET B.E.S.T. OPERATING COSTS

The process cost of \$35.81/ton for B.E.S.T. drying, derived in Section 8.0, can be credited with the steam surplus shown in step (7) of Figure 9-3.

With the steam assessed at $\$2.59/10^6$ Btu, the credit amounts to 9.3×10^6 Btu/hr x 24/50 x $\$2.59/10^6$ Btu = \$11.56/ton.

B.E.S.T. Net Processing Cost \$/Ton

	\$/Ton
Chemical	12.14
Electrical	8.87
Steam Credit	- 11.56
Net Cost	\$ 9.45

Capital costs for a 50 TPD system were estimated as 3.7 million (turnkey plant). Amortization costs (15 years at 8%) calculated as \$23.72/ton.

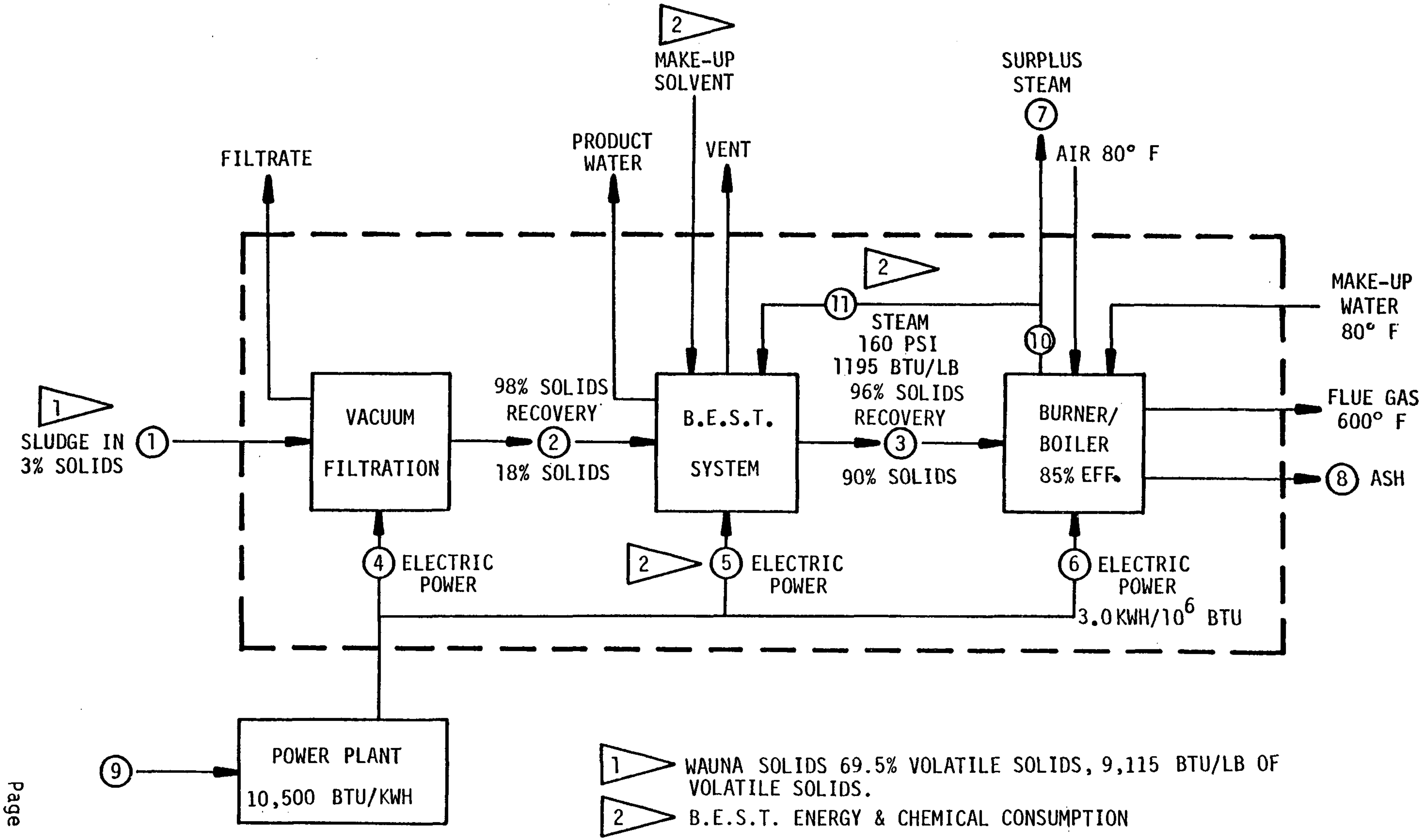

FIGURE 9-1 MODEL (A) B.E.S.T. SYSTEM SCHEMATIC

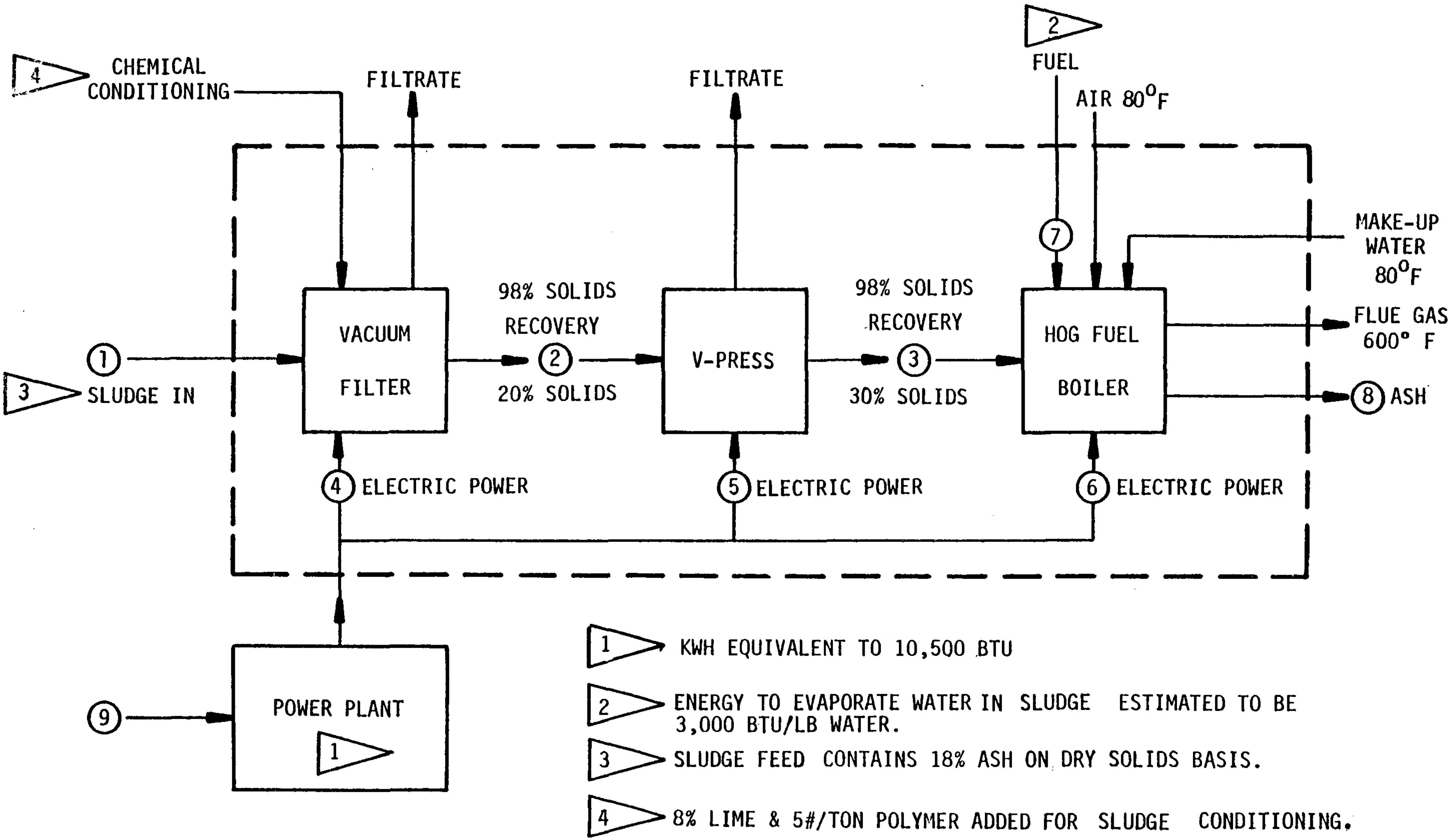

FIGURE 9-2 MODEL (B) CONVENTIONAL SYSTEM SCHEMATIC

FIGURE: 9-3 SLUDGE DEWATERING/DRYING/DISPOSAL HEAT BALANCE

(100,000 lbs. SOLIDS/DAY)

	STREAM (1) SLUDGE				STREAM (2) SLUDGE				STREAM (3) DRY SOLIDS				(4)	(5)	(6)	(7)	(8)	(9)	[5] (10)	(11)	NET HEAT BALANCE
	WATER lb/Hr	VOLATILE SOLIDS lb/Hr	ASH lb/Hr	TOTAL lb/Hr	WATER lb/Hr	VOLATILE SOLIDS lb/Hr	ASH lb/Hr	TOTAL lb/Hr	WATER lb/Hr	VOLATILE SOLIDS lb/Hr	ASH lb/Hr	TOTAL lb/Hr	ELECT. POWER KWH	ELECT. POWER KWH	ELECT. POWER KWH	NET(10-11) STEAM/FUEL 10^6 BTU/Hr	ASH lb/Hr	ELECTRICAL AS FUEL 10^6 BTU/Hr	STEAM PRODUCED 10^6 BTU/Hr	STEAM USED 10^6 BTU/Hr	(7)-(9) 10^6 BTU/Hr
[4] MODEL (A) B.E.S.T. SYSTEM TO BOILER	134,733	2896	750	138,367	18,985	2838	735	22,558	340	2724	706	3770	28	1478	64	9.3 / [3] 10.9	706	-16.5	21.2	11.9	-5.6
MODEL (B) V-PRESS WET TO BOILER	134,733	2896	750	138,367	16,332	2838	[2] 1062	20,234	9338	2781	1043	13,162	30	33	84	-9	1043	-1.5	19.0	[1] 28.0	-10.5

NOTE: BASED ON THESE CALCULATIONS THE B.E.S.T. SYSTEM SAVES 4.9 x 10^6 BTU/Hr

[1] FUEL REQUIRED TO INCINERATE 30% SOLIDS SLUDGE

[2] 8% ADDITION OF LIME FOR SLUDGE CONDITIONING

[3] FUEL EQUIVALENT AT 85% EFFICIENCY

[4] B.E.S.T. TEST DATA:
STREAM (5) .355 KWH/lb SOLIDS
STREAM (11) 2857 BTU/lb SOLIDS

[5] 92% SOLIDS COMBUSTED @ 85% EFFICIENCY
30% SOLIDS COMBUSTED @ 75% EFFICIENCY

EVENTS LOG SYNOPSIS

DATE	TIME	COMMENT
2-21	1137-1217	Feed Off, plugged valve at post mix HX.
	1222-1247	" , centrate tank overfilled.
	1430	Two barrels of TEA added to Tank #1
2-22	0640	Water still pressure 2.8 psig, antifoam on Max.
	0720	Still pressure at 0, antifoam set at 3.
	1053-1215	Feed off, emulsion in decanter.
	1330-1450	Feed off, recycle solvent pump seal leak.
	1500	42 barrels of sludge cake collected.
	2015-2125	Feed off, Moyno feed pump suction plugged.
	2130-2205	Feed off, rupture disc blew.
	2205-2230	Feed off, rupture disc blew, Moyno plug.
	2400-0235	Feed off, clean 2nd stage condenser, centrifuge and dryer inlet.
2-23	0400-0531	Feed off, water still plugged.
	0920-0935	Feed off, centrifuge feed tube blocked.
	1140-1218	Feed off, pressure alarm, plug not found.
	1225	Added 1 barrel solvent.
	1235-1255	Feed off, chip in mix port tee.
	1450-1520	Feed off, rupture disc blew, Moyno discharge.
	1540	Dryer conveyor plugged with solids.
	1750-1900	Feed off, dryer inlet plugged.
	1945-2030	Feed off, rupture inlet plugged.
	2039-2125	Feed off, rupture disc blew, Moyno discharge.
	2235-2255	Feed off, rupture disc blew, Moyno discharge.

EVENTS LOG SYNOPSIS

DATE	TIME	COMMENT
2-24	0100-0202	Feed off, rupture disc blew.
	0235-0315	Feed off, feed plug
	1430-0510	Feed off to clean dryer and centrifuge.
	0520-0715	Feed off, feed plug.
	0740-1550	Moyno feed pump cleaned, #1 condenser cleaned, dryer conveyor removed.
	1600-1646	Feed off, dryer inlet plugged.
2-25	0725-1200	Feed off, dryer inlet plugged, centrifuge bowl jammed with solids.
	1250	1 barrel TEA added.
	1500-1735	Feed off, centrifuge plugged.
	1800-0615	Feed off, dryer jammed by tubing, cleared centrifuge centrate drains of fiber.
2-26	1007-1430	Feed off, decanter inlet plugged.
	2215	Feed off, water still plugged.
2-27	0045-0320	Centrifuge feed tube flange weld failed.
	0600-1500	Feed off, to clean water still, repair feed tube.
	2015	Water still pressurized and blew over.
	2230-0140	Feed off, to clean water still.
2-28	1100	Feed off, feed pump PVC flange cracked.
	1300	Collected 26 barrels of feed.
	1720	Feed ON
	1955-2030	Feed off, feed pump suction plug.

A

EVENTS LOG SYNOPSIS

DATE	TIME	COMMENT
3-1	0530	Feed off, inlet to centrifuge is plugged, dryer inlet bridged, 1st stage condenser 80% plugged, water still flushed.
	0900	1 barrel TEA added.
	1647	Feed ON
3-2	0240	Cleared dryer bridge
	0500-0750	Feed off, plant air failed, recirc. loop plug.
	0915-1000	Feed off, mix tee plug.
3-3	0021	Decanter emulsified temp. up 10^{o} in 1 hour.
	0300	Emulsion gone.
	0530-0610	Feed off, dryer inlet plug.
	1255	Feed off, water still vent plugged.
3-6	0515	Water feed in, base line data
	1300	Sludge feed in.
	1950	22 barrels of feed collected. Sweco screen overflowed, flooding office below.
3-7	0440-1005	Feed off, centrifuge and dryer inlet cleaned.
	1620-1704	Feed off, emulsion in decanter caused by C/A regeneration steam condensate.
3-8	0850	Added 50 gallons TEA
	1540	Feed off, over to water feed.
	1720-1830	Water feed off, centrifuge plug.
	2050	System shutdown.

EVENTS LOG SYNOPSIS

DATE	TIME	COMMENT
3-9	1330	Loaded 49 barrels of feed.
	1615	Feed in.
3-10	0630	Increase in negative pressure in #1 condenser sump results in hotter solids and cooler dryer vapor.
	1000-1039	Feed off, plug in feed system.
	1900	Decanter emulsion formed during last 10 minutes of C/A regeneration.
3-11	0100-0310	Feed off to flush water still
	0910	Feed off to clean 1st Stage condenser.
	1000	1 barrel TEA added.
	1553	Feed ON.
3-12	0950	Dryer inlet bridge cleared.
	1018	Water still pressure = 3.6 psig.; antifoam increased to setting of 5
	1110	Water still P = 4.3 psig, control T lowered.
	1400	Plant clarifier water being used for cooling.
	1515	Still P = 0
	2200	Still blew over P = 6 psig.
3-13	0935-1045	Feed off to flush water still.
	1205-1233	Feed off, post mix chiller plug.
	1300	On water, no sludge left.
	1630	Water feed off. System cleaned.

EVENTS LOG SYNOPSIS

DATE	TIME	COMMENT
3-14	1255	Water Feed in.
	1300	45 barrels of feed collected.
	1720	Sludge feed in.
	2300	Emulsion in decanter from C/A regeneration.
	2320	Solids capture has been good with backdrive setting of 6.
3-15	0650	Capture poor, backdrive of centrifuge set at one.
	0900-0940	Feed off, centrifuge feed tube plugged.
	1015	Dryer inlet cleaned.
	1610	Product water pump failed.
	1800	Water still P=3 psig
	2315	Water still feed valve plugged
3-16	0330-1425	Feed off, still feed valve cleaned.
	0545-0820	Feed off to flush water still.
	0900	Decanter sidearm plugged.
3-17	1225-1250	Feed off, plug at outlet of post mix HX.
	1340	Feed off - sludge gone.
	1445	System shutdown-test completed.

