A METHOD OF MANURE DISPOSAL
FOR A BEEF PACKING OPERATION
First Interim Technical Report

by

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Project 12060 EOF

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ABSTRACT

This report contains the preliminary studies, process development, process calculations, and process design for a system to successfully handle the paunch manure in a beef slaughtering operation.

These studies resulted in a system in which the paunch manure is collected from the slaughtering operation and is fed to a screening device which separates the coarse solids. The screenings are dewatered to a solids content of 37 percent. This dewatered material is then sent to a fluid bed incinerator via a screw conveyor. The liquid stream from the screening is fed to a settler and is combined with the liquid stream from the dewaterer prior to sand filtering. The filter cake is fed to the incinerator using a screw conveyor. The filtrate is recycled back to the settler.

Work was completed as of February 1971.

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SECTION I

INTRODUCTION

This report is concerned with the preconstruction research and development activities under an FWQA Research Development and Demonstration Grant, Project Number 12060 EOF, to the Illinois Packing Co., Chicago, Ill. The objective of the project is to develop, design, install and demonstrate a fluidized bed incineration system for the efficient disposal of paunch manure waste generated in beef packing operations. The process, shown schematically in Figure 1, is being engineered and fabricated by Procedyne Corporation for installation at the Illinois Packing Co., Chicago, Ill.

Paunch Manure is partially digested feed material removed from the stomach of cattle during preparation for market. It is the only major portion of the animal with no practical commercial value. Disposal is costly -- approximately $12/ton when hauled from the premises for dumping. In addition, a significant portion of the solid waste is disposed as sewage and adds a substantial burden to municipal sewage disposal facilities.

The USDA statistical report on commercial slaughter for the United States indicates that a total of 35,026,400 head were slaughtered in 1968. When related to the paunch manure disposal problem of approximately 50 lbs/stock (including