B.E.S.T. PILOT PLANT LOG SHEET

SHT. ____ OF ____

OPERATING DATA	DATE	3/12					
	TIME	0400	0600	0800	1000	1200	1400
RAG DRAW PUMP - DIAL		—	—	—	—	—	—
SLUDGE FEED PRESS - PSIG		25	25	35	20	25	20
SLUDGE FEED PUMP - DIAL		2.2	2.2	2.2	2.2	2.2	2.2
POST MIX HX FEED P IN/OUT PSI		—/—	—/—	—/—	—/—	—/—	—/—
SOLV. COOL SOLV. TEMP OUT-°F		56	54	59	56	60	60
W.C.DECANT H_2O T IN/OUT, °F		100/190	100/184	102/187	107/183	109/204	105/190
W.C.DECANT H_2O P IN/OUT, PSI		—/30	—/28	—/29	—/28	—/28	—/29
W.C. BOTTOMS T IN, °F		204	204	209	202	216	206
W.C.BOTTOMS P IN/OUT,PSI		5/4	5/4	5.5/4.5	5.5/5.5	9/7	6/5.5
WATER STILL STEAM-ROTO		—	—	—	—	—	—
TRIM HTR. CENT T IN/OUT, °F		—/—	—/—	—/—	—/—	—/—	—/—
TRIM HTR. CENT. FLOW - GPM		—	—	—	—	—	—
RAG HX RAG TEMP OUT -°F		—	—	—	—	—	—
RAG HX COOLANT FLOW - GPM		—	—	—	—	—	—
RAG HX COOLANT T IN/OUT, °F		—/—	—/—	—/—	—/—	—/—	—/—
SOLV.COOL. CENT. FLOW - ROTO		2.	2.	2.	2.	2.	2.
SOLV.COOL.CENT. T IN/OUT, °F		50/74	48/74	48/76	48/76	48/77	48/79
SOLV.COOL SOLV. P IN/OUT,PSI		—/—	—/—	—/—	—/—	—/—	—/—
SOLV.COOL CENT. P IN/OUT,PSI		4/4	4/4	4/4	4/4	4/4	4/4
SOLVENT STILL FEED - ROTO		.6	.6	.3	.75	.65	.75
SIDEARM DECANT IN		112	112	110	113	113	122

OPERATING DATA	DATE	3/12					
	TIME	0400	0600	0800	1000	1200	1400
2ND STG.COOL. SOLV. - ROTO		75	75	76	79	74	75
2ND STG. GAS PR. IN. H_2O		-.3	-.3	-.3	-.4	-.3	-.2
1ST STG.GAS PR.-IN. H_2O		-.7	-.6	-.8	-1.0	-.9	-.9
1ST STG.SPRAY NOZ.FLOW - ROTO		—	—	—	—	—	—
1ST STG. COOL WATER TEMP OUT		40/70	40/68	40/62	40/64	40/64	
1ST STG.COOL.WATER FLOW, GPM		1.5	1.5	1.1	1.1	1.0	5.7
DRYER STEAM PRESS. #1/#2, PSI		141/136	138/136	135/135	191/136	142/142	190/133
DRYER STEAM PRESS. #3/#4,PSI		142/142	140/138	138/137	142/140	142/140	142/140
RECYCLE SOLV.FLOW, ROTO		1.8	1.8	2.0	2.2	2.0	2.0
SOLV/SLUDGE MIX FLOW, M/GPM		32	32	31	31	31	/
WATER STILL FEED FLOW, M/GPM		28	29	26	32	28	/
PRODUCT WATER FLOW, M/GPM		18	38	27	10	14	/
THIN SLUDGE FEED FLOW,M/GPM		—/—	—/—	—/—	—/—	—/—	—/—
RAG DRAW FLOW, M/GPM		—/—	—/—	—/—	—/—	—/—	—/—
PRETREAT FLOW, M/GPM		—/—	—/—	—/—	—/—	—/—	—/—
ELECT., PHASE 1 - AMP/VOLT		62/460	62/460	62/460	61/455	64/455	62/455
ELECT., PHASE 2 - AMP/VOLT		61/460	60/460	62/	59/	63/	61/
ELECT., PHASE 3 - AMP/VOLT		60/460	58/460	58/	55/	58/	57/
O_2 MONITOR - READING							
VENT TEA - READING/%		.089	.140	.128	.075	.075	.063
VENT AMMONIA - READING/%		/	/	/	/	/	/
VENT METHANE - READING/%		/	/	/	/	/	/
WATER STILL SUMP LEVEL		50	55	50	53	54	50
RAG LEVEL		68	69	67	67	73	72
CENTRATE TANK LEVEL		45	44	91	92	92	91

RECORDER TEMPS	DATE	3/12					
	TIME	0400	0600	0800	1000	1200	1400
#1 C/A UNIT CARBON		57	55	59	59	54	55
#2 THIN SLUDGE FEED		—	—	—	—	—	—
#3 CHILLED SLUDGE FEED		—	—	—	—	—	—
#4 RECYCLE SOLVENT		108	108	108	106	112	121
#5 COOLED SOLVENT		44	45	45	46	46	48
#6 CHILLED SOLVENT		20	20	21	22	22	25
#7 MIX		—	—	—	—	—	—
#8 POST MIX FEED		57	56	55	58	59	60
#9 CENTRIFUGE FEED		38	37	37	41	43	43
#10 WATER TO STILL		180	176	178	158	187	176
#11 WATER STILL, TOP		166	164	166	166	168	168
#12 WATER STILL, MID-TOP		—	—	—	—	—	—
#13 WATER STILL, MID		—	—	—	—	—	—
#14 WATER STILL, MID-BOTTOM		—	—	—	—	—	—
#15 WATER STILL, BOTTOM		208	208	209	216	218	211
#16 PRODUCT WATER DISCHARGE		114	117	119	112	122	115
#17 SOLVENT STILL, TOP		—	—	—	—	—	—
#18 SOLVENT STILL, BOTTOM		—	—	—	—	—	—
#19 DRYER SOLIDS		259	256	257	257	240	255
#20 DRYER GAS		192	190	189	190	189	192
#21 1ST STAGE GAS OUT		—	—	—	—	—	—
#22 2ND STAGE GAS OUT		—	—	—	—	—	—
#23 1ST STAGE SUMP		—	—	—	—	—	—
#24 WATER STILL SUMP		—	—	—	—	~~223~~	216

MATERIAL BALANCE DATA*	DATE	3/12					
	TIME	0400	0600	0800	1000	1200	1400
PRETREAT TANK LEVEL, GAL		—	—	—	—	—	—
ANTI-FOAM TANK LEVEL, LTR		8.0	7.8	7.3	7.0	6.3	5.5
WATER STILL STEAM, COUNT		7756	8772	9771	0652	1591	266[illegible]
REFRG COOLING H_2O TOTAL		1354	1953	2540	3054	3637	499[illegible]
DRYER STEAM TOTAL, COUNT		4092	4748	5389	6031	6827	7385
VENT GAS TOTAL, CF		8702	9205	9721	0202	0733	125[illegible]
SOLV. STILL STEAM, COUNT		6686	6967	7266	7606	7931	8240
BOILER WATER TOTAL, GAL		6621	6663	6709	6740	6775	6815
BOILER H_2O MAKEUP TOTAL		692.1	690.3	707.7	713.7	718.8	72[illegible]
SOLVENT STILL WATER, LB		—	—	—	—	—	—
SOLVENT STILL OIL, LB		—	—	—	—	—	—
DRIED CAKE TOTAL, LB		—	—	—	—	—	—
FUEL TANK LEVEL, IN		28¾	28½	28	27½	27	26½
SLUDGE TANK LEVEL,		—	—	—	—	—	—
KWH METER TOTAL, KWH		3035	3069	3106	3137	3168	3201
SLUDGE MIX TOTAL, GAL		3954	7617	1556	4953	8631	2191
PRODUCT WATER TOTAL, GAL		2883	3471	4136	4698	6152	573[illegible]
RECYC SOLV COUNT		8882	9115	9401	9650	9927	018[illegible]

SLUDGE TYPE/
FEED RATE/PRETREAT ____________

TIME SLUDGE IN ____________

NR = NO READING

TABLE 1. DATA REDUCTION: INPUTS/OUTPUTS

		T	W	T	F	S	S	M	T	W	T	F	
DATE		2/21	2/22	2/23	2/24	2/25	2/26	2/27	2/28	3/1	3/2	3/3	Avg
HOURS ON	Sludge 23	23.3	17.9	15.3	11.0	10.9	11.6	10.2	16.0	12.7	20.4	10.1	350 15.2
	Water												
SLUDGE INPUT ◁1 21 days	~~Gal~~ or lb	5884	3760	3159	1814	2573	3464	1922	3890	4002	5440	2592	4034
	lb/hr GPH	253 29.4	210 24.3	206 26.8	165 24.5	236 35.9	298 44.3	188 28.0	254 33.0	315 36.4	267 31.8	240 27.7	265 32
PRODUCT WATER ◁2	Gal hrs	630.2	156.9 7	☆	75.1 4	150.8 6.5	184.0 6	87.6 4	200.8 8	103 5.5	372.5 13	154.9 5.7	
	lb/hr X1.07 GPH	240 28.9	200 24		167 20	207 24.8	273 32.8	195 23.4	224 26.8	165 18.8	255 30.6	242 29	246
DRY SOLIDS	lb 9 1.11	971	727	528	344	418	370	370	689	476	871	452	11363 712
	lb/hr W.B.	41.7	40.6	34.5	31.3	38.3	31.9	36.3	43.1	37.5	42.7	44.8	46.8
SOLVENT STILL OIL	lb Dry solids D.B. fraction	36.4 .874	38.7 .956	33.6 .975	28.4 .99	37.0 .966	29.7 .932	35.8 .987	41.2 .956	35.9 .957	41.76 .978	44.0 .982	41.9
	lb/hr												.52
SOLVENT STILL WATER	lb												
	lb/hr												12.6
VENT GAS	ft^3 hrs	8896	2446 7	☆	1572 4	3390 6.5	2301 6	1520 4	2861 8	2140 5.5	5603 13	2011 5.7	
	ft^3/hr	382	349		143	522	384	380	358	389	431	353	333
NET BOILER WATER MAKEUP ◁3	~~lb~~ GAL hrs	99.6 20	3.0 7	☆	NR	72 6.5	NR 6	200 4	29.7 8	19 5.5	19.2 13	NR 5.7	
	lb/hr	41	3.6			9.2		42	31	3.5	12		
PRETREAT ◁4	gal												
	GAL/HR												
	gm/l Sludge												
ANTIFOAM ◁5	ml	4000	5480	5860	460	2500	4100	0	2000	0	4640	1800	
	ml/hr	171	306	383	42	230	353	0	125	0	227	178	187
	mg/l	1562	2200	☆	2204	2448	2841	0	1231	0	1958	1620	162[illegible]

☆ NO CONTINUOUS OPERATION

8.655.6 with 30-50% secondary

6.745.6 primary.

½ - 7.75.6.

7:00
Lehler —

TABLE 1. DATA REDUCTION: INPUTS/OUTPUTS

8.65

			M	T	W	T	F	S	S	M	T	W	T	
DATE			3/6	3/7	3/8	3/9	3/10	3/11	3/12	3/13	3/14	3/15	3/16	3/17
HOURS ON	Sludge		11.0	17.8	15.7	8.0	23.4	15.1	24.0	11.3	7.7	23.3	20.5	13.1
	Water	5												
SLUDGE INPUT ◁1	~~Gal~~ or lb	1749	3015	5092	4376	2180	6654	4362	7066	3045	2352	5875	6435	3829
	lb/hr GH	350 \ 42	274 \ 31.7	286 \ 33.1	279 \ 32.2	273 \ 31.5	284 \ 32.9	289 \ 33.4	294 \ 34	269 \ 31	305 \ 39	252 \ 32.3	314 \ 40.2	294 \ [illegible]
PRODUCT WATER ◁2	Gal	?	237.8 \ 8	128.2 \ 4.5	340 \ 12	181.6 \ 6	300.5 \ 10	118.8 \ 4	622.5 \ 20	124.0 \ 4	171.0 \ 6.3	552.5 \ 20	337.9 \ 11	294.
	lb/hr X 1.07 GH		265 \ 31.8	254 \ 30.5	253 \ 30.3	270 \ 32.4	267 \ 32.1	265 \ 31.8	277 \ 33.3	276 \ 38.2	242 \ 29	246 \ 29.6	274 \ 32.9	269 \ 32.2
DRY SOLIDS	lb	—	596	952	778	352	971	722	1038	470	350	1392	1515	1017
	lb/hr	—	54.2	53.5	49.6	44.0	41.5	47.8	43.3	41.6	45.5	59.7	73.9	77.6
SOLVENT STILL OIL	lb D.B. (drysolids)		37.2 \ .687	38.1 \ .713	36.0 \ .725	38.7 \ .679	39.0 \ .958	42.7 \ .934	41.8 \ .966	40.4 \ .974	42.2 \ .927	55.5 \ .929	60.1 \ .613	68.3 \ [illegible]
	lb/hr													
SOLVENT STILL WATER	lb													
	lb/hr													
VENT GAS	ft^3		2123 \ 8	1512 \ 4.5	3779 \ 12	2158 \ 6	3467 \ 10	1093 \ 4	5094 \ 20	916 \ 4	2323 \ 6.3	6298 \ 20	3455 \ 11	2526 \ 10
	ft^3/hr		265	336	315	360	347	273	255	229	369	315	314	255
NET BOILER WATER MAKEUP ◁3	~~lb~~ GAL	49.5	15.6 \ 8	49.1 \ 4.5	20.4 \ 12	12.9 \ 6	43.2 \ 10	12.7 \ 4	42.1 \ 20	16.7 \ 4	18.0 \ 6.3	70.9 \ 20	19.8 \ 11	14.[illegible]
	lb/hr	82	16	91	14	18	36	26	17.5	35	24	30	15	12
PRETREAT ◁4	gal								21.7					
	GAL/HR													
	gm/l Sludge													
ANTIFOAM ◁5	ml		1530	2620	2540	1400	1640	3310	4750	4000	2100	3300	3200	2000
	ml/hr		139	147	162	175	70	219	198	519	273	142	156	143
	mg/l		1154	1272	1411	1425	576	1818	1570	4205	2532	1266	1252	1172

TABLE 2. DATA REDUCTION: THERMAL & ELECTRICAL LOADS

DATE		H_2O										
HOURS ON	Sludge											
	Water	5										
ELECTRICAL	kWh	29.1										
	KW kWh/~~lb D.S.~~ hr	5.8										
FUEL OIL ◁6	gal	12.25										
	lb/hr	17.25										
as steam	Fuel Oil Btu / ~~lb D.S.~~ #/hr	319										
METERED BOILER STEAM ◁7	gal	87										
	lb/hr	28.6										
	Btu/lb D.S.											
NET PROCESS STEAM ◁8	Btu/lb D.S.											
METERED DRYER STEAM ◁9	Counts	956										
	Counts/hr	191										
	lb/hr	57										
	Btu/lb D.S.											
NET DRYER STEAM ◁10	Btu/lb D.S.											
	Theor. Dryer Stm Btu/lb D.S.											
THERMAL EFFICIENCIES ◁11	Overall Dryer											
	Overall Process											
	Boiler											
COOLING WATER ◁12	Gal											
	lb/hr											
Water still steam, counts/1.05		187										
Solvent still steam, counts/1.??		212										
WATER STILL STEAM, lb/hr		37										
SOLVENT STILL STEAM, lb/hr		42										

TABLE 2. DATA REDUCTION: THERMAL & ELECTRICAL LOADS

350 hr avgs

DATE		avg #D.S.	avg #1KPYT								
HOURS ON	Sludge										
	Water										
ELECTRICAL	kWh										
	kWh/lb D.S.	.76									
FUEL OIL ◁6	gal										
	lb/hr	16.3									
	Fuel Oil Btu / lb D.S.	7189									
METERED BOILER STEAM ◁7 −52	gal										
	lb/hr	171									
	Btu/lb D.S. 1667b	4445	4178								
NET PROCESS STEAM ◁8	Btu/lb D.S.	3093	2908								
METERED DRYER STEAM ◁9 x.337	Counts										
	Counts/hr	388									
	lb/hr	131									
	Btu/lb D.S.	2869	2611								
NET DRYER STEAM ◁10	Btu/lb D.S.	2172	1977								
	Theor. Dryer Stm Btu/lb D.S.	2137	1944								
THERMAL EFFICIENCIES ◁11	Overall Dryer	75.5									
	Overall Process										
	Boiler	61.7									
COOLING WATER ◁12	~~Gal~~										
	lb/hr										
water still steam, #/hr		19.8									
Silver still steam, #/hr		15.7									
net water still steam, #/hr		10.8									
net Silver still steam, #/hr		9.0									
wat still steam Btu/#D.S.		432									

TABLE 2. DATA REDUCTION: THERMAL & ELECTRICAL LOADS

		@ M	T	W	Th	@ F	½@ S	S	M	T	@ W	Th	F
DATE		3/6	3/7	3/8	3/9	3/10	3/11	3/12	3/13	3/14	3/15	3/16	3/17
HOURS ON	Sludge	11.0	17.8	15.7	8.0	23.4	15.1	24.0	11.3	6.7	23.3	20.5	13.0
	Water 0.5/hr	54.0	53.5	49.6	44.0	41.5	47.8	43.3	41.6	45.5	59.7	73.9	77.6
ELECTRICAL hr	kWh (x2)	135 \ 9	114 \ 4.5	201 \ 12	105 \ 6	167 \ 10	76 \ 4	334 \ 20	71 \ 4	104 \ 6.3	310 \ 20	186 \ 11	175 \ 10
	kWh/lb D.S.	.62	.94	.67	.80	.80	.73	.77	.85	.72	.52	.46	.45
FUEL OIL ①5.7 Gpi 32-42 ①5.85 Gpi 5"-12 ◁6	~~gal~~ INCHES	2.0 \ 8	1.5 \ 4.5	3.0 \ 12	1.75 \ 6	4.5 \ 10	1.5 \ 4	6.0 \ 20	1.25 \ 4	2.0 \ 6.3	1.5 \ 4	3.25 \ 11	2.5 \ 10
	Gal ~~lb~~/hr	2.85	2.33	1.75	2.04	2.57	2.63	1.91	2.19	2.22	2.14	2.07	1.75
	Fuel Oil Btu / lb D.S. #/hr	20 6829	16.3 5656	12.3 4582	14.3 6021	18.0 8042	18.5 7146	13.4 5729	15.4 6837	15.6 6337	15.0 4655	14.5 3638	12.3 2929
METERED BOILER ~~STEAM~~ H₂O x 6.71 ◁7	gal	130 \ 8	113 \ 4.5	233 \ 12	129 \ 6	213 \ 10	93 \ 4	398 \ 20	71 \ 4	126 \ 6.3	431 \ 20	228 \ 11	193 \ 10
	lb/hr	134	207	160	177	175	191	164	146	165	177	171	159
	Btu/lb D.S.	2690	4211	3512	4384	4604	4364	4123	3828	3943	3238	2516	2231
NET PROCESS STEAM ◁8	Btu/lb D.S. corr	1146	2649	1824	2477	2574	2595	2190	1808	2108	1826	1386	1152
METERED DRYER STEAM x.3 920 ◁9	Counts	3018 \ 8	2120 \ 4.5	4313 \ 12	1876 \ 6	3215 \ 10	1308 \ 4	6809 \ 20	1369 \ 4	2303 \ 6.3	7797 \ 20	4416 \ 11	4094 \ 10
	Counts/hr	377	471	359	313	322	327	340	342	366	390	401	409
	corr lb/hr	113	141	108	94	97	98	102	103	110	117	120	123
	Btu/lb D.S.	2022	2430	1998	1963	2141	1888	2167	2269	2220	1803	1498	1455
NET DRYER STEAM ◁10	Btu/lb D.S.	1374	1874	1409	1296	1441	1270	1487	1570	1577	1309	1096	1079
	Theor. Dryer Stm Btu/lb D.S.	2692	2224	2624	1794	1794	2287	1939	1775	1717	1615	1817	1841
THERMAL EFFICIENCIES ◁11	Overall Dryer	68	77	71	66	67	67	69	69	71	73	73	74
	Overall Process	16.8	46.8	39.8	41.1	32.0	36.3	38.2	26.4	33.5	39.2	38.1	39.3
	Boiler	39.3	74.4	76.6	72.8	57.2	61.1	72.0	56.0	62.2	69.6	69.1	76.1
COOLING WATER ◁12	Gal												
	lb/hr		41.8										82.5 72.6
Water still steam counts x.05		3352 \ 8	2905 \ 4.5	5296 \ 12	3054 \ 6	5234 \ 10	2135 \ 4	9701 \ 20	2244 \ 4	2847 \ 6.3	9241 \ 20	4993 \ 11	4528 \ 10
Solvent still steam counts x.075		2129 \ 4	851 \ 4.5	3623 \ 12	1536 \ 6	1941 \ 10	466 \ 4	3119 \ 20	558 \ 4	1147 \ 6.3	3464 \ 20	1214 \ 11	945 \ 10
Water still # steam/hr		24.1	32.3	22.1	25.5	26.2	26.7	24.2	25.1	22.6	23.1	22.7	22.6
Solvent still # steam/hr		19.0	14.2	22.6	19.2	14.6	8.74	11.7	10.5	13.7	13.0	8.3	7.1