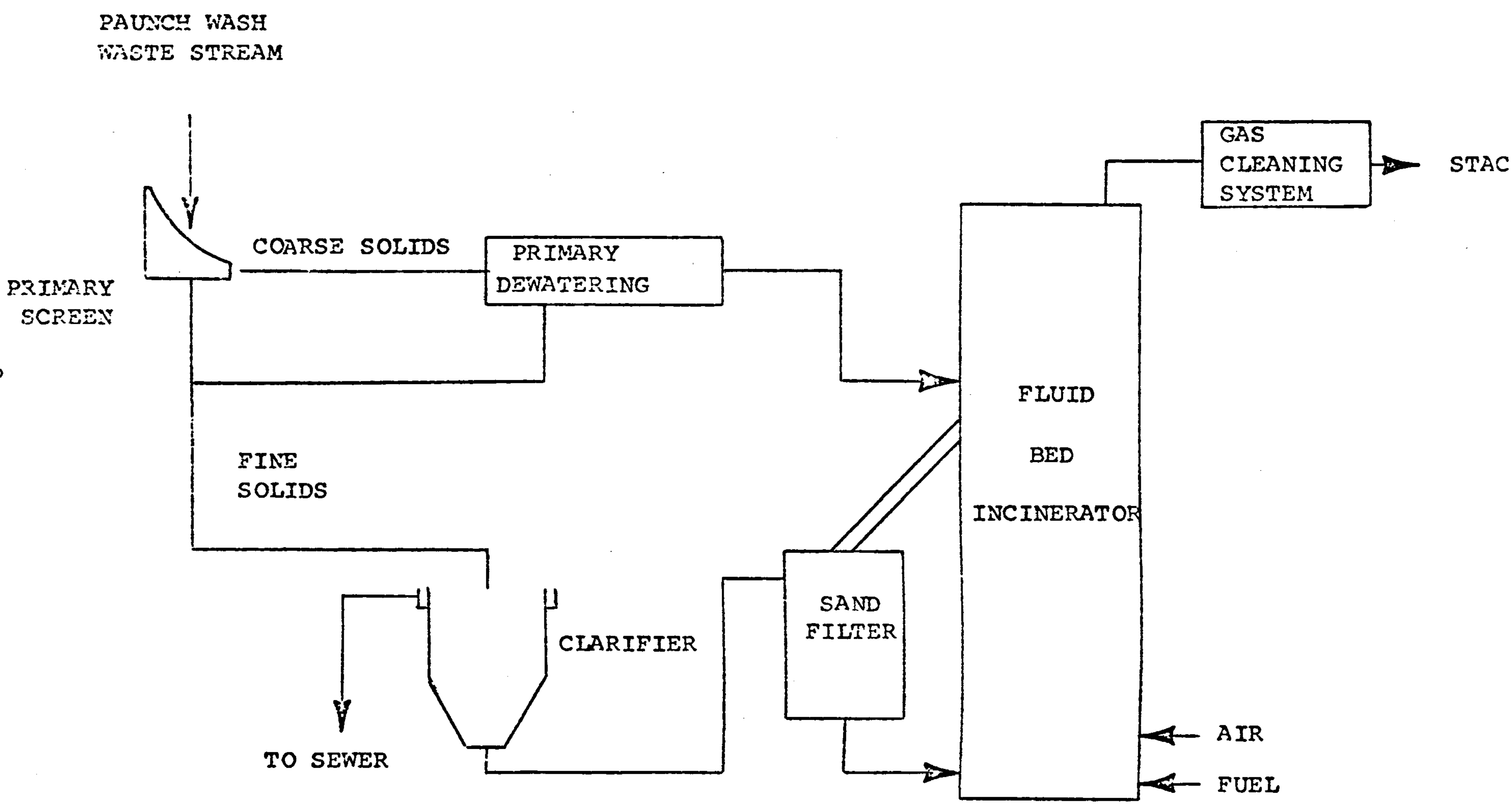


FIGURE 1. PAUNCH MANURE DISPOSAL PROCESS

sack waters), it is seen that 1,751,320,000 lbs. of waste that must somehow be disposed at minimal cost and with maximum consideration to the reduction of the pollution problems resulting from present disposal practice. The same statistical data indicates that the 1968 slaughter for the state of Illinois was 1,407,000 head; thus 70,350,00 pounds of paunch manure.

Paunch manure is untreatable in conventional sewage treatment plants (1) for the following reasons:

1. It has a very high biochemical oxygen demand.
2. Its high solids content tends to mat into masses which clog bar screens.
3. It settles out in conventional tanks and in time hardens to the consistency of low density rock.
4. It clogs hopper bottoms, pits and pump suctions.
5. It sets up like concrete in pipe lines, requiring augering for removal.
6. The cellulose material will not decompose in digesters and forms straw blankets which clog and eventually fill digesters.
7. The entrapped moisture in the cellular material can not be dewatered by vacuum filters.
8. The material has an objectionable odor which rapidly decomposes into an intolerable stench.
9. Because of its cellular type moisture it cannot be dried in a flash dryer nor can it be burned in suspension.

For these major reasons and many minor ones, no community treats paunch manure in sewage treatment plants but disposes of it separately, usually by land fill.

The research and development activities described in this report have culminated in a final process design which is currently being engineered for construction. A summary of the process design is presented in the last section of this report, Section V.

SECTION II

PRELIMINARY STUDIES

Paunch Manure Characteristics

The quantity and physical-chemical composition of paunch manure are dependent upon the composition of the animal feed and the environment in which the animals are held during the processing period. Due to this fact, it is difficult to relate existing literature data to the present problem. Furthermore, data obtained in the past do not accurately reflect present conditions because of changing feed practices (2). Physical and chemical data as required for the present process development were obtained as presented in this section.

The quantity of paunch manure was established in terms of pounds of wet paunch waste per animal. A number of measurements on fresh paunch sacks shows that the average weight of the paunch contents is approximately 55 lbs./ animal. That number compares to the typical value of 50 lbs/animal reported in the literature.

Simultaneous research conducted by the Federal Water Quality Office Robert S. Kerr Research Center produced physical and chemical data on dried paunch material.

BOD analysis on the material established the ultimate first stage BOD of the soluble material to be 84,000 mg/l of paunch content and that of the nonsoluble fraction to be

24,000 mg/l of paunch content. These data suggest that approximately 80% of the total BOD in a paunch wash stream will be in the form of dissolved paunch solids. Composition of the dried paunch material was reported as follows:

Dehydrated paunch	Average % of 10 samples
Moisture	15.3
Protein	10.3
Ash	6.7
Fat	4.4
Calcium	0.5
P_2O_5	1.4
Crude Fiber	21.2
Carbohydrates	42.0**

** Calculated by subtracting the total percentage of moisture, protein, ash, fat, and crude fiber from 100%.

For purposes of material balance calculations the combustible fraction of the paunch material was assumed to be cellulose with the molecular formula($C_6H_{10}O_5$).

In order to establish data for heat balance calculations, calorimetric measurements on dried paunch material were made with Parr Oxygen Bomb Calorimeter. Data on two samples produced heat of combustion, ΔH, values of 3890 and 4120 Cal/gm for an average value of approximately 4000 cal/gm or 7200 Btu/lb.

The data presented above adequately establish the chemical and thermodynamic properties of the paunch material required for process design purposes. Physical and chemical characteristics pertinent to the paunch disposal operation and unit operations in the disposal system under design are presented in sections which follow.

Present Disposal System

In the present paunch handling system the unbroken paunch sacks are conveyed to a table where they are carefully washed and any excess fat is manually trimmed before emptying as shown in Figure 2. The empty paunch sacks are then thoroughly washed and sent on for packaging. All three streams, the paunch sack preliminary wash, the paunch manure slurry, and the paunch final wash are combined and fed through a trough to one of two paunch manure drain bins as shown in Figure 3. The solids content of that manure slurry has been estimated to be 5% solids by weight.

The paunch manure is settled periodically in each of the drain bins during the day shift by alternating the feed between the two bins. The bins are unloaded each evening into a disposal truck which carts the material to a farm outside of the City of Chicago to be spread on the ground. The filtrate from the drain bins is discharged at a rate of approximately 50-60 gpm into a plant sanitary sewer line

Paunch Processing Area
Illinois Packing Co.