TABLE 2. DATA REDUCTION: THERMAL & ELECTRICAL LOADS

		T	ⓐ W	T	F	S	ⓑ S	M	T	W	T	ⓑ F
DATE		2/21	2/22	2/23	2/24	2/25	2/26	2/27	2/28	3/1	3/2	3/3
HOURS ON	Sludge	23.3	17.9	15.3	11.0	10.9	11.6	10.2	16.0	12.7	20.4	10.
	Water # d.s.	41.7	46.6	34.5	31.3	38.3	31.9	36.3	43.1	37.4	42.7	44.8
ELECTRICAL	kWh	328 \ 20	135 \ 8	☆	70 \ 4	106 \ 6.5	98 \ 6	66 \ 4	136 \ 8	89 \ 5.5	212 \ 13	94
	kWh/lb D.S.	.78	.83		1.12	.61	1.02	.91	.79	.86	.76	.72
FUEL @5.7 gpi 32"-42" @ 5.85 gpi 5"-12" OIL ⊲6	gal INCHES	3.3 \ 8	2.5 \ 8	☆	2.0 \ 4	1.75 \ 6.5	3.0 \ 6	1.0 \ 4	3.25 \ 8	1.8 \ 5.5	5.25 \ 13	2.1
	~~lb~~ gal/hr	2.89	1.78		3.5	1.88	2.93	1.75	2.84	2.29	2.82	2.1
	Fuel Oil Btu / lb D.S. #/hr	20.3 9000	12.5 5693		24.6 14522	13.2 6390	20.6 11928	12.3 6261	19.9 8556	16.1 7918	19.8 8577	15 6262
METERED X8971 BOILER STEAM ⊲7	gal hrs	416 \ 20	160 \ 8	☆	95 \ 4	138 \ 6.5	122 \ 6	85 \ 4	141 \ 8	127 \ 5.5	295 \ 13	120
	lb/hr	173	165		198	175	167	175	145	190	187	173
	Btu/lb D.S.	4474	4419		6807	4973	5718	5252	3669	5539	4768	42
T -77 PROCESS STEAM ⊲8	Btu/lb D.S. .52	2509	2362		2681	2789	3675	2942	1721	3293	2807	245
METERED DRYER STEAM X.3 ⊲9 should be /337	Counts hrs	7047 \ 20	2701 \ 8	5425	1536 \ 4	2717 \ 6.5	2539 \ 6	1724 \ 4	3718 \ 8	3133 \ 5.5	4795 \ 13	219 \ 5.7
	Counts/hr	352	338	☆	384	418	423	431	465	570	369	377
	lb/hr	106	101		115	125	127	129	140	171	111	113
	Btu/lb D.S.	2330	2298		3380	3012	3660	3269	2978	4204	2385	2323
NET DRYER STEAM ⊲10	Btu/lb D.S.	1632	1563		2440	2234	2740	2458	2305	3443	1702	1665
	Theor. Dryer Stm Btu/lb D.S.	2390	2343	2175	2523	2504	2418	2438	2212	1993	1966	211
THERMAL EFFICIENCIES ⊲11	Overall Dryer	70	68		72	74	75	75	77	83	71	72
	Overall Process	27.9	41.5		18	43.6	25.8	47.0	20.1	41.2	32.7	39.
	Boiler	49.7	77.6		46.9	77.8	47.9	83.9	42.3	69.4	55.5	67.3
COOLING WATER ⊲12	Gal											
	lb/hr				33.3		31.6		45.8	38.2		[illegible]
Water Still steam count x.05		8553 \ 20	3277 \ 8	☆	1796 \ 4	3318 \ 6.5	2283 \ 6	2167 \ 4	3628 \ 8	1732 \ 5.5	7268 \ 13	2715 \ 5.7
Solv Still steam count x.075		6674 \ 20	1203 \ 8	☆	834 \ 4	633 \ 6.5	1329 \ 6	1092 \ 4	1428 \ 8	1235 \ 5.5	3232 \ 13	2253
Water Still #steam/hr		21.4	20.4		22.4	25.5	19.0	27.1	22.7	15.7	28.0	23.8
Solvent Still #steam/hr		25.0	20.6		15.6	7.3	16.6	20.4	13.4	16.8	18.6	29.

TABLE 3. ANALYTICAL DATA

Average of 23 Days of Daily Averages

DATE		Avg			RANGE MAX	MIN					
INPUT SLUDGE	T.S., %	17.7			21.4	15.1					
	S.S., % by calc.										
	Oil, %										
DRY SOLIDS	TEA, %	0.24			.56	.004					
	T.S., %	89.6			98.2	68.7					
	Oil, %										
PRODUCT WATER	TEA, ppm	922			4000	0					
	T.S., %	5216			15160	1000					
	S.S., %	3276			11400	200					
	pH	8.7			9.6	7.4					
WET CAKE	Solids, %	25.7			29.1	20.8					
	TEA, %	35.3			49.5	23.0					
	Water, %	39.0	Q = 2133		51.6	25.5					
CENTRIFUGE CAPTURE, %		94			99	84					
CENTRATE, TEA/WATER RATIO		5.4			6.7	4.3					
RECYCLE TEA	Oil, %	1.07			2.23	0.61					
	Cloud, Pt., ^{o}C										
SOLVENT STILL OIL	TEA, %										
	Water, %										
	Solids, %	1.2			1.53	.83					
SOLVENT STILL WATER	TEA, %										
	Solids, %										
RAG RECYCLE	T.S., %										
	S.S., %										
input sludge pH.		7.4			8.3	6.5					
input sludge, % volatiles		75			81	70					

NEW FEED 0300

50% 2nd
NEW FEED 0550

TABLE 3. ANALYTICAL DATA

NEW Feed 1340 10% 2nd.

		T	W	T	F	S	S	M	T	W	T	F
DATE		2/21	2/22	2/23	2/24	2/25	2/26	2/27	2/28	3/1	3/2	3/
INPUT SLUDGE	T.S., %	17.0	16.5	—	—	19.1	—	—	20.1	—	16.2	16
	S.S., %											
	volatile solids % Oil, %	78	78			81			79		71	–
DRY SOLIDS	TEA, %	.42	.57	.50	.007	.004	.017	.004	.013	.069	.21	.1
	T.S., %	87.2	75.6	97.5	90.9	96.6	93.2	98.7	95.6	95.7	97.8	98.2
	Oil, %											
PRODUCT WATER	TEA, ppm	894	680	657	160	183	130	1500	423	560	390	0
	T.S., ~~%~~ PPM	6600	4800	2870	1600	1950	1600	1800	4000	1700	3270	27
	S.S., %	4500	3100	870	200	350	200	1300	2230	750	1530	600
	pH	8.9	8.5	7.4	9.6	9.0	8.9	9.1	8.8	8.9	8.9	8.
WET CAKE	Solids, %	23.8	25.7	25.8	25.2	26.4	26.3	25.8	26.9	29.1	28.1	25.
	TEA, %	35.2	26.6	33.7	26.0	23.0	24.7	26.1	28.6	27.7	32.0	37.5
	Water, %	41.0	47.7	40.5	48.8	51.6	49.0	48.1	44.5	43.2	39.9	37.
CENTRIFUGE CAPTURE, %		89	87	92	98	98	96	99	97	94	98	98
CENTRATE, TEA/WATER RATIO		4.9	5.5	6.7	6.1	6.7	—	6.7	6.5	5.7	5.7	4.
RECYCLE TEA	Oil, %	0.61	0.79	0.78	0.74	0.84	0.75	0.69	0.75	0.69	0.80	0.8
	Cloud, Pt., °C											
SOLVENT STILL OIL	TEA, %											
	Water, %											
	Solids, %											
SOLVENT STILL WATER	TEA, %											
	Solids, %											
RAG RECYCLE	T.S., %											
	S.S., %											
Input sludge ph		7.1	7.1	—	—	8.3	—	—	7.8	—	7.7	7.

TABLE 3. ANALYTICAL DATA

new feed 2025 (3/6) — NEW FEED 1615 (3/9) — NEW FEED 1720 (3/14)

DATE		3/6	3/7	3/8	3/9	3/10	3/11	3/12	3/13	3/14	3/15	3/16	3/17
INPUT SLUDGE	T.S., %	—	15.9	15.1	—	17.1	—	—	16.8	—	—	20.6	21.4
	S.S., %												
	Volatile solids, % / Oil, %		81	73		73			70			72	73
DRY SOLIDS	TEA, %	.19	.38	.49	.21	.29	.30	.25	.29	.19	.29	.36	.32
	T.S., %	68.7	71.3	72.5	81.9	95.8	89.4	96.6	91.2	92.7	92.9	81.3	88.0
	Oil, %												
PRODUCT WATER	TEA, ppm	790	1075	1400	1200	737	320	1290	790	920	4000	2000	1100
	T.S., %	5300	4650	9225	2500	7130	3700	15100	12200	4200	9930	4650	9100
	S.S., %	3500	1750	6350	300	3500	500	11400	8000	2400	5130	2100	6800
	pH	8.7	8.5	8.2	8.8	8.9	8.5	8.9	8.7	8.7	8.7	8.6	8.4
WET CAKE	Solids, %	24.2	25.0	20.8	27.3	26.8	24.8	24.3	24.4	25.0	26.7	27.3	26.7
	TEA, %	25.4	36.5	41.3	40.0	41.6	34.2	45.6	49.8	49.5	47.3	39.3	40.5
	Water, %	50.4	39.5	37.9	32.7	31.6	41.0	30.1	25.8	25.5	26.0	33.4	32.8
CENTRIFUGE CAPTURE, %		95	96	95	95	95	94	93	94	99	91	84	93
CENTRATE, TEA/WATER RATIO		5.7/1	4.6/1	5.0/1	4.6/1	4.5/1	4.3/1	4.6/1	4.6/1	4.9/1	5.4/1	5.3/1	6/1
RECYCLE TEA	Oil, %	0.74	1.0	1.0	0.95	1.29	1.44	1.36	1.30	1.39	1.51	2.00	2.23
	Cloud, Pt., °C												
SOLVENT STILL OIL	TEA, %												
	Water, %												
	Solids, %												
SOLVENT STILL WATER	TEA, %												
	Solids, %												
RAG RECYCLE	T.S., %												
	S.S., %												
Input Sludge pH		—	7.0	6.5	—	7.2	—	—	7.5	—	—	7.3	7.2

APPENDIX D

JOHN ZINK PROCESS SYSTEMS

TULSA **OKLAHOMA**

FUEL EVALUATION TEST REPORT

Customer: Resource Conservation Co. Renton, Washington
Customer P.O.: 46-5385
John Zink S.O.: 097151
Test Dates: May 30-31, June 1-3, 1978
Personnel: Joe Bostjancic
John Cegielski - John Zink Company
Jake Campbell - John Zink Company
Dan Banks - John Zink Company
Joe Gilliland - John Zink Company

OBJECTIVES:

1. Evaluate the performance of a John Zink cyclonic type high temperature oxidation unit on burning a solid waste submitted by the RCC. This objective is to be accomplished by visual observation of burner operation, analyses of gaseous products of combustion and residual solids.

2. Conduct a "heat balance" on the burner/boiler system to determine heat recovery, thermal efficiencies, and heating value of the waste. Compare heating value as determined by burner test with laboratory calorimeter results.

3. Observe stoking characteristics of the waste and determine suitable methods for feeding.

4. Conduct a material balance on the system based on analyses of waste and feed rate of waste and combustion air.

5. Determine fuel/air ratio used in the test and evaluate expected ratios to be used in commercial operations for complete combustion, operating temperatures, and boiler efficiency.

6. Determine quality of flue gases with respect to SOx and ash contamination.

7. Determine distribution and condition of residual ash and arrive at methods of removal from the unit and flue gas stream.

WASTE DESCRIPTION:

The solid waste was received in plastic bags ranging in size from 20 lbs. to 105 lbs. Composition of the waste varied from bag to bag. A handful of waste was sampled from each of the 324 bags fed (14,810 lbs total). These samples were thoroughly mixed and a representative sample taken for analyses. Samples were also taken for analyses of waste representation of the dry, wet, and rancid waste. Analyses of these samples are given below.

Component	Composite Wt. %	Rancid Wt. %	Dry Wt. %	Wet Wt. %
Water Content	13.3	21.0	2.8	29.5
V.C.M.	50.8	49.2	65.8	38.3
Fixed Carbon	14.5	16.6	15.8	12.6
Ash	21.3	13.2	15.6	19.5
Sulfur	0.51	0.42	0.17	0.36
	BTU/#	BTU/#	BTU/#	BTU/#
Heating Value (HHV)	5370	5120	6360	4430

The material was rather light weight ranging in bulk density from 8 to 20 lbs/ft^3.

TEST EQUIPMENT:

1. Waste Feed System -

   One of the major problems, if not the major one, was obtaining a feed system that had the desired capacity - 500-600 lbs/hr. Several systems were tried. The following is a description of each system.

   A. Solids were placed in a feed hopper with a Sprout-Waldron rotary discharge valve on the bottom. Solids were discharged through the valve into a 2" air conveying system. The air conveying system fed the solids and conveying air into the Combusticlone. The conveying system entered the combusticlone tangentially on the side at 45° (0°Top) approximately two feet back of the front plate. Near the Combusticlone an air seal was installed in the two inch pipe to overcome the positive pressure in the Combusticlone. The air seal operated on a 15-30 psig pressure drop across 6-1/8" ports jetting air towards the Combusticlone. Several different poking devices were used to stir the solids in the hopper in a effort to keep the rotary valve full of solids.

B. Same system as described unaer (A) only a portable mixer was substituted for the poker to keep the solids from packing.

C. Same system as described under (A) only a compressed air hose instead of a poker was used to keep solids from packing the feed hopper.

D. Same system as described under (A) with the addition of an Acrison variable speed screw feed was placed above the feed hopper in a effort to maintain just enough solids in the feed hopper to cover the rotary valve and eliminate bridging in the hopper.

E. Same system as described under (D) with a portable mixer installed in the feed hopper on the screw feeder to prevent the solids from bridging.

F. Same system as described under (E) using two Acrison screw feeders instead of one.

G. The rotary valve was removed from the system and solids were discharged from the Acrison screw feeders into a single open funnel - 18" dia. at the top and connecting in to a 2" pipe at the bottom. A John Zink standard air seal was installed just below the small hopper educting the solids into the 2" air conveying line. The air seal installed earlier near the combusticlone was left in service.

H. Same system as described under (G) only two funnels were used (one for each feeder) instead of one.

I. Same system as described under (H) only another 2" air conveying line was installed (one for each feeder). The second air conveying line was installed in the sight port located in the man-way in the front plate of the Combusticlone. The inlet feed was located 12 in. below center line pointing upwards. Initially, the feed hoppers on the screw feeders were agitated manually with a stick. About two-thirds through the run mechanical agitators were placed in the hoppers. The agitators were six "fingers" made of 1/2" dia. rod welded on the agitator shaft space 4 inches apart. Length varied from 6" near the bottom to 12" at the top.

2. Burner Assembly -

The Combusticlone was equipped with three 1 x 10^6 BTU/hr spider gas burners firing tangentially into the Combusticlone. The burners were positioned at 270° (front end view with 0° at top), and the middle burner was located 18" from the front plate of the Combusticlone. Combustion air for the three burners was tapped off the main combustion air duct.

3. Fuel -

The Combusticlone burners were operated on natural gas (930 BTU/scf LHV). The burners were used to bring the unit up to operating temperature before adding waste. The burners were operated on the minimum amount of gas required to maintain flame stability.

4. Combustion Air -

Combustion and conveying (low pressure) air were supplied by a Spencer blower through a main duct to the Combusticlone. Air for the gas burner and, solids conveying system were tapped off the main duct. Flow was controlled with butterfly dampers located in the individual supply ducts off the main duct. Air flow for the combusticlone and solids conveying system were metered with orifice flanges located in the individual ducts. Total air was measured with a 5" ASME bell located on the suction of the blower. Air for combustion was regulated according to oxygen content or temperature of the combustion gas exit from the combusticlone.

5. Combusticlone -

Combustion of waste solids was carried out in a cyclonic type combustion vessel. The unit was a horizontal refractory lined vessel 5'-5" I.D. x 16' long. Discharge was 15" I.D. Combustion air entered tangentially through a 2" x 9" slot with centerline located 18" from front plate of Combusticlone. Orientation (front end view) was (a) waste feed - 50°, (b) combustion air - 45°, and (c) gas burners - 270°. Combustion air swirled in a counter clockwise motion.

6. No. 1 Tower - Cross over duct - No. 2 Tower

Primary purpose of the towers was to transport the process stream from the Combusticlone to the waste heat boiler. No. 1 Tower was a vertical refractory lined tower 3'-8-1/2" I.D. x 15'-0" high. The Combusticlone was mounted horizontally with its center 4" above grade. Tower exit nozzle was 14" I.D. and located horizontally 12' above grade.

7. Waste Heat Boiler -

Flue gases were discharged from Tower No. 2 into a water tube boiler. The boiler consisted of one steam and one mud drum, with 15 rows of tubes. The first 4 rows were bare and the other 11 rows finned. The boiler was designed for clean gas service, and no provision had been made for operating with soot blowers. Two "home made" soot blowers were installed for removing solids build up. One was located in the boiler entrance and the other in the boiler exit. The soot blowers were constructed of 1' pipe with 4 holes on 6" centers.

8. Vent Stack -

Flue gases were discharged from the boiler into a 22'-0" high vent stack. A 14" O.D. duct ("U" section), connected to the top of the vent stack, reversed the gas flow downwards to a vertical venturi scrubber. A water quench spray nozzle was located in the top of the "U" duct discharging downwards in to the section connected to the Venturi Scrubber.