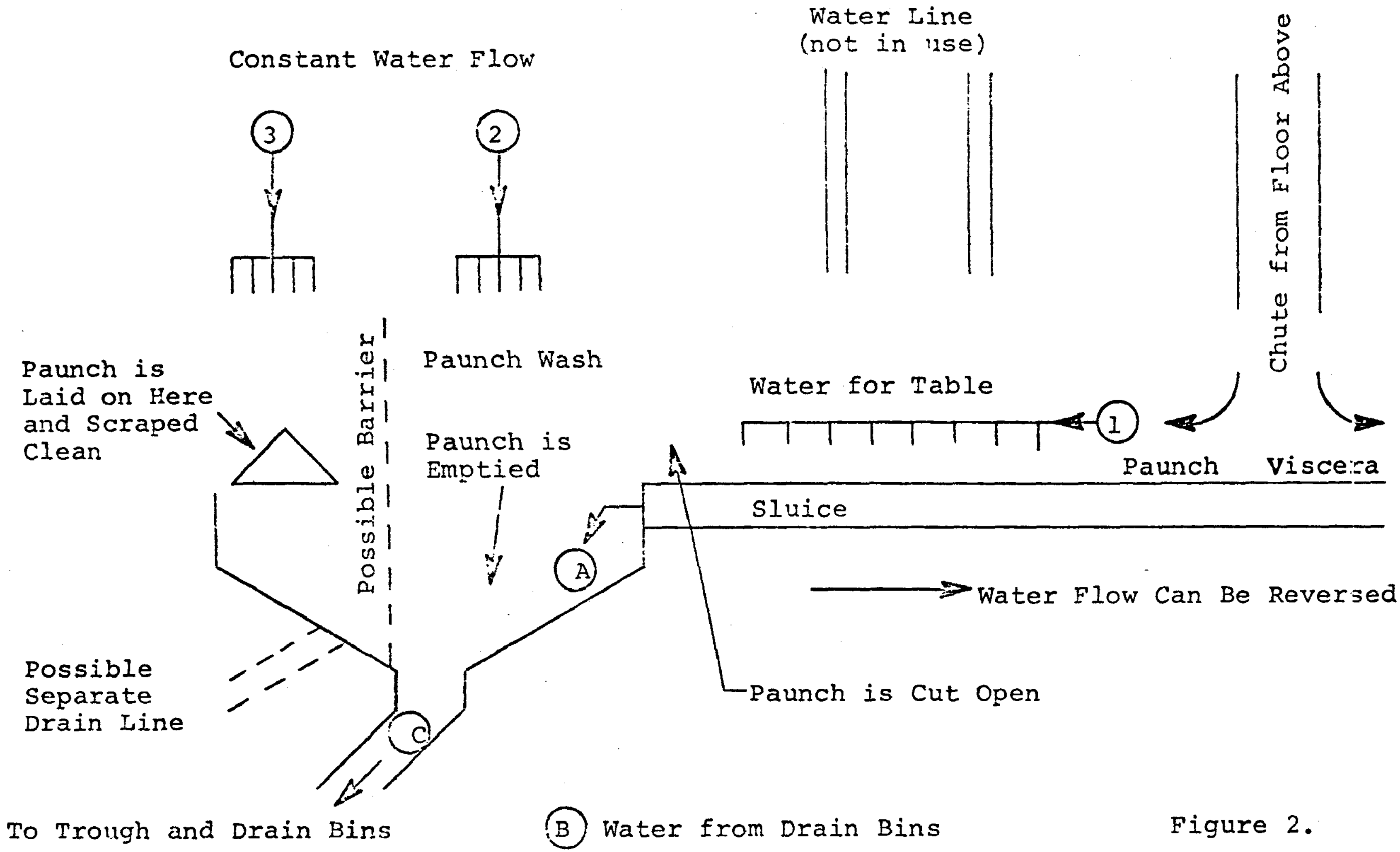


Figure 2.