9. Venturi Scrubber -

The flue gas stream went through a standard John Zink trimmable Venturi Scrubber to remove solid particles from the flue gas stream. Flue gases were scrubbed with recycle water. The water was recycled with an Oberdorfer gear pump connected to a 4' diameter x 4' high catch tank located directly underneath the Venturi Scrubber discharge.

10. Instrumentation -

Six thermocouples (Type K) were used to measure the temperature of the flue gas stream at different points in the system. Thermocouples were placed in the following locations:

A. Combusticlone front - thermocouple was located on the side of the Combusticlone approximately 4' back of the front plate.

B. Combusticlone exit - Thermocouple was located near the middle of the horizontal discharge nozzle going from the Combusticlone to No. 1 Tower.

C. No. 1 Tower exit - Thermocouple was located in the entrance of the cross over duct between No. 1 Tower and No. 2 Tower.

D. No. 2 Tower entrance - Thermocouple was located in the exit of the cross over duct between No. 1 Tower and No. 2 Tower.

E. No. 2 Tower exit - Boiler Inlet - Thermocouple was located in the duct connecting No. 2 Tower to the boiler approximately 1'-6" downstream from No. 2 Tower exit and 2' upstream from boiler entrance.

F. Boiler exit - Thermocouple was located in the duct connecting the boiler exit to the Vent Stack approximately 2' downstream from the boiler exit.

Thermocouples were connected to a Westronics multipoint temperature indicator.

A flue gas sample tap for oxygen analysis was located in the crossover duct near No. 1 Tower exit. The oxygen analyzer was a Teledyne oxygen/combustibles continuous analyzer. Analyses for other components in the gas stream were conducted on a Hewlett Packard 5840A Gas Chromoatograph.

Combustion air supplied by the blower, was measured with a metering orifice located in the duct to the Combusticlone. Total air supplied by the blower, was measured with a 5" ASME bell located on the suction of the blower. Waste feed was measured by weighing each bag of waste using a platform scale. Feed rate was determined by determining the amount fed on definite time intervals. Auxiliary fuel gas and boiler feed water flow rates were measured with rotameters. Fuel gas, educting air, and steam pressure were measured with Bourdon type pressure gauges.

Static pressures of the flue gas stream at the boiler and Venturi Scrubber entrance were measured with magnehellic type gauges. Flue gas ΔP across the boiler was measured with a U tube mannometer.

PROCEDURE:

Pilot run - A pilot run was conducted Friday, May 26, 1978, prior to conducting the main test the week of May 29th to determine what operating difficulties might be encountered. Our main concern was coming up with a feed system capable of feeding 500-600 lbs/hr of light weight waste material.

The unit was brought up to operating temperature (1800°F) on natural gas prior to introducing waste feed. As waste was brought on stream, the auxiliary fuel gas was reduced to 0.308 x 10^6 BTU/hr. Approximately 200 lbs of waste was fed with feed rates of 500 and 700 lbs/hr. Results from the pilot run indicated the feed system described under Test Equipment: 1. Waste Feed System, system (A) (Rotary valve feeder with air conveying system) would be satisfactory for feeding 500-600 lbs/hr continuously over an extended period of time. Based on the pilot run, we could not forsee any problems operating the test unit for an extended period of time.

Main run - The unit was brought up to operating temperature the morning of 5/31/78 in the same manner as the pilot run prior to adding waste. Waste feed was started at 2:15 p.m. 5/31/78. By 2:30 p.m., it was evident the feed system used the previous Friday would not feed 500-600 lbs/hr on a continuous basis. Different poking devices were tried to maintain the rotary valve full of solids and still not pack hard enough to prevent gravity dumping of the solids as each section rotated into the air conveyer.

3:00 p.m. - A mechanical agitator was installed in the feed hopper to stir the solids and maintain uniform flow from the rotary valve. The feed was still spasmotic and far below capacity.

4:00 p.m. - An air hose was placed in the feed hopper to suspend the solids in an effort to maintain uniform flow from the rotary valve. The feed was still spasmodic.

6:00 p.m. - An Acrison screw feeder was placed over the hopper on the rotary valve for two stage feeding. The idea being to just cover the roatry valve and eliminate bridging of solids in the rotary valve hopper. The screw feeder was discharging at a faster rate than the rotary valve could handle.

7:00 p.m. - A trial run was made removing the rotary valve feeder and discharging from the screw feeder in to a funnel with an air seal educting the solids into the air conveying line. This method was more satisfactory with respect to uniform feed and capacity than any of the previous methods. The balance of the night was spent installing the system described under Test Equipment: 1. Waste feed system, system (g). An air compressor was also obtained to assure a dependable and ample source of compressed air for the air seals.

6/1/78 7:00 a.m. - The feed system installed overnight was tried and appeared to be satisfactory only the single screw feeder was of insufficient capacity and two feeding systems were installed into a common air conveying line (system h).

3:00 p.m. - The dual feed system was completed and tested. It was able to achieve uniform feed at 600-700 lbs/hr on a continuous basis.

The continuous 36 hour test was started officially at 4:10 p.m. 6/1/78 and continued until 4:00 a.m. 6/3/78.

Every effort was made to achieve continuous steady-state operating conditions. Each bag of waste was weighed and the contents placed in the screw feeder hoppers. No attempt was made to blend the wet, rancid and dry material. The screw feeders were set at the maximum discharge rate the air eductors would handle. Even though a constant feed rate of 600 lbs/hr was our goal, the feed rate varied from 250 lbs/hr to 750 lbs/hr and as low as 100 lbs/hr near the end of the 36 hr. run. The above feed rates were calculated from the amount fed on 30 minute intervals. Tabulation of waste added and feed rate is given in Table 1. The slower feed rates occurred when feeding wet material, which would plug in the feeder causing an occassional down time. Each feed hopper was continuously stirred by an operator with a rod to keep the material from caking or bridging over in the hopper. For the last quarter of the run, manual agitation was replaced with mechanical agitation. Mechanical agitation was achieved using a mixer with "fingers" on the rotating shaft. The fingers would "shear" the solids preventing bridging and caking. Even with agitation in the hopper, discharge from the screw conveyer was not as uniform as desired. Lack of uniform feeding gave considerable fluctuations in combusticlone exit temperature and oxygen content (excess air) in the flue gas. The bulk of the combustion air was supplied by the air seal eductors on the waste feed system. The control damper in the combustion air duct was adjusted to maintain the air blower just above surging. Combustion air was maintained constant for the first 9 hours of operation. It was increased proportionally when the auxiliary fuel was increased from 0.304×10^6 BTU/hr to 0.655×10^6 BTU/hr. The auxilary fuel was increased to maintain constant operating temperature. The auxiliary fuel and air were increased again after 26 hours of operation to maintain constant operating temperature.

Boiler feed water was added continuously to the boiler. It was adjusted to maintain as constant a water level in the boiler as possible. The "home made" soot blowers were operated just once during the test. The boiler was "blown" after 17 hours of operation.

Recycle scrubber solution to the Venturi Scrubber was regulated to 18 gpm maintaining 15-20" H_2O ΔP on the Venturi Scrubber. Fresh water make-up was regulated to maintain a constant level in the catch tank.

A "grab sample" of the waste was taken from each bag. The "grab sample" was placed in sample container. Each container represented four hours of operation.

Flue gas samples for gas chromatograph analyses by John Zink Company were taken according to the following schedule

6/1/78	6:20 p.m.
	11:00 p.m.
6/2/78	4:00 a.m.
	12:30 p.m.
	3:30 p.m.

Two EPA stack gas particulate runs were conducted, one at 10:00 and one at 11:00 a.m. 6/2/78 by Wilson Wright Laboratories.

Two Anderson classifier runs were made at the same time by Wilson-Wright Laboratories to determine particle size distribution of the particulates in the stack gases.

Flue gas samples for SO_2 and SO_3 analyses were taken by Wilson-Wright and John Zink Company according to the following schedule.

Date	Time	Sampler
6/1/78	6:00 p.m.	John Zink
6/2/78	10:00 a.m.	Wilson Wright
6/2/78	11:00 a.m.	Wilson Wright
6/2/78	11:45 a.m.	John Zink
6/2/78	3:00 p.m.	John Zink

A post test inspection was made of the combusticlone, boiler, and venturi scrubber.

RESULTS AND COMMENTS:

Visual observations of burning operations indicated the combusticlone performed according to design. The waste entered the combusticlone just below the combustion air entrance discharging upwards into the air stream. The waste solids appeared to be dispersed immediately into the combustion air stream stretch in a band approximately 18" wide. At low and average feed rates (200-400 lbs/hr) solids were observed from the entrance to a point 25-30% around the periphery of the Combusticlone. In this region, combustion appeared to be taking place, but the bulk of combustion appeared to take place from the burner region until the waste had completed the better part of a revolution. Velocity of the air stream appeared to be sufficient to keep the solids slung against the wall. Flame appeared to be 5-6" thick staying along the wall of the refractory. It was difficult to observe exactly what effect the waste stream entering the front of the Combusticlone had on the gas and solids flow pattern. There was more combustion taking place when the stream entered, and combustion of this stream seem to take place more on the bottom of the Combusticlone indicating poor suspension of solids. Combustion appeared to be complete in the front 1/3 of the Combusticlone. Very few "fire flies" were observed. These disappeared before they reached the Combusticlone exit. Observing the solids entering the Combusticlone, feed rate was very spasmodic even though solids were discharging continuously from the screw feeder. Every few seconds a large "slug" would enter the Combusticlone which was 200-300% larger than the normal "dribble" coming in. These "slugs" did not appear to seriously effect burner operations. Slugging did cause a wider band of fire, but combustion appeared to still be complete in the front 1/3 of the Combusticlone.

At average to high feed rates (400-750 lbs/hr), it was difficult to observe solids distribution due to increase intensity of the flame and larger quantity of fire. The flow path had increased from 5-6" to 12-24" thick. Length of combustion increased from 1/3 to 1/2-2/3 of the length of the Combusticlone. "Fire flies" produced had increased, with 95-98% disappearing by the time they had reached the Combusticlone exit. Waste feed "slugging" was similar to lower feed rates. At the maximum average feed rate (10:30-11:00 p.m. 6/1/78), instantaneous feed rate when slugging occurred was probably in the range of 1200 lbs/hr which is near the capacity of the Combusticlone.

Even though the Combusticlone was capable of absorbing "slugs" of waste. For smooth operation particularly at high waste feed rates, it is felt effort should made to minimize "slugging" in the waste feed.

Recorded Notes - Joe Gilliland RCC - 5/30-6/1-2/78

After the test was completed, the Combusticlone was opened for inspection. It contained a large amount of solid deposits in the front half. The solids ran from about 2' thick at the front end to about 3' thick near the middle. Right beyond the middle, the solids sloped off to about a foot deep. The solids started at 90° (0° top) and extended around to 200°. Most of the buildup was between 90° and 180°, then tapered off to 200°. The material, for the bulk, was small particles and fairly free-flowing. There was a crust over the entire mass; but once this crust was broken, which was fairly easy to do, the material underneath was free flowing. When standing in the Combusticlone, I would sink 1-1/2' into the material. Beginning at 90°, the buildup was on a 45° angle to 2-1/2' depth. From 120° to 180°, the buildup was about a 30° slope. Above 90°, there was about a 1" deposit almost the entire length of the combusticlone. This material was harder and ran from about 90° up to 45°. From 45° up to 200° the unit was almost fairly free of deposit on the refractory sidewall. The material deposited right below the burner entrance was very hard. Color of the deposit ran from a light brown to a light gray. The light brown material was the harder material. Some of the material was in chunks of approximately 3" in diameter. It was easy to see the air pattern in the Combusticlone. Starting about 1/3 of the way down, a ridge progressed towards the front and from left to right (front end view). The three thermowells had a hard deposit. The deposit appeared to be built up over a period of time. The deposit was in the form of a V on the top side or the upstream side of the flow of gas.

Build up on the two thermowells inside the combusticlone was about 1" high. Both thermocouples looked about the same. The thermowell located in the exit nozzle had a higher buildup (3" near the top to less than 1/4" half way down the thermowell). 2,485 lbs. of ash was removed from the Combusticlone. Bulk density was 34.5 lbs/ft^3 giving 5.14 ft^3 of solids in the Combusticlone. This represents 22% of the total volume. As described earlier, the bulk of the solids was in the front half of the Combusticlone. Cause for solids build up in the Combusticlone is not known for sure. Possible explanations are (1) poor location of the second waste feed entrance (inserted in Combusticlone front plate) causing disruption in the centrifugal effect of the gas stream. (2) "Slugging" effects in the feed giving too high a concentration of solids in the gas stream which caused solids "fall out". (3) agglomeration of the ash at high temperatures producing particles too large for conveying. (4) Insufficient gas velocity to convey the solids. Solids build up in the Combusticlone can be eleminated by positioning the Combusticlone vertically rather than horizontally. In fact, the majority of John Zink's combusticlone installations are vertical.

Analyses of the flue gas stream indicated virtually complete oxidation of combustibles in the flue gas stream. CO was < 100 ppm. These values are considered good for industrial combustion equipment. Tabulation of flue gas analyses for different samples is given in Table 2.

Analyses of a representative sample of the residue in the Combusticlone is as follows:

Moisture	- 0.063% (wt.)
VCM	- 0.000
Fixed Carbon	- 1.88
Ash	- 98.05

The above analyses indicates lack of contact of the ash settling in the bottom with combustion air. Based on analyses of the composite waste residue samples, combustion was calculated to be 99.5% complete.

It is felt better contact of combustion air with the residue in the Combusticlone will greatly reduce the amount of residual carbon.

The observed heat recovery was 54% of the total heat generated. This was low for two reasons. First, due to the large amount of connecting ducts between the combusticlone and the boiler, radiation losses were high. Second, the unusually large percent excess of air leads to a reduced combusticlone exit temperature and also a reduced boiler inlet temperature.

Calculated radiation losses from the combusticlone alone accounted for ∿ 22% of the total heat generated. Calculated radiation losses from the ducting joining combusticlone and boiler accounted for another 29% of the total heat generated. Thus, radiation losses claimed ∿ 51% of the total heat generated.

As these high radiation losses are not typical of our commercial units, several calculations were made to determine the heat recovery possible had the radiation losses not been so large. By using a boiler performance computer program, a boiler was designed, capable of replicating the data obtained during the test. The following table lists observed steam rate and calculated rate at four different times.

Time of Reading	Observed Steam Rate (lb/hr)	Calculated Steam Rate (lb/hr)
4:00	835.0	807.5
4:40	817.7	797.7
5:00	848.7	837.0
5:30	850.1	841.1

In order to more closely simulate commercial operation, the steam rate was recalculated assuming no heat losses in the ducting joining combusticlone and boiler (ie. combusticlone exit temp. = boiler inlet temp.)

The table below lists calculated steam rates. For this case, heat recovery ran ∿ 77% of the total release.

Time of Reading	Calculated Steam Rate (lb/hr)
4:00	1187.9
4:40	1181.9
5:00	1300.6
5:30	1262.6

The high percent oxygen (10.3%) in the flue gas was a result of using air to control flame temperature. A more typical % oxygen in the flue gas would be 4.5%. By subtracting O_2 + N_2 (atmospheric ratio) from the flue gas and performing a heat balance, a theoretical combusticlone exit temperature (∿ 2600°F) was calculated. The following table lists observed and calculated flue gas compositions and flowrates.

Component Name	Observed Mol/hr	Observed Mol%	Calculated Mol/hr	Calculated Mol%
CO_2	9.86	8.0	9.86	12.4
H_2O	10.03	8.2	10.03	12.6
O_2	12.66	10.3	3.58	4.5
N_2	89.97	73.3	55.8	70.2
NO	0.13	0.11	0.13	0.16
SO_2	0.052	0.04	0.052	0.06
Total	122.71		79.44	

Again assuming no heat loss from connecting ducts, the steam rate for 4.5% O_2 in the flue gas was calculated, and is presented in the following table.

Time of Reading	Calculated Steam Rate (lb/hr)
4:00	1368.5
4:40	1361.9
5:00	1493.2
5:30	1451.3

In this final case, heat recovered (as 5 psig steam) accounts for 89% of the total heat generated. This thermal efficiency is indicative of recovery when using flue gas recycle to quench combustion and maintain a stable operating temperature. However, a unit using recycle quench (as well as having higher thermal efficiencies) is larger than units using air or water as quench.

Due to erratic feeding, heat balances on a "spot" basis were also erratic. However, a heat balance for the entire run and on thirty minute time intervals were good. For the 24 hour period, when data was taken every 30 minutes, total feed was 9098 lbs. of waste. Using Wilson Wrights heating value for composite waste (HHV=5370 BTU/lb, LHV-4884 BTU/lb), total heat release was 44,434,000 BTU. Heat recovered, based on actual steam production was 31,028,000 BTU giving a thermal efficiency of 69.8%. Using the average heating value derived from the test - 5331 BTU/lb. Heat release was 48,501,000 BTU lowering the thermal efficiency to 64.0%.

Comparison of calculated steam production, (based on flue gas temperatures and flow rates) and actual steam production (based on boiler feed water make up) were very good. This comparison for different time intervals is given in Table 3. Steam production varied according to waste feed rate with a minimum of 722.8 lb/hr and a maximum of 1606.5 lb/hr. Total steam production for the 24 hr. period was 28,481.5 lbs. giving an average rate of 1,186.7 lbs/hr. Calculated steam production was 29,115.2 lbs. The above figures gives a 1.4% radiation loss in the boiler system which is normal.

Comparison of calculated radiation losses (based on surface area, refractory thermal conductivities and temperature differentials) and actual radiation losses (based on flue gas temperatures and flow rates) was good. The following is a tabulation of calculated and actual heat losses for each section.

Section	Temperature (Avg) °F	Heat Flux BTU/hr-°F ft^2	Heat Loss (Calculated) BTU/hr	Heat Loss (Actual) BTU/hr
Tower #1	1700	824	226,600	243,000
Cross over	1550	754	43,300	50,000
Tower #2	1400	669	183,975	197,000
Combusticlone	2000	930	373,000	---

Heating values (LHV) of the waste were determined as the heat released by the waste divided by the pounds of waste fed. Heat released by the waste was calculated as the summation of heat recovered and radiation loss and sensible heat in flue gas boiler exit less heat released by the auxiliary fuel. Average heating value for the 24 hour period was:

$$\frac{48,501,000}{9,098 \text{ lbs.}} = 5331 \text{ BTU/lb.}$$

This value compares to 4884 BTU/lb obtained by Wilson Wright Laboratories on a composite sample.