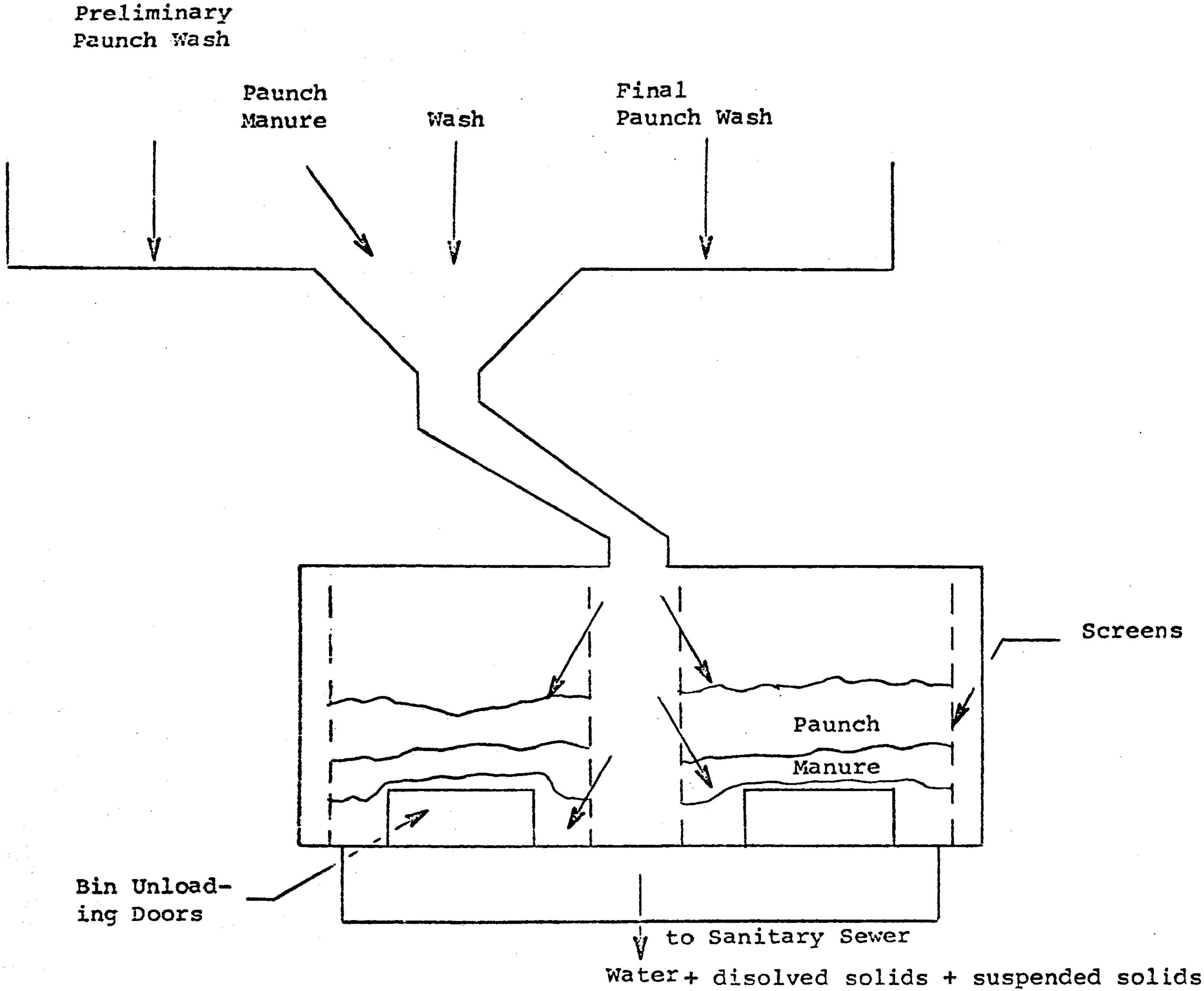


FIGURE 3. PRESENT PAUNCH MANURE HANDLING OPERATION

which runs from the plant to a Chicago municipal sewer line and then to the Chicago Sanitary District Treatment Plant.

Several field studies were made of the operation of the paunch table under actual operating conditions. Flowrates of water were measured and these are shown in Figure 4. Original estimates of paunch manure weights were checked and it was found that the contents of 14 sacks weighed an average of 55 lbs. each. Taking the stated maximum capacity at the paunch table of 95 sacks/hour, the material balance at the table can be summarized in the following block diagram Figure 5.

The concentration of solids in paunch manure was determined at random intervals throughout the preconstruction study period. Examples of results of these laboratory tests are presented in Table I.

TABLE I

PAUNCH SOLIDS CONCENTRATION

Sample Number	Date	% Total Solids
1 Paunch & Fluid	1/30/70	14
2 Paunch	6/15/70	17.0
3 Paunch	10/13/70	18.5
4 Paunch	10/22/70	18.4

Paunch Processing Area
Illinois Packing Co.