Table 4 gives a break down for the different time intervals on heating values (based on heat balances around the Combusticlone and entire system), steam production rate, and heat recoveries.

The last system tried for feeding the waste, two screw feeders equipped with agitated hoppers discharging into air eductors appear to be the best method. The system used was rather crude and certainly needs improvement in design. The biggest problem is conditioning the waste in the screw feeder hopper so it will flow freely into the screw. Considerable difficulty was experienced with bridging and packing. The wet material had more of a tendency to pack. The shearing action of the mixer on the waste in the hopper helped tremendously. A mixer designed to give better shearing action should help. It is also felt a better educting system and larger air conveying line are necessary to prevent plugging between the screw feeder discharge and feed entrance to the Combusticlone. The screw feeder itself appeared to be satisfactory. A large quantity of air was required (0.5-1 cu. ft. per lb of waste) to convey the waste. This figure represents 25-50% of the combustion air requirements.

Due to erratic feeding, material balances, like heat balances, were difficult to obtain on a "spot" basis. Unfortunately, flue gas samples for analyses had to be taken on a short time basis. Comparison of flue gas composition based on stoichiometric calculations with flue gas analysis were very poor. To illustrate the problem, O_2 concentration in the flue gas stream was recorded every five minutes for a 30 minute period. During this time, the combustion air flow and auziliary fuel were held constant, and O_2 fluctuation were the result of feed rate fluctuations.

6/2/78	Time	% O_2	Combusticlone Temperature °F	Combusticlone Exit Gas Temp °F
	5:15 pm	6.0	2138	2110
	5:20	9.2	2024	2007
	5:25	10.8	2017	1982
	5:30	5.0	2125	2147
	5:35	8.0	2135	2119
	5:40	7.3	2152	2142

Flue gas composition was calculated on the following basis:

Waste Composition

VCM	% (wt.)
H	3.66
C	18.92
O	27.51
N	0.56
Fired C	14.50
Ash	21.3
S	0.51
H_2O	13.3

Waste feed rate - 323 lb/hr.
Combustion air rate - 720 SCFM
Comparison of calculated and chromatograph analyses are given below:

	Mol %		
Chromatograph	Co_2	0_2	N_2
6/1/78-6:22 pm	5.30	13.91	80.79
" -6:25 pm	4.87	15.13	80.01
" -11:04 pm	5.45	14.37	80.18
" -11:09 pm	2.10	17.08	80.82
6/2/78-4:00 am	5.83	13.38	80.78
" -4:05 am	0.48	18.23	81.30
" -3:32 pm	6.82	11.58	81.61
Calculated	8.75	11.24	79.85

Since good correlation was achieved on measured air and gas flow rates and heat balances, one would assume calculated values for average flue gas composition are more nearly correct.

The following tabulation gives SO_2 and SO_3 concentrations in the flue gas based on gas analyses and stoichiometric calculations:

Sample	Analyzer	SO_2 ppm	SO_3 ppm
6/1/78-6:15 pm	John Zink	19	123
6/1/78-11:50 pm	John Zink	50	130
6/2/78-10:00 am	Wilson Wright	52.5	1.1
" -11:00 am	Wilson Wright	164.8	2.4
" -3:12 pm	Wilson Wright	19	126
Calculated		500	---

Two EPA particulate runs were conducted by Wilson-Wright Laboratories to determine ash content in the flue gas. Particulate concentration was found to be 0.2166 and 0.2671 grains/std cu. ft. dry. Based on a flue gas flow rate of 850 SCFM, particulate discharge to the atmosphere would be 1.76 lbs/hr based on an average concentration of 0.242 grains/st. cu. ft. dry, for the 36 hour test, this would be 63.4 lbs. This figure does not correlate at all with the ash collected in the venturi scrubber catch tank. 12.6 cu. ft. of solids were collected in the catch tank. Using the bulk density determined for the solids in the Combusticlone of 34.5 $\#/ft^3$, total amount of solids collected by the venturi scrubber was 433 lbs. One possible explanation for the large discrepancy would be major changes in the amount of solids retained in the Combusticlone as the test progressed. The Combusticlone contained 2,485.5 lbs of solids. Based on the composite waste feed ash content of 21.3%, total ash fed to the unit was 3,154.5 lbs. Distribution was:

Combusticlone	- 78.8%
Venturi Scrubber	- 13.7%
Boiler (est.)	- .5%
Towers (est.)	- 0.8%
Vented to atms.	- 5.2%

TABLE 2

Flue Gas Analyses

Date	Sample Time	Mole % O_2	N_2	H_2	CO_2	CO	Mole ppm CH_4	C_2H_6	C_3-C_5	C_5^+	Total
6/1/78	6:22 pm	13.91	80.79	<0.01	5.30	<0.01	24.26	0.41	>0.1	2.16	26.82
	6:25 pm	15.13	80.01	<0.01	4.87	<0.01	11.28	>0.1	>0.1	6.44	17.82
	11:04 pm	14.37	80.18	<0.01	5.45	<0.01	32.98	23.28	21.96	1.40	79.61
	11:09 pm	17.08	80.82	<0.01	2.10	<0.01	1.68	>0.1	24.37	72.54	98.54
6/2/78	4:00 am	13.40	80.78	<0.01	5.83	<0.01	37.19	>0.1	19.71	1.29	58.19
	4:05 am	18.23	81.30	<0.01	0.48	<0.01	0.82	>0.1	15.20	>0.1	16.02
	3:32 pm	11.58	81.61	<0.01	6.82	<0.01	12.83	>0.1	2.52	>0.1	15.34

Summary and Conclusions

1. The Combusticlone was readily adaptable to burning solid waste material from paper pulp operations submitted by Resources Conservation Co. Maximum capacity of the Combusticlone was difficult to determine due to limited feeder capacity and erratic feeding problems. A conservative estimate would be 1,000 lbs/hr of 3.11 lbs per hour of waste feed per cubic of Combusticlone volume. Oxidation in the gaseous phase was essentially complete with an average hydrocarbon and CO concentration in the flue gas of 45 ppm and 50 ppm, respectively. Combustion was 99.55% complete based on analyses of residual ash. Elimination of solids build up in the Combusticlone to give better air-solids contact should improve percent conversion.

2. Comparison of calculated and observed heat balances on the system was very good giving a high degree of confidence in the experimental data. Average steam production from the waste only was 1,001 lbs/hr or 2.64 lbs steam per pound of waste. Thermal efficiency (heat recovered as steam/heat released) was 54%. Heat recovery was greatly reduced due to the physical set up of the system. Namely, the flue gases had to be transported through two large towers to get from the Combusticlone to the waste heat boiler. Deducting the radiation losses in the towers, thermal efficiency was increased to 83%. This efficiency would give an average steam production of 1,557 lbs/hr or 4.11 lbs steam per pound of waste. Air to fuel ratio was higher than normally used for two reasons, (1) Maintain Combusticlone temperature below 2000°F to prevent "clinkering" of ash in the Combusticlone, (2) Use of a large amount of air to prevent plugging in the solids conveying duct. Reducing the amount of excess air from 100 to 25% and using recycle gas to maintain 2000°F operating temperature would increase thermal efficiency to 89%. This efficiency would increase steam production to 1,670 lbs/hr or 4.41 lbs steam per pound of waste. Based on heat balances, average heat heating value of the waste was 5,331 BTU/lb (LHV). This value is slightly higher than the value obtained by Wilson-Wright Laboratories, Tulsa, Oklahoma - 4,884 BTU/lb (LHV).

3. One of the major problems was the waste feed system. Due to low bulk density of the waste and high moisture content in some of the waste, feeding could be a problem unless properly designed equipment is used. Further experimentation should be done to arrive at an acceptable design. The following guidelines should be followed on additional experimentation.

   A. Proper conditioning of waste to prevent bridging and caking prior to introduction into the feed system. The best method appears to be a good shearing device placed in the feed hopper.

B. Larger diameter conveying duct and shortest distance possible between feeder and Combusticlone to minimize plugging.

4. Maximum SO_x concentration in the flue gas vent was 168 ppm. This concentration is relatively low and flue gases can be vented to the atmosphere with out further treatment and still meet all known EPA requirements. A stack of reasonable height (<100 ft) would probably be required to meet EPA ground level concentrations. Ash concentration in the flue gas stream was 0.25 grains/ft^3. This value is low when compared to ash collected in the venturi scrubber. Calculated concentration based on venturi scrubber collection was 1.71 grains/ft^3. This concentration would require particulate removal from the flue gas stream.

5. Residual ash was distributed ~80% in the Combusticlone and ~20% in the vent stack flue gases. Build up in the Combusticlone was too great a magnitude to cope with in a commercial unit. Cause of build up and subsequent remedy was difficult to determine. Bulk of the ash was free flowing, and one positive solution to minimize build up would be to install the Combusticlone vertically with solids removal equipment connected to the discharge. A small amount of hard solids were deposited in the Combusticlone. These were probably formed from localized temperatures in the 2300°F range. One of the following types of equipment would be required to remove particulates from the flue gas streams before venting to the atmosphere: (a) bag filter, (b) venturi scrubber, (c) electrostatic precipitator.

	a.m.															p.m.					
Time	4:00	4:40	5:00	5:30	6:00	6:30	7:00	7:30	8:00	8:30	9:00	9:30	10:00	10:30	11:00	12:15	12:45	2:30	3:30	4:00	4:30
Temperature Combusticlone (front) °F	1870	1972	2050	2060	1894	1790	1927	2033	2052	1951	1914	1864	1706	1801	1922	1750	1800	1759	1947	1845	2077
Temperature Combusticlone (exit) °F	1775	1768	1906	1862	1696	1662	1751	1880	1937	1826	1764	1759	1600	1690	1767	1695	1760	1737	1921	1826	2089
Temperature Tower No. 1 (exit) °F	1552	1525	1610	1599	1526	1475	1521	1559	1635	1546	1513	1478	1405	1424	1473	1473	1500	1484	1602	1584	1721
Temperature Tower No. 2 (entrance) °F	1516	1487	1568	1565	1496	1443	1486	1522	1600	1507	1473	1435	1368	1392	1439	1446	1480	1455	1574	1560	1654
Temperature Boiler Entrance °F	1321	1309	1357	1362	1332	1293	1309	1316	1355	1338	1324	1292	1253	1235	1271	1301	1315	1327	1377	1388	1457
Temperature Boiler Exit °F	405	412	426	429.5	423	417.5	420.0	423.6	440	442	432	392	306	377	390	433	429	413	438	437	449
Static Pressure Boiler Entrance "H_2O	10	12	12	10	10	10	10	10	10	12	12	10	10	10	10	15	20	20	16	16	15
Static Pressure Venturi Entrance "H_2O	5.0	5.0	10.0	10.0	10.0	10.0	5.0	5.0	5.0	6.0	6.0	6	5	3.4	4	9	10	9	8	11	8
Flue Gas ΔP across Boiler "H_2O	5/16	5/16	5/16	5/16	5/16	5/16	5/16	5/16	3/8	3/8	3/8	5/16	3/8	5/16	5/16	1/2	3/8	7/16	1/2	7/16	3/8
Boiler water level of 9" sight glass	80	80	80	75	80	85	90	90	80	75	70		70	70	70	30	50	75	90	90	80
Oxygen content in Flue gas - % O_2 (vol)	11.0	8.7	7.0	10.5	13.40	11.5	11.3	9.0	8.0	10.5	12.5		12.0	12.0	9.0	10	9	10.6	7.5	10	7.5
Flow rate Total air (blower) SCFM	390	390	390	390	390	390	390	390	390	390	390	390	390	390	805	805	805	805	805	805	805
Flow rate Combustion air (blower) SCFM	320	320	320	320	320	320	320	320	320	320	320	320	320	320	590	590	590	590	590	590	590
Flow rate Aux. fuel gas 10^6 BTU/hr	0.304	0.304	0.304	0.304	0.304	0.304	0.304	0.304	0.304	0.304	0.304	0.304	0.304	0.304	0.655	0.655	0.655	0.655	0.655	0.655	0.655
Flow rate Boiler feed water - #/hr	722.8	843.3	843.3	923.6	923.6	763.0	763.0	763.0	763.0	722.8	722.8	722.8	722.8	722.8	722.8	1285	1947	1365	1285	1004	1004
Accumulative Boiler feed water added-lbs	-	361.4	783.0	1204	1646	2108	2570	2951	3333	3715	4076	4438	4799	5161	5522	6516	7158	9559	11254	11096	12398
Feed rate waste - lbs/hr	332	332	384	436	206	276	226	394	273	300	300	300	300	300	300	435	435	540	482	495	495
Accumulative waste fed - lbs	-	-	332	524	742	845	983	1096	1293	-	1566	-	-	-	-	2615		3538	3944	4105	
				A		B				C		D									

A. Stop screw feeding and use manual feeding
B. Revert back to screw feeding
C. Wilson-Wright obtaining particulate samples
D. Soot blowing boiler

Resource Conservation Corp.
S.O. 097151
June 2, 1978
June 3, 1978

6/2/78 p.m.

Time	5:00	5:30	6:00	6:30	7:00	7:30	8:00	8:30	9:09	9:30	10:00	10:30
Temperature Combusticlone (front) °F	1955	2128	1348	1561	1674	1558	2011	1964	1921	1903	1858	1676
Temperature Combusticlone (exit) °F	1937	2114	1916	1727	1692	1597	2108	2022	1920	1930	1894	1719
Temperature Tower No. 1 (exit) °F	1552	1765	1656	1539	1518	1469	1762	1739	1638	1678	1670	1539
Temperature Tower No. 2 (entrance) °F	1613	1724	1621	1507	1486	1439	1723	1700	1600	1644	1532	1493
Temperature Boiler Entrance °F	1434	1498	1466	1389	1372	1347	1499	1514	1449	1482	1488	1408
Temperature Boiler Exit °F	437	471	441	428	424	427.8	477	482.7	456	468	463	460
Static Pressure Boiler Entrance "H_2O	18	18	18	20	20	20	25	25	23	25	25	25
Static Pressure Venturi Entrance "H_2O	8	9	12	12	13	13	17	20	15	20	13	18
Flue Gas ΔP Across Boiler "H_2O	7/16	8/16	1/2	9/16	9/16	1/2	9/16	9/16	9/16	9/16	10/16	10/16
Boiler water level % of 9" sight glass	65	45	40	40	60	60	60	55	70	80	80	90
Oxygen content in Flue gas - % O_2 (vol)	11	7.2	11	13.5	12-1/4	12.8	6.4	7-1/2	12	12	11	14.5
Flow rate Total air (blower) SCFM	805	840	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000
Flow rate Combustion air (blower) SCFM	590	590	764	764	764	764	764	764	764	764	754	764
Flow rate Aux. fuel gas 10^6 BTU/hr	0.380	0.304	0.304	0.304	0.304	1.292	0.456	0.361	1.292	0.988	0.988	0.988
Flow rate Boiler feed water - #/hr	1004	1124	1285	1606	1606	1205	1365	1566	1605	1486	1486	1486
Accumulative Boiler feed water added-lbs	12009	13402	13965	14607	15410	16214	16816	17409	18232	19085	--	20571
Feed rate Waste - lbs/hr	522	522	535	535	535	558	390	332	424	552	228	300
Accumulative waste fed - lbs	4600	4861	5122	--	--	5325	6204	6399	6565	6777	7053	7197

6/3/78 a.m.

Time	11:00	11:30	12:00	12:30	1:00	1:30	2:00	2:35	3:00	3:30	4:00
Temperature Combusticlone (front) °F	1774	2047	1832	1849	2019	2000	1755	1800	1714	1676	1731
Temperature Combusticlone (exit) °F	1787	2002	1868	1855	2014	1989	1778	1759	1709	1671	1750
Temperature Tower No. 1 (exit) °F	1560	1699	1615	1606	1690	1681	1577	1534	1514	1493	1520
Temperature Tower No. 2 (entrance) °F	1514	1657	1572	1564	1639	1653	1551	1503	1484	1461	1680
Temperature Boiler Entrance °F	1417	1491	1452	1470	1483	1488	1437	1361	1352	1350	1362
Temperature Boiler Exit °F	456	478	468	469	478	475	463	436	435	436	434
Static Pressure Boiler Entrance "H_2O	25	25	25	25	27	28	25	23	20	23	23
Static Pressure Venturi Entrance "H_2O	20	20	20	20	20	20	20	20	17	18	17
Flue Gas ΔP Across Boiler "H_2O	10/16	10/16	10/16	10/16	10/16	11/16	10/16	9/16	1/2	1/2	1/2
Boiler water level % of 9" sight glass	85	65	65	70	80	80	70	90	90	75	65
Oxygen content in Flue gas - % O_2 (vol)	13.9	11.0	13.9	13.9	12.2	10.7	13.3	15.5	15.0	14.0	14.2
Flow rate Total air (blower) SCFM	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000	1000
Flow rate Combustion air (blower) SCFM	764	764	754	764	764	764	764	764	764	764	764
Flow rate Aux. fuel gas 10^6 BTU/hr	0.988	0.988	0.988	0.988	0.988	0.798	0.798	1.824	1.330	1.672	1.672
Flow rate Boiler feed water - #/hr	1205	1365	1446	1556	1486	1486	1446	1446	1446	1446	1446
Accumulative Boiler feed water added-lbs	21314	21916	22599	23321	24105	24847	25590	26313	27036	27759	28482
Feed rate Waste - lbs/hr	448	382	496	406	704	384	160	282	152	82	82
Accumulative waste fed - lbs	7347	7571	7762	8010	8211	8568	8760	8890	8981	9057	9098

WILSON-WRIGHT, INC. PETROLEUM LABORATORIES

BOX 2526 TULSA, OKLA 74101 (918) 584-6060 1427 E. ADMIRAL BLVD.

File No. 4976
June 21, 1978

John Zink Company
P. O. Box 7388
Tulsa, Ok. 74105

Attn: Mr. Russell Moody

Gentlemen:

This letter reports the results of the testing done on the four (4) Waste Samples sent to us June 8, 1978. All samples were taken June 2, 1978, except for the Composite Waste which was taken June 1 and 2.

Our Sample No.:	7908	7909	7910	7911
Your I.D.:	Composite	Rancid	Dry	Wet
Component	Wt.%	Wt.%	Wt.%	Wt.%
Water Content	13.3	21.0	2.8	29.5
V.C.M.	50.8	49.2	65.8	38.3
Fixed Carbon	14.5	16.6	15.8	12.6
Ash	21.3	13.2	15.6	19.5
Sulfur	0.51	0.42	0.17	0.36
	BTU/#	BTU/#	BTU/#	BTU/#
Heating Value	5370	5120	6360	4430

We enjoy serving you.