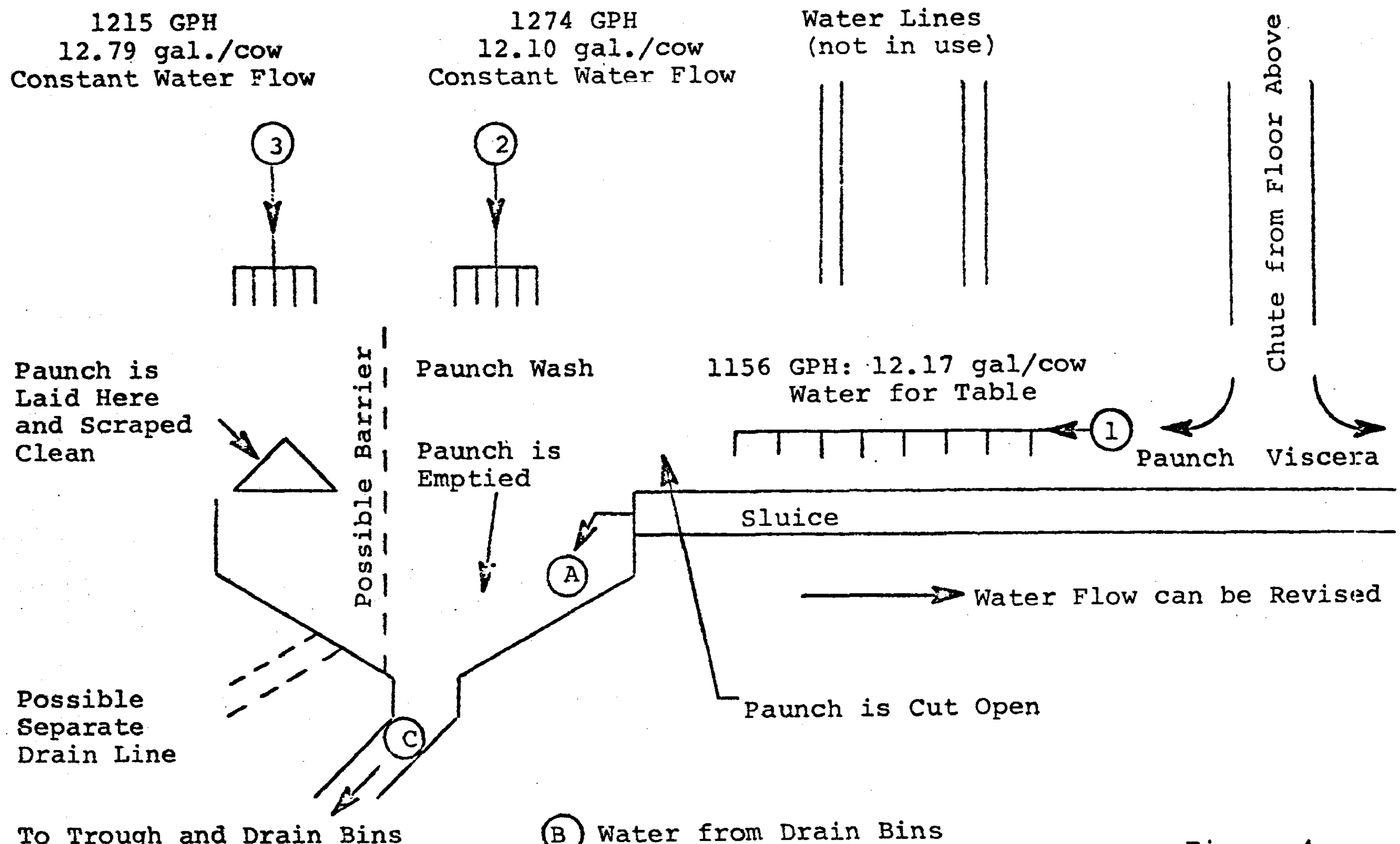


Figure 4.