Yours truly,

Clair C. Wilson

Clair C. Wilson

/gm
Encl.

WILSON - WRIGHT, INC. POLLUTION DIVISION

BOX 2526 TULSA, OKLA. 74101 (918) 584-6060

File No. 4966
June 30, 1978

John Zink Company
P. O. Box 7388
Tulsa, Oklahoma 74105

Attn: Mr. John Ciegelski

Gentlemen:

The following pages report data measured during sampling of the stack from your waste heat recovery furnace while a waste Bio-sludge material from Resource Conservation Company was being burned.

The sampling and analysis requested in our conference June 1, 1978 with Joe Gilliland, Jake Campbell and Resource Conservation Company's representitive was done. These were two particulate tests using, E.P.A. Procedure, two particulate classifier tests, and sulfur oxides tests.

The filters are being retained. No other samples remain except the waste material. The waste material was reported June 21, 1978, our file no. 4976.

We enjoy serving you.

JUL 5 1978

Yours truly,

Clair C Wilson

Clair C. Wilson

CCW/gm

WILSON-WRIGHT, INC. POLLUTION DIVISION

BOX 2526 TULSA, OKLA. 74101 (918) 584-6060

John Zink Company
Resource Conservation

File No. 4966
June 30, 1978

Sampling:

A 3" neptune valve was installed on one port of the stack from your waste heat recovery furnace and ahead of the venturi scrubber, to allow insertion of sampling probes into the positive pressure in the stack at the sampling point.

The E.P.A. sampling train with a cyclone and filter in a heated box, and 4 impingers in an ice water bath was hung at the sampling point and connected to the sampling probe. The first impinger contained isopropyl alcohol, the second impinger contained hydrogen peroxide, the third was dry and the fourth contained silica gel.

The impingers were connected to a vacuum pump and the discharge from the pump metered.

When burn conditions stabilized, the sampling probe was run into the center of the stack and the test started.

A second S.P.A. particulate test run was made.

Two particulate classifier tests were made with the sample nozzle at the center of the stack and the velocity through the tip approximately equal to the velocity during the E.P.A. sampling.

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PARTICULATE GRAIN LOADING DATA Run No. 1

Container Ident.	Prob & cyc.	Filter # 133	Impinger-1	Impinger-2	Impinger-3	TOTAL
Final Wt. gms.	90.1432	1.2567	101.8334	98.0281	98.2526	-
Initial Wt. gms.	89.8858	1.0615	101.8324	98.0281	98.2505	-
Net Wt. gms.	.2574	.1952	.0010	0.0000	.0021	-
Aliquiot Ratio aliq./total vol. (mls./mls.)	1.0000	1.0000	.4000	.3333	.6000	-
Grains/Std. Cu. Ft. dry	.1215	.0922	.0011	0.0000	.0016	.2166
% of Total Collected	56.1273	42.5643	.5451	0.0000	.7631	100.00

PARTICULATE GRAIN LOADING DATA Run No. 2

Container Ident.	Prob & cyc.	Filter # 134	Impinger-1	Impinger-2	Impinger-3	TOTAL
Final Wt. gms.	97.8661	1.3592	85.5548	99.2959	98.8557	-
Initial Wt. gms.	97.6096	1.0546	85.5547	99.2959	98.8554	-
Net Wt. gms.	.2565	.3046	.0001	0.0000	.0003	-
Aliquiot Ratio aliq./total vol. (mls./mls.)	1.0000	1.0000	.4000	.3333	.6000	-
Grains/Std. Cu. Ft. dry	.1219	.1448	.0001	0.0000	.0002	.2671
% of Total Collected	45.6527	54.2137	.0444	0.0000	.0889	100.00

WILSON-WRIGHT, INC. PETROLEUM LABORATORIES

BOX 2526 TULSA, OKLA. 74101 (918) 584-6060 1427 E. ADMIRAL BLVD.

John Zink Company
Resource Conservation Co.

File No. 4966
June 30, 1978

SULFUR OXIDES MOLS PER MILLION MOLS OF GAS

Run	1	2
Impinger 1 SO_3	1.1	2.4
Impinger 2 SO_2	52.3	161.2
Impinger 3 SO_2	0.2	3.6
Total	53.6	167.2

WILSON-WRIGHT, INC. POLLUTION DIVISION
BOX 2528 TULSA, OKLA. 74101 (918) 584-6060

PARTICULATE FIELD DATA

Job No. 4966 Stack No. 1 Run No. Date June 2, 1978 Time 10:00

Probe Lenght, m 3 Heater 2 Coef. 0.64 Nozzle Dia., In. .500
Sample Box No. 1 Heater Setting °F 240 Ambient Temperature °F 75
Meter Box No. 1 Delta Ha 1 Assumed Water % 15

Run No. 1 Barometer, In. Hg. 29.48 Operator H McM

SAMPLE POINT NO.	TIME θ min.	PRESSURE Stack Ps In. Hg.	TEMP. Stack Ts °F	VELOCITY HEAD ΔPs In. H2O	METER Diff. (H) In. H2O	VOLUME Meter Vm Cu. Ft.	TEMP. Meter Ts °F	TEMPERATURES Box °F	TEMPERATURES Impinger °F	DRY STANDARD Vmc Cu. Ft.
1	0.0	29.99	444	.050	.100	0.000	76	240	32	0.000
1	5.0	29.99	390	.050	.170	4.450	76	240	32	4.306
1	5.0	29.99	383	.050	.170	4.750	76	240	32	4.596
1	5.0	29.99	369	.050	.170	4.800	76	240	32	4.644
1	5.0	29.99	334	.050	.170	5.770	76	240	32	5.583
1	5.0	29.99	374	.050	.170	3.230	76	240	32	3.125
1	5.0	29.99	373	.050	.170	6.100	76	240	32	5.902
1	5.0	29.99	370	.050	.170	4.560	77	240	32	4.404
Total	8.0					33.660				32.563
Average		29.99	384	.050	.161		76	240	32	

WILSON-WRIGHT, INC. POLLUTION DIVISION
BOX 2526 TULSA, OKLA. 74101 (918) 584-6060

PARTICULATE FIELD DATA

Job No. 4966 Stack No. 1 Run No. 2 Date June 2, 1978 Time 11:00 am

Probe Lenght, m 3 Heater 1 Coef. 0.64 Nozzle Dia., In. .500
Sample Box No. 1 Heater Setting °F 240 Ambient Temperature °F 75
Meter Box No. 1 Delta Ha 1 Assumed Water %

Run No. 2 Barometer, In. Hg. 29.48 Operator H. McM

SAMPLE POINT NO.	TIME θ min.	PRESSURE Stack Ps In. Hg.	TEMP. Stack Ts °F	VELOCITY HEAD ΔPs In. H2O	METER Diff. (H) In. H2O	VOLUME Meter Vm Cu. Ft.	TEMP. Meter Ts °F	TEMPERATURES Box °F	Impinger °F	DRY STANDARD Vmc Cu. Ft.
1	0.0	29.99	389	.050	.170	0.000	75	240	32	0.000
1	0.0	29.99	389	.050	.170	4.440	75	240	32	4.304
1	5.0	29.99	394	.050	.170	5.410	76	240	32	5.235
1	5.0	29.99	399	.050	.170	3.950	77	240	32	3.815
1	5.0	29.99	399	.050	.170	5.960	77	240	32	5.756
1	5.0	29.99	397	.050	.170	3.890	76	240	32	3.764
1	5.0	29.99	395	.050	.170	4.750	79	240	32	4.570
1	393.0	29.99	393	.050	.170	5.042	79	240	32	4.851
Total	8.0					33.442				32.298
Average		29.99	394	.050	.170		76	240	32	

WILSON-WRIGHT, INC. POLLUTION DIVISION
BOX 2528 TULSA, OKLA. 74101 (918) 584-8060

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MOISTURE CONTENT Run No. 1

Standard Cubic Feet Dry Gas	32.669
Gram Moles Dry Gas @ STP	38.517

CONTAINER	Impinger #1	Impinger #2	Impinger #3	Silica Gel
FINAL WEIGHT	562.00	655.60	452.60	410.30
INITIAL WEIGHT	522.80	612.00	445.50	394.50
NET WEIGHT	39.20	43.60	7.10	15.80

Total Grams Water Collected	105.70
Gram Moles of Water Collected	5.87
Water Content (% by Volume)	13.22

MOISTURE CONTENT Run No. 2

Standard Cubic Feet Dry Gas	32.458
Gram Moles Dry Gas @ STP	38.268

CONTAINER	Impinger #1	Impinger #2	Impinger #3	Silica Gel
FINAL WEIGHT	556.50	648.00	455.40	402.00
INITIAL WEIGHT	522.00	595.50	446.60	383.50
NET WEIGHT	34.50	52.50	8.80	18.50

Total Grams Water Collected	114.30
Gram Moles of Water Collected	6.35
Water Content (% by Volume)	14.23

WILSON-WRIGHT, INC. POLLUTION DIVISION
BOX 2526 TULSA, OKLA. 74101 (918) 584-6060

ISOKINETICITY DATA

Job No. 4966 Stack No. 1 Run No. 2 Date June 2, 1978 Time 11:00

Nozzle Dia., in. .5000 Tip Area, Sq. Ft. .001364 Pitot Coef. .640
Fraction Water .1400 Fraction Gas .8600 Mol. Wt. Gas Dry 30.00 Mol. Wt. Gas Wet 28.32

Barometer, In. Hg. 29.48 Static Stack Pres. In. Water 7.00 Stack Pres. In. Hg. Abs. 29.99

SAMPLE Point No.	TIME θ min.	TEMP. Stack Ts °R	VELOCITY Head ΔPs In. H2O	VOLUME Meter Vm Cu. Ft.	TEMP. Meter Tm °F	ACTUAL CF Dry Std. Cu. F.	THEOR. CF Dry Std. Cu. Ft.	% ISOKIN.
1	0	849	.170	0.0000	75	0.0000	0.0000	0.0
1	5	849	.170	4.4400	75	4.3174	4.9479	87.2
1	5	854	.170	5.4100	76	5.2508	4.9334	106.4
1	5	859	.170	3.9500	77	3.8266	4.9190	77.7
1	5	859	.170	4.9600	77	4.8051	4.9190	97.6
1	5	857	.170	4.8900	78	4.7285	4.9248	96.0
1	5	855	.170	4.7500	79	4.5846	4.9305	92.9
1	5	853	.170	5.0420	79	4.8664	4.9363	98.5
AVERAGE								
		854	.170		77			
TOTALS								
8	35			33.4420		32.3979	34.5109	93.9

WILSON-WRIGHT, INC. POLLUTION DIVISION
BOX 2526 TULSA, OKLA. 74101 (918) 584-6060

ISOKINETICITY DATA

Job No. 4966 Stack No. 1 Run No. 1 Date June 2, 1978 Time 10:00

Nozzle Dia., in. .5000 Tip Area, Sq. Ft. .001364 Pitot Coef. .640
Fraction Water .1400 Fraction Gas .8600 Mol. Wt. Gas Dry 30.00 Mol. Wt. Gas Wet 28.32

Barometer, In. Hg. 29.48 Static Stack Pres. In. Water 7.00 Stack Pres. In. Hg. Abs. 29.99

SAMPLE Point No.	TIME θ min.	TEMP. Stack Ts °R	VELOCITY Head ΔPs In. H2O	VOLUME Meter Vm Cu. Ft.	TEMP. Meter Tm °F	ACTUAL CF Dry Std. Cu. F.	THEOR. CF Dry Std. Cu. Ft.	% ISOKIN.
1	0	904	.170	0.0000	76	0.0000	0.0000	0.0
1	5	850	.170	4.4500	76	4.3191	4.9450	87.3
1	5	843	.170	4.7500	76	4.6102	4.9655	92.8
1	5	829	.170	4.8000	76	4.6588	5.0072	93.0
1	5	834	.170	5.7700	76	5.6002	4.9922	112.1
1	5	834	.170	4.0300	76	3.9114	4.9922	78.3
1	5	833	.170	5.3000	77	5.1345	4.9952	102.7
1	5	830	.170	4.5000	77	4.3595	5.0042	87.1
AVERAGE								
		844	.170		76			
TOTALS								
8	35			33.6000		32.5965	34.9015	93.4

WILSON-WRIGHT, INC. POLLUTION DIVISION

BOX 2526 TULSA, OKLA. 74101 (918) 584-6060

John Zink Co.
Resource Conservation Co.

File No. 4966
June 30, 1978

PARTICLE SIZE DISTRIBUTION DATA

Run No. 1

Std Cu. Ft. 3.131 Minutes 2.41 Fraction Water .0654
Classifier Temp. °F 393 Viscosity, micropoise 254

Particle Density 2.220 Impaction Parameter .14

Plate No.	Hole Diameter Centimeters	Particule Diameter Microns
1	1.1500	18.10
2	.6350	7.38
3	.3969	3.60
4	.2489	1.75
5	.1778	1.02
6	.1400	.69
7	.0940	.35
8	.0734	.22

Particle Density 2.220 Impaction Parameter .28

Plate No.	Hole Diameter Centimeters	Particule Diameter Microns
1	1.1500	25.63
2	.6350	10.47
3	.3969	5.13
4	.2489	2.51
5	.1778	1.48
6	.1400	1.01
7	.0940	.52
8	.0734	.34

Total wt. of particulates in classifier train (grams) .0372

Plate No.	1	2	3	4	5	6	7	8
Wt. gain (gms.)	.0049	.0045	.0056	.0066	.0056	.0041	.0028	.0023
% of total particulates	13.17	12.09	15.05	17.74	15.05	11.02	7.52	6.18

Total grams of particulate collected on plates .0364
Total grams of particulate through plates .0008

% of total particulates collected on plates 97.84
% of total particulates through plates 2.15

Total particulates, grains per Std. Cu. Ft. .1834

WILSON-WRIGHT, INC. POLLUTION DIVISION

BOX 2526 TULSA, OKLA. 74101 (918) 584-6060

John Zink Co.
Resource Conservation Co.

File No. 4966
June 30, 1978

PARTICLE SIZE DISTRIBUTION DATA

Run No. 2

Std Cu. Ft. 3.054 Minutes 2.41 Fraction Water .0654
Classifier Temp. °F 393 Viscosity, micropoise 254

Particle Density 2.220 Impaction Parameter .14

Plate No.	Hole Diameter Centimeters	Particule Diameter Microns
1	1.1500	18.33
2	.6350	7.47
3	.3969	3.65
4	.2489	1.77
5	.1778	1.04
6	.1400	.70
7	.0940	.35
8	.0734	.22

Particle Density 2.220 Impaction Parameter .28

Plate No.	Hole Diameter Centimeters	Particule Diameter Microns
1	1.1500	25.95
2	.6350	10.60
3	.3969	5.20
4	.2489	2.54
5	.1778	1.50
6	.1400	1.02
7	.0940	.53
8	.0734	.34

Total wt. of particulates in classifier train (grams) .0415

Plate No.	1	2	3	4	5	6	7	8
Wt. gain (gms.)	.0033	.0035	.0050	.0086	.0047	.0045	.0028	.0016
% of total particulates	7.95	8.43	12.04	20.72	11.32	10.84	6.74	3.85

Total grams of particulate collected on plates .0340
Total grams of particulate through plates .0075

% of total particulates collected on plates 81.92
% of total particulates through plates 18.07

Total Particulated, grains per Std, Cu. Ft. .2097 33

APPENDIX E

WORK PERFORMED BY

ENERGY INCORPORATED
IDAHO FALLS, IDAHO 83401

TEST REPORT

COMBUSTION OF DRIED SLUDGE

BY

L. A. POWERS
J. W. STALLINGS
C. M. YOUNG

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1.0 SUMMARY

Resources Conservation Company has developed a process to chemically dewater sludges using triethylamine (TEA). Energy Incorporated has demonstrated the feasibility of combustion of the resultant dried sludge in a fluidized-bed burner in their pilot plant in Idaho Falls, Idaho. Prior to the combustion tests, laboratory analyses were performed on a sample of the sludge to ascertain heating value and concentrations of the major constituents of concern with respect to system operation and effluent standards. Based on the results, no major problems with slagging, corrosion, or emissions are expected in the operation of a commercial-size unit. Special tests were performed to monitor NO_x emissions, which were found to be less than one ppm in each case.

2.0 OBJECTIVES

The objectives for the combustion tests of dried sludge were as follows:

(1) To demonstrate the feasibility of fluidized-bed combustion of dried sludge;

(2) To identify potential problems which would inhibit commercial operation of a system to recover energy from dried sludge;

(3) To analyze the effect of ash content on bed behavior and on particulate formation and removal; and

(4) To analyze the gaseous effluent for identification of primary pollutants.

3.0 RESULTS, CONCLUSIONS, AND RECOMMENDATIONS

3.1 Results

The feasibility of fluidized-bed combustion of dried sludge was demonstrated in the six-inch-inner-diameter pilot plant in Idaho Falls, Idaho. No agglomerations were found in the bed material, and there was no evidence of corrosion in the burner. Tests to determine NO_x concentrations in the effluent gases indicated values less than one ppm in all instances. Although the feed rates were erratic due to the nature of the fuel, average rates could be calculated to perform material balances.

3.2 Conclusions

No problems are foreseen in the commercial operation of a fluidized-bed burner to recover energy from dried sludge. Despite the relatively high concentration of bound nitrogen in the fuel (1.46%), the tests for NO_x emissions indicate that virtually none of the nitrogen leaves the burner in the form of NO_x. Sulfur concentration is low in the fuel and will not present problems with meeting SO_2 emission requirements.

One problem which will have to be approached before the advent of commercial operation is in the area of materials handling. In order to run an effective combustion system, feed rates of the fuel entering the burner must be known. The screw feeder was adequate for the demonstration of feasibility but will not suffice for commercial operation.

3.3 Recommendations

If a demonstration program is undertaken, Energy Incorporated recommends that a high-pressure transport system be employed to convey the fuel from the storage bin to the burner. A screw feeder may be used to meter the fuel into the transport system.

An economic analysis should also be performed to ascertain whether the blending of dried sludge with incoming wet sludge will provide an overall system with more favorable economics. Sludges such as those considered can be burned in a fluidized-bed system with moisture contents of fifty percent or more.

4.0 DESCRIPTION OF TESTS

4.1 Analytical Laboratory Tests

Prior to a demonstration of combustion, a series of tests were performed on a sample of dried sludge in the Energy Incorporated Analytical Laboratory. The results of the tests are provided in Table I.

4.1.1 Moisture Content

The moisture content of 4.12 percent was lower than the original value provided by RCC. Although no problems were encountered in burning this dry a fuel, an analysis should be made of the economic effects on an overall system of blending wet and dry sludges to a mixture of 50-percent moisture.

4.1.2 Volatiles and Fixed Carbon

The high value for percentage of volatiles in relation to that for percentage of fixed carbon suggests that in-bed feeding will provide more effective fluidized-bed combustion. When above-bed feeding is employed, much of the volatile material will burn either on the bed or in the vapor space. In order to maintain adequate bed temperatures, it is assumed that some of the volatile material must burn in the bed itself.

4.1.3 Ash Analysis

The ash concentration of 15.02 percent on a dry basis was higher than originally expected. However, as indicated by the softening and melting temperatures, no problems are expected with eutectic mixtures or other slagging effects. The low chloride content helps insure that corrosion will be at a minimum. This high concentration of ash will require a more extensive dry collection system than is employed in the pilot plant burner.

TABLE I

ANALYTICAL LABORATORY DATA

Moisture (wt% - wet basis)	4.12
Volatiles (wt% - dry basis)	73.70
Fixed Carbon (wt% - dry basis)	11.28
Ash (wt% - dry basis)	15.02
Sulfur (wt% - dry basis)	0.15
Nitrogen (wt% - dry basis)	1.46
Ash Softening Temperature (°C)	>1100
Ash Melting Temperature (°C)	>1100
Bomb Calorimetry (Btu/lb - dry basis)	
#1	6715
#2	6647
Average	6681
Ash Analysis (wt%)	
Na_2O	4.00
K_2O	1.08
Fe_2O_3	1.43
SiO_2	37.06
Cl^-	0.43
CaO	4.61
MgO	0.93
V_2O_5	<0.1
Al_2O_3	27.8
TiO_2	7.0
P_2O_5	1.65
SO_3	0.70

4.1.4 Sulfate Concentrations

The concentration of sulfur is low and will not present any problems in meeting local air pollution standards. Assuming all of the sulfur left the burner in the form of SO_2, only 0.45 lb SO_2/MM Btu heat input should be present. However, some of the sulfur will remain in the ash in the form of alkali sulfates.

4.1.5 Bomb Calorimetry

The heating values for dried sludge on a dry basis are higher than originally thought and show room for optimism. Two bomb calorimetry tests were performed. The average of the two tests was 6681 Btu/lb.

4.1.6 Nitrogen Concentration

The nitrogen concentration of 1.46 percent is higher than most fuels and could present problems in high-temperature combustion. Whether the nitrogen is from amine contamination or from the sludge itself is unknown.

4.1.7 NO_x Emissions

All of the results of the tests for emission of nitrogen oxides show concentrations to be less than one ppm. (The largest value was 24.3 μg/dry SCF, which is equivalent to 0.66 ppm). The bound nitrogen in the dried sludge probably forms a nitrate which joins the solid particulate. However, no tests were performed to determine whether this is the case. Also, no tests of nitrogen oxide concentration of ambient air were performed.

4.2 Combustion Tests

Three combustion tests were performed in the six-inch fluidized-bed burner at the Energy Incorporated pilot plant in order to demonstrate the combustion of dried sludge. Figure 1 is a schematic diagram of the

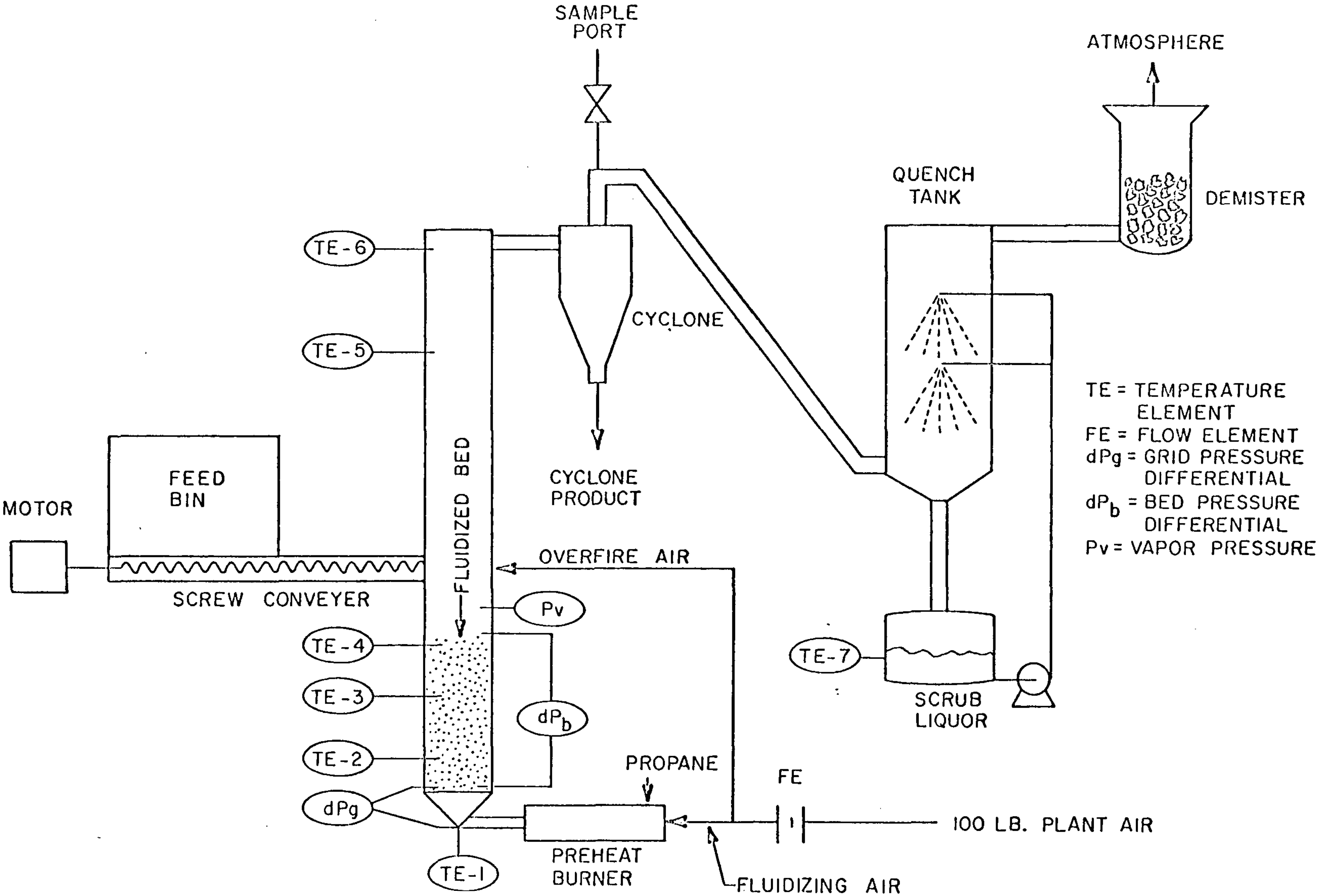

FIGURE 1
SCHEMATIC OF ENERGY INCORPORATED SIX-INCH DIAMETER FLUIDIZED BED BURNER

TABLE II

TEST DATA

COMBUSTION TEST	#1	#2	#3
Bed Temperatures (°C)			
TE-2	725	800	810
TE-3	890	840	850
Freeboard Temperatures (°C)			
TE-5	920	850	860
TE-6	870	820	820
Total Air Input (SCFM)	15.6	16.1	19.0 [17.9]*
Fluidizing air	7.0	7.0	8.1 [7.0]*
Sight-glass air (above bed)	3.8	3.9	5.2
Fuel-tube air (in bed)	2.6	2.6	2.3
Feed-hopper air (in bed)	2.2	2.6	3.4
Feed Rate (lb/hr-wet basis)	Varied	12.4	10.6
Bed Pressure Differential (in. H_2O)	17.0	16.5	15.0

Exhaust Gas (Volume % - dry basis)□	#1	#2	#1	#2
H_2	0	trace	0	0
O_2	7.8	8.9	16.4	19.0
N_2	74.5	73.2	76.0	75.4
CH_4	0	0	0	0
CO	trace	2.2	0	0

CO_2		14.8	14.5	7.6	5.6
C_2H_4		0	trace	0	0
C_2H_6		0	trace	0	0
Cyclone Particulate					
Char (lb/hr)	.09	0.10		0.04	
Ash (lb/hr)	1.67	1.75		1.21	
Loss on ignition (%)	4.84	5.62		3.32	
Total Sludge Burned (lb)	77	64.5		70	
Total Time (min)	190	280		390	
NO_x (μg/dry SCF)					
Midget Bubbler	<4.4	<4.23		24.3	
Four-hole Impinger	-	0.91		21.5	
Cyclone Particulate					
Size Range (%)=					
+50	31.8	16.3		16.4	
-50 + 80	10.5	8.5		9.5	
-80 + 140	9.4	8.6		8.3	
-140 + 200	6.3	5.8		4.7	
-200 + 270	7.1	7.1		4.8	
-270 + 400	17.8	35.3		21.2	
-400	17.2	17.4		34.0	

---

* Values in brackets were recorded for the second of two gas chromatograph samples.

□ Two gas chromatograph samples were taken for each of the last two combustion tests.

= Measured with U.S.A. Standard Testing Sieves with A.S.T.M. E-11 Specifications.

burner. The first two tests were really one long test conducted over a period of two working days. Test One was performed on Friday, June 16, and Test Two continued the efforts on Monday, June 19. Test Three, the final demonstration test, was performed on June 28, 1978. The data for the three tests are portrayed in Table II.

4.2.1 Test One

In preparation for Test One the screw feeder was calibrated for feeding dried sludge without the positive pressure in the hopper which is used during the tests to prevent movement of smoke and combustion gases from the burner to the hopper. In order to effect more uniform feed rates, rubber tips were installed on the ends of the spike roll to insure that bridging would not take place.

Test One was performed on a Friday afternoon. Propane gas was used as a start-up fuel to heat the bed material. A total of 77 pounds of dried sludge were burned during a period of over three hours. Both feed and air rates were varied over a number combinations to attain acceptable steady-state conditions. At the end of the afternoon the burner was simply turned off and left until the following Monday for Test Two. Data for the test are recorded in Table II.

4.2.2 Test Two

On Monday the burner was again heated with propane gas prior to introducing the feed of dried sludge. Feed rates were generally lower on the second day. A total of 64 pounds of sludge were burned over a period of four hours and forty minutes.

Two samples for the gas chromatograph were taken, and measurements were made of particulate formation. Tests were also performed for NO_x emissions. All results are tabulated in Table II.

4.2.3 Test Three

On Friday June 28, 1978 a demonstration of combustion was performed for personnel from Resources Conservation Company and the Idaho Office of the Department of Energy. Calibration tests were performed two days in advance with the screw feeder and hopper. During these calibration tests positive air pressure was added to the hopper to simulate combustion conditions. However, the back pressure exerted by the bed in the burner could not be simulated. Therefore, the calibration values are all relative rather than absolute.

Although the data state that seventy pounds of dried sludge were fed to the burner during a period of six hours and thirty minutes, the results of the gas chromatograph test suggest that this value for the feed rate is low. The material balance calculations at the end of the report verify this conclusion.

Data from Test Three are compiled in Table II as in the previous tests.

4.2.4 Discussion of Results

The major problem in analyzing the data from the combustion tests is the fact that the actual feed rates are unknown. Despite the use of rubber tips on the spike roll to prevent bridging, rates of feed into the burner varied at given settings of the Reeves Drive which controls the speed of the screw. Although the tests were successful, a method must be found to meter the sludge into the burner at known rates.

Calibrated rates with the screw feeder were much higher than actual rates entering the burner, due to the effect of pressure in the bottom of the bed pushing against the feed in the screw. If further tests are performed, care will be taken to insure that the actual feed rates are higher. The resulting excess air rates will then be lower. However, a certain amount of excess air is necessary to insure that the temperature of the gases leaving the burner is well below the softening temperature of the ash.

Material balances were performed for the four sets of data corresponding to the gas chromatograph tests. The poor results obtained can be explained by two factors. The first problem is the unknown feed rates discussed earlier. A second source of error involves the use of an ultimate analysis performed by Combustion Engineering for Resources Conservation Company as a source for concentrations of oxygen and carbon. Other discrepancies between CE data and those of EI, such as fixed carbon and heating value, suggests that the two samples of dried sludge varied in composition. A copy of the Combustion Engineering data is included at the end of this report (Appendix K).

5.0 PROCEDURE FOR DATA COLLECTION

During the scoping tests, data for bed and vapor pressure differentials, temperatures, and feed rates of air and dried sludge were recorded every half hour. Other tests were performed when steady-state conditions were reached.

5.1 Gas Chromatograph

Samples for gas chromatograph analysis were taken at points in time when steady-state conditions existed. Thus, all data could be recorded for each sample and approximate material balances could be performed. The tests were performed on a Varian Aerograph Series 1400.

5.2 Cyclone Particulate

Samples of particulate collected in the cyclone were taken periodically during the scoping tests. Based on these samples, calculations can be made for rate of particulate collection. Tests performed in the laboratory for loss of weight on ignition provided the breakdown between char and ash in the particulate. Particule size distribution was also determined for the particulate. Extensive data collection is not performed with particulate generation, primarily because the pilot plant is not representative of a commercial unit. In a larger burner, proper design can insure that combustion is virtually complete within the burner for most fuels. However, the low results for loss on ignition indicate that very little carbon is escaping the pilot unit unburned.

5.3 NO_x Emissions

Samples were taken during each of the three combustion tests for nitrogen oxide emissions. The samples were collected at a sampling port downstream from the cyclone in the offgas ducting. Both a four-hole sonic impinger and a midget gas bubbler were used in order to attain two different sets of data for comparison.

The two systems are used to bubble the gas effluent directly into the test solution. While a sonic-flow orifice is included in the four-hole impinger, the midget bubbler employs an orifice downstream from the test solution. In the test the gaseous effluent is pulled by means of a vacuuum pump into a 500-ml Erlenmeyer flask filled with 175 ml of sodium hydroxide/sodium arsenite solution. The nitrogen oxides are absorbed by the solution, which is then analyzed in the laboratory using spectrophotometric techniques.

The absolute pressures on either side of the orifices are monitored along with the gas temperatures. Then the total volume sampled is determined by timing the duration of the test.

This testing method is detailed in the Federal Register, Volume 38, No. 110, Friday, June 8, 1973.

5.4 Laboratory Analysis

All analytical tests were performed in the laboratory according to ASTM procedures when available.

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LAUCKS TESTING LABORATORIES

INCORPORATED

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1008 WESTERN AVENUE
SEATTLE, WASHINGTON 98104

CHEMISTS
SAMPLERS • INSPECTORS
ASSAYERS • SPECTROGRAPHERS
BIO-CLINICAL CHEMISTRIES

LABORATORY NO. 61768

DATE April 13, 1978

P.O. No. 46-5306

CLIENT Resources Conservation
P. O. Box 936
Renton, WA 98055

REPORT ON MATERIAL

SAMPLE IDENTIFICATION Eighteen samples of various materials were submitted to our laboratory for analysis. The samples were identified as follows.

TESTS PERFORMED AND RESULTS:

Sample No.	Marks	Sample Type
1	RCC 3-03-78 318	Recycle Solvent
2	RCC 3-07-78 329	Recycle Solvent
3	RCC 3-15-78 461	Recycle Solvent
4	RCC 3-03-78 320	Dry Solids
5	RCC 3-07-78 331	Dry Solids
6	RCC 3-15-78 463	Dry Solids
7	RCC 3-11-78 432	Vent Gas
8	RCC 3-12-78 433	Vent Gas
9	RCC 3-15-78 464	Vent Gas
10	RCC 3-03-78 319	Product H_2O
11	RCC 3-07-78 330	Product H_2O
12	RCC 3-15-78 462	Product H_2O
13	RCC 3-07-78 328	Feed Sludge
14	RCC 3-14-78 456	Feed Sludge
15	RCC 3-03-78 333	Unknown
16	RCC 3-08-78 334	Still Oil
17	RCC 3-22-78 518	Vent Gas
18	RCC 3-22-78 519	Vent Gas

The samples (except #15) were analyzed to determine the presence and amount of the following contaminants.

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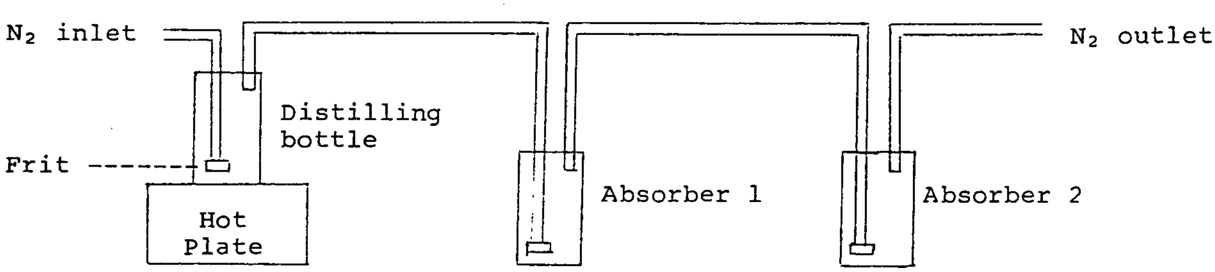
A. Ethylamine
B. Diethylamine
C. N-Chlorodiethylamine
D. Diethyl Abietamide
E. Polychlorinated Biphenyls (PCBs)

With the exception of the PCB's there was no officially recognized protocols for the analysis of these materials. The analyses were broken down into three groups. The procedures used for each group are specified in the appropriate section below.

Amine Analysis

Due to the volatile nature of these compounds we decided to isolate them by distillation. Aliquots of the sample were mixed with water and a carbonate buffer to maintain an alkaline distillation medium. The mixture was then heated to approximately 90°C and purged with nitrogen into two flasks containing an excess of hydrochloric acid. The apparatus used is shown in Figure 1 below. The amines were thus converted into the hydrochloride salts in a fairly pure form. The undistilled solution was retained for the analysis of the diethyl abietamide.

Figure 1



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The distillate was then evaporated to isolate the hydrochloride salts. This residue was then reacted with trifluoroacetic anhydride to prepare derivatives suitable for analysis by gas chromatography. The derivatives were extracted into ethyl ether, washed with sodium bicarbonate and analyzed using the following gas chromatography conditions.

1. G. C. Column------------- 6 ft. X 1/8 inch stainless steel column with 3% Carbowax 20 M on Chromosorb W 80/100 AW-DMCS
2. Detector----------------- FID
3. Temperature-------------- 140°C
4. Carrier Gas-------------- N_2 at 30 CC/min

Results of these analyses were as follows: (ppm)

Sample No.	Ethylamine	Diethylamine	N-Chlorodiethylamine
1	Less than/400	Less than/200	Less than/200
2	Less than/400	870	Less than/200
3	Less than/400	690	Less than/200
4	Less than/20	97	Less than/5
5	Less than/20	Less than/5	Less than/5
6	Less than/20	200	Less than/5
7	Less than/5	56	Less than/2
10	Less than/2	Less than/1	Less than/1
11	Less than/2	3	Less than/1
12	Less than/2	240	Less than/1
13	Less than/2	Less than/1	Less than/1
14	Less than/2	Less than/1	Less than/1
16	Less than/2	230	Less than/1
17	Less than/20	625	Less than/10
18	Less than/15	960	Less than/10

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PAGE 4

Diethylabietamide

No standard material of diethylabietamide could be obtained commercially. As a reference material we prepared some abietamide at our laboratory. This was prepared from abietic acid by reaction with thionyl chloride and ammonia. Subsequent analysis by infrared spectrophotometry and mass spectrometry confirmed the material as abietamide. This material was used as a reference material in subsequent thin layer chromatography and gas chromatography.

The reserved solution from the amine analysis was extracted with chloroform to isolate any diethylabietamide which might be present in the samples. This extract was evaporated on a steam bath. The residue was then purified using thin layer chromatography. The system used was Analtech silica gel GHF plates 250 micron thick. The appropriate area of the plate was then scraped off the thin layer plate and eluted with chloroform for analysis by gas chromatography/mass spectrometry.

The extracts were analyzed using a GC/MS using the following conditions:

Column --------------------- 3 feet X 3 mm glass column with 3% OV 101 on Chrom W 80/100 mesh AW-DMCS

Carrier Gas ---------------- He at 30 cc/min

Temperature ---------------- 230°C

Results of these analyses were as follows: (ppm)

Sample No.	Diethylabietamide
1	Less/50
2	Less/50
3	Less/50
4	Less/50
5	Less/50
6	Less/50

APPENDIX F

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SEATTLE, WASHINGTON 98104

LABORATORY NO. 61768

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Sample No.	Diethylabietamide
7	Less/5
8	Less/5
9	Less/5
10	Less/5
11	Less/5
12	Less/5
13	Less/5
14	Less/5
16	*

*Due to excessive interferring material diethylabietamide could not be analyzed for in this sample.

Polychlorinated biphenyls

Analysis of these samples was performed by routine procedures which involved solvent extraction, solvent partition, and cleanup using florisil. The resulting extracts were then analyzed using gas chromatography/mass spectrometry utilizing the selective ion monitoring to eliminate interference. Results are calculated as Aroclor 1254.** Results were as follows: (ppm)

Sample No.	Polychlorinated Biphenyls
1	Less/.009
2	Less/.009
3	Less/.009
4	Less/.050
5	Less/.050
6	Less/.050
7	Less/.009
8	Less/.009
9	Less/.009
10	Less/.009
11	Less/.009
12	Less/.009
13	Less/.34
14	Less/.34
16	Less/10

**Arochlor 1254 is a tradename for a typical commercial PCB mixture.

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PAGE 6

Sample 15 was analyzed to determine its nature. Results of the analyses indicated the sample was essentially triethylamine with an oily residue amounting to about 1,000 parts per million. Analysis of this residue indicated it was essentially the same material as the oil fraction in sample 16.

Respectfully submitted,

LAUCKS TESTING LABORATORIES, INC.

G. F. Anderson

G. F. Anderson

GFA:mjt

Method #2 Calculation of Suspended Solids

Note: fr = fraction, equations in lbs/hr

(1) Total solids, lbs/hr = lbs/hr input sludge X fr. total solids
" = 265 (.177)
" = 46.9 lbs/hr

(2) Suspended solids in product water = lbs water/hr x S.S. fr.
" = 265 (.0033)
" = .87 lbs/hr.

(3) Total solids in dry cake = lbs cake/hr X T.S. fr
= 46.8 (.896)
= 41.9 lbs/hr.

(4) Oil as suspended solids = total solids x calc. oil fr
= 46.9 x .0107
= .50 lbs/hr

(5) Dissolved solids in dry cake = .292 of total dissolved solids, by ratio of:
water to dryer/water input

(6) Total dissolved solids, X = input total solids -(suspended solids) - (total cake solids - dissolved solids in cake)

X = (1)-(2)-(3)-(4)+.292X
X = 46.9-.87-41.9-.50+.292X
TDS = 5.13 lbs/hr

(7) Suspended Solids as fraction input solids = $\frac{(1)-(6)}{(1)}$

= .890

(8) Average Suspended Solids as fraction of input solids = .986+.890/2
" = .938

Method #1 Calculation of Suspended Solids

Ave. for 23 days
Input solids = 46.9 lbs/hr.

(1) Total dissolved solids (T.D.S.) = D.S. product water + D.S. in dryer cake

(2) Dissolved solids in dryer cake proportional to the ratio of water into dryer to water in feed, lbs/hr. or

$$63.6/218 = .292$$

(3) D.S. product water = % total solids -% suspended solids X $\frac{\text{lbs/hr P.W.}}{100}$

$$= (.52 - .33)\ (241.6)/100$$

$$= .46 \text{ lbs/hr.}$$

(4) D.S. in cake, X = $\frac{X}{.46+X} = .292$

D.S. in cake, X = .189 lbs/hr

(5) T.D.S. = .46 + .189 = .648 lbs/hr

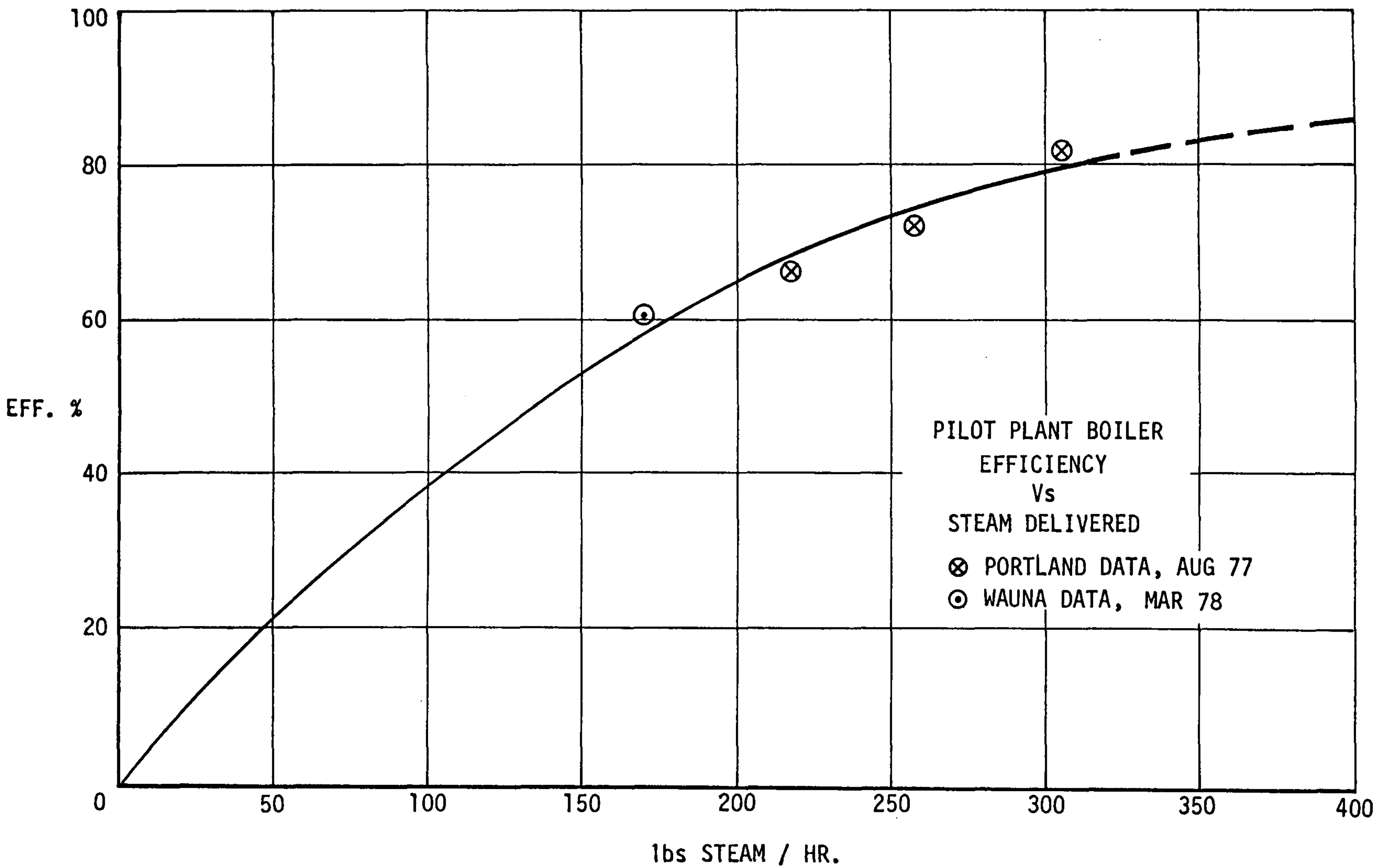
(6) Suspended solids = total solids - Dissolved Solids

= 46.9-.648

S.S = 46.25

(7) Suspended solids as fraction of input solids = 46.25/46.9

= .986


PILOT PLANT BOILER
EFFICIENCY
Vs
STEAM DELIVERED
⊗ PORTLAND DATA, AUG 77
⊙ WAUNA DATA, MAR 78
EFF. %
0
20
40
60
80
100
50
100
150
200
250
300
350
400
lbs STEAM / HR.

Calculation of Oil as Percent of Input Solids via Solvent Still Operation

March 10 through March 14

5 days

oil in TEA avg. = 1.35%

Total solids = 16.0%

Steam = 11.85#/hr-6.7#loss

Solids = 43.9#/hr.

X = fraction oil in input solids

wt(1) = 1/.16 = 6.25

svt(1) = 6.25(6) (6)/7 = 32.14

wt (11) = 32.14/(1-.0135-.04) = 36.0

wt(14) = X/.0135 = 74X
svt (14) = 74X(32.14)/36 = 66.0X

wt(29) = 66X(1.11) = 73.33X
wtr(29) = 73.33X-1.11

$\frac{\text{Steam/hr}}{\text{Solids/hr.}}$ = 73.33X-1.11 - 2.96X+140/(1150-210)

.117 = 73.33X-1.11-2.96X+.149

.117+1.11-.149 = +70.37X

1.078 = 70.37X

X = .0153 oil in solids = 1.53%

Appendix J

Calculation of Oil as Percent of Input Solids via Solvent Still Operation

February 24 through February 28

5 days

NOTE: For sta. locations refer to Fig. 8-3

Total solids input sludge = 19.6%
Steam = 14.72 #/hr.-6.7 loss = 8.02#/hr theor.
dry solids = 36.2#/hr.
steam/# solids = .407
recycle teA % oil = .75%
X = fraction oil input solids
wt (1) = 1/.196 = 5.10
svt (11) = 5.1 (6.5) 6/7 = 28.4
wt (11) = 28.4 /(1-.0075-.04) = 29.83
wt (14) = X/.0075 = 133X
svt (14) = 133X (28.4)/29.83 = 126.94X
N1 = 126.94X
wt (29) = 126.94X (1.11) = 140.91X
wtr (29) = 140.91X-1.11
wtr(15) = wt(14) C(74) = 5.32X

$$\frac{\text{Steam}}{\text{Solids}} = \frac{8.02}{36.2} = 140.91X - 1.11 - 5.32X + .205$$

$$.22 = 135.59X - .905$$

$$1.126 = 135.59X$$

$$X = .0083 = .83\% \text{ oil in solids.}$$

Resources Conservation Company
Renton, Washington
Project 500501

CROWN ZELLERBACH, WAUNA MILL DRIED SLUDGE

KDL#	1150-M
Proximate, Wt. %	as rec'd
Moisture	14.4%
Volatile Matter	63.9
Fixed Carbon	11.7
Ash	10.0
TOTAL	100.0%
HHV, Btu/Lb.	6000
Ultimate;	
Moisture	14.4%
Hydrogen	4.6 5.37 dry basis
Carbon	35.5
Sulfur	0.2
Nitrogen	0.7
Oxygen	34.6
Ash	10.0
TOTAL	100.0%
Ash Fusibility; ASTM D-271-48	
I.T. °F	2210 initial deformation temp.
S.T.	2270 softening temp.
H.T.	2340 hemispherical temp.
F.T.	2530 fluid temp.
Ash Composition;	
SiO_2	50.3%
Al_2O_3	28.7
Fe_2O_3	8.5
CaO	3.8
MgO	1.8
Na_2O	2.9
K_2O	0.6
TiO_2	1.7
P_2O_5	0.1
SO_3	1.5
TOTAL	99.9%

David Vaughan
David E. Vaughan
March 7, 1978

THE BOEING COMPANY.

LABORATORY REPORT

NO. 2-4809-0001-258

Purpose ______ Model ______ Date July 20, 1978

To: Jerry Bannon Org'n. ______ Part No. ______ Appendix L

Subject: Ash Samples

Source ______ Reinsp. Req. ______

Purchase Order ______ R.R. ______ Date Rec'd. ______ Quan. ______ Acc. ______ Rej. ______

Material ______ Spec. ______

☐ Chem. Lab. ______ ☐ Sonic ______ ☐ Met. Lab. ______ ☐ Mechanical ______

☐ X-Ray ______ ☐ Mag/Penetrant ______ ☒ Particle Identification Laboratory

Reference: C.C. to:

Two samples of ash were received for analysis. These were labeled "Ash Front Sight Port" and "Ash Back Section". These samples were examined using petrographic techniques.

Results: The material consisted largely of fused mixtures of glass similar to the ash from coal. The pyrolysis products of calcite ($CaCO_3$), clays (complex of Ca,Na,K, Al,Mg,Fe,Si,Al,O,H), alumina (Al_2O_3), quartz (SiO_2), and rutile (TiO_2) were identifiable in the ash as separate entities as well as in combination. Iron was typically part of a mixed particle imparting an orange to red color and a higher refractive index to the particle. Some quartz was present in the sample as irregular particles with rounded edges as well as partially crystallized spheres. This indicates that some of the quartz is present in cooler areas of the burn. The presence of the individual pyrolysis products of calcite, clays, alumina, quartz, and rutile indicates local concentrations and a minimum of mixing during combustion. This is supported by the grossly segregated appearance of the "Ash Back Section" sample. The elements from the ash are distributed as follows:

Probable Source	Free Oxides	Mixed Oxides
quartz, clays	Si	Si
alumina, clays	Al	Al
process		Fe
carbonates, clays	Ca & Mg	Ca & Mg
process, clays		Na
clays		K
rutile	Ti	Ti
--		P
process		S

An additional source of these elements is the residue from the combusted organic portion of the material.

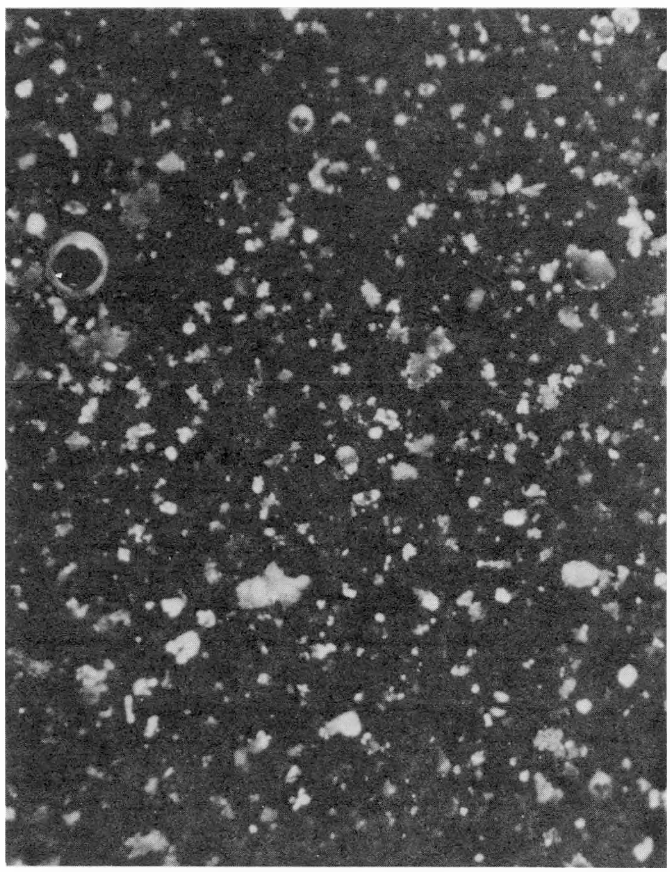
Prepared by E. R. Crutcher
E. R. Crutcher
Approved by R. P. Bossler
R. P. Bossler
Org'n. 2-4809

D6 6000 2100 ORIG.3/71

RCC: Ash - Front Sight Port

Magnification 71x

Crossed Polars



Thermal Cost of Vaporizing Sludge Cake Water in a Hog Fuel Incinerator

There is general agreement in the pulp and paper industry that the "breakeven point" for burning sludge cake in a hog fuel incinerator is somewhere between 30 and 40% solids,[(1)] i.e., the heat generated by the solids is enough to vaporize the associated water, no additional fuel being required. The thermal cost for vaporizing the water is calculated as follows.

$$\frac{\text{lbs water}}{\text{lb solids}} \text{ @ 30\% solids} = \frac{1}{.3} - 1 = 2.3$$

$$\frac{\text{lbs water}}{\text{lb solids}} \text{ @ 35\% solids} = \frac{1}{.35} - 1 = 1.86$$

From the fuel value results of sec. 7.0, the solids were shown to have a low heating value of 6335 Btu/lb solids; therefore the range of thermal cost for vaporizing the water is,

Btu/Lb H_2O (30% solids) = 6335/2.33 = 2718

Btu/Lb H_2O (35% solids) = 6335/1.86 = 3406

Avg. = 3062 $\frac{\text{Btu}}{\text{Lb } H_2O}$

For calculation say 3000 $\frac{\text{Btu}}{\text{Lb}H_2O}$

(1) personal communication with members of the engineering staff of ITT Rayonier, Hoquium, Washington, Crown Zellerbach, Camas, Washington and Longview Fiber, Longview, Washington