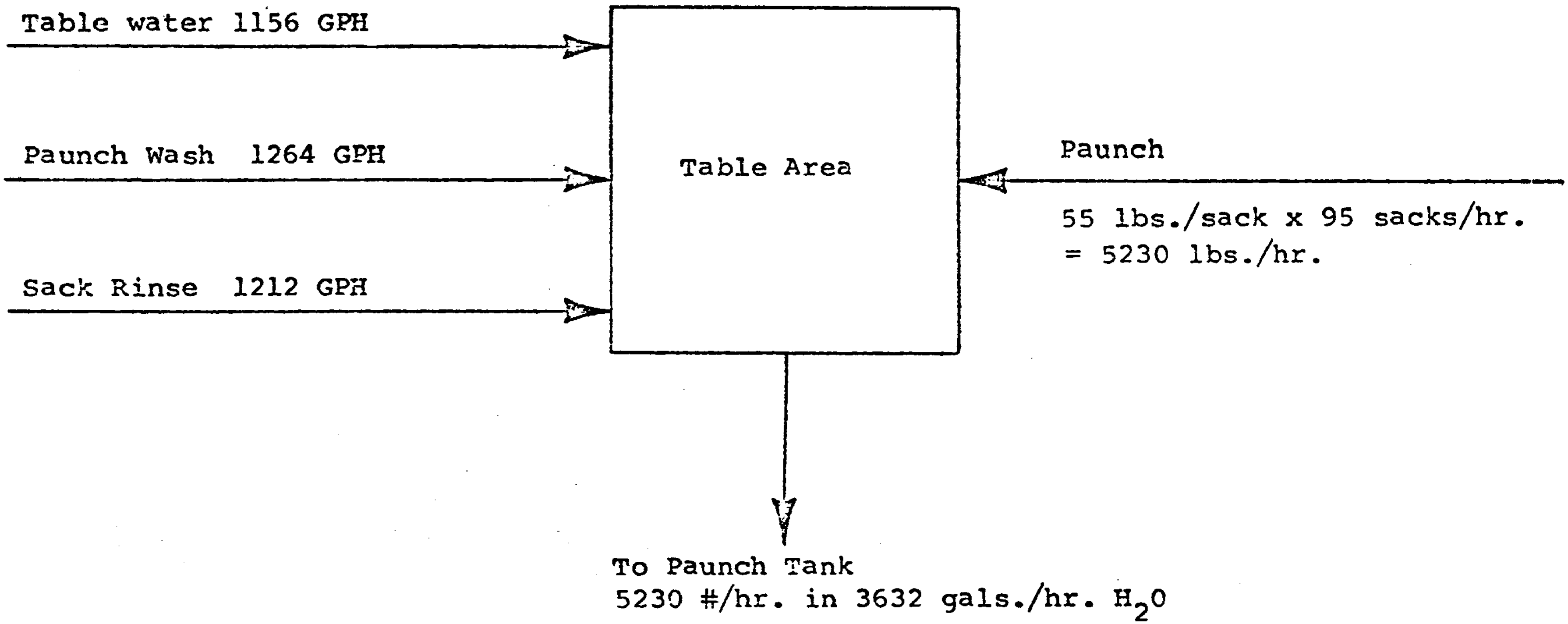


FIGURE 5. MATERIAL BALANCE FOR PAUNCH WASH OPERATION

The first result represents a sample of manure and sack fluids, the others a sample of free drained paunch manure. Samples of the waste paunch stream running to the sewer were taken monthly by Illinois Packing personnel during the four month period - April to July 1970 and these samples were sent to Pollution Control Laboratories Inc. Chicago, Ill. Results obtained are shown in Table II.

TABLE II

PAUNCH WASTE STREAM CHARACTERISTICS

Parameter	Date (1970) April 1	May 2	June 3	July 4
BOD (mg/ l)	8,353	5,693	2,990	10,500
COD (mg/ l)	16,334	7,108	11,178	12,534
DO (mg/ l)	0.1	0.1	0.1	0.1
Total Solids (mg/ l)	18,064	8,274	4,244	2,361
pH	7.40	7.25	6.20	6.55
Nitrogen, Kjedhal (mg/l)	238	463	259	291

Except for supplying data regarding ranges of efficient concentrations, the above information was of little value in further defining plant design parameters. Because of the fluctuating stream concentration found, no further sampling was deemed useful at the time. A scheduled, periodic sampling program will be initiated six to eight weeks prior to the disposal plant start-up in order to establish base line measurements for system evaluation purposes.

Paunch Manure Filtration Studies

The first experiments with the filtration of paunch manure and water mixtures were made to develop compressibility data for paunch manure on sand beds using classical techniques (3). These involve a determination of filtration rate, filtration volume, and pressure drop, and then calculating specific cake resistance at several levels of pressure. The log of cake resistance is plotted against log of P.; the slope of the curve is designated as the compressibility S of the cake. For paunch manure, which produced a mat-like cake because of the presence of many straw like particles, the compressibility S, was found to be 0.9 (vs. 1.0 max.). A second group of experiments involved batch filtration in a 3 1/2" Plexiglass column with varying static heads of water above the beds. Results are shown graphically in Figures 6 and 7 for two conditions of sand and manure in layers and sand and manure mixed. Filtration rates were better for mixed material but quite low for both cases. Several different fluidizing materials are used by Procedyne in its fluid bed systems. For incineration of waste materials sand is the economical choice. Two specific sands were used in the project and these are designated as Sand MG and Sand P.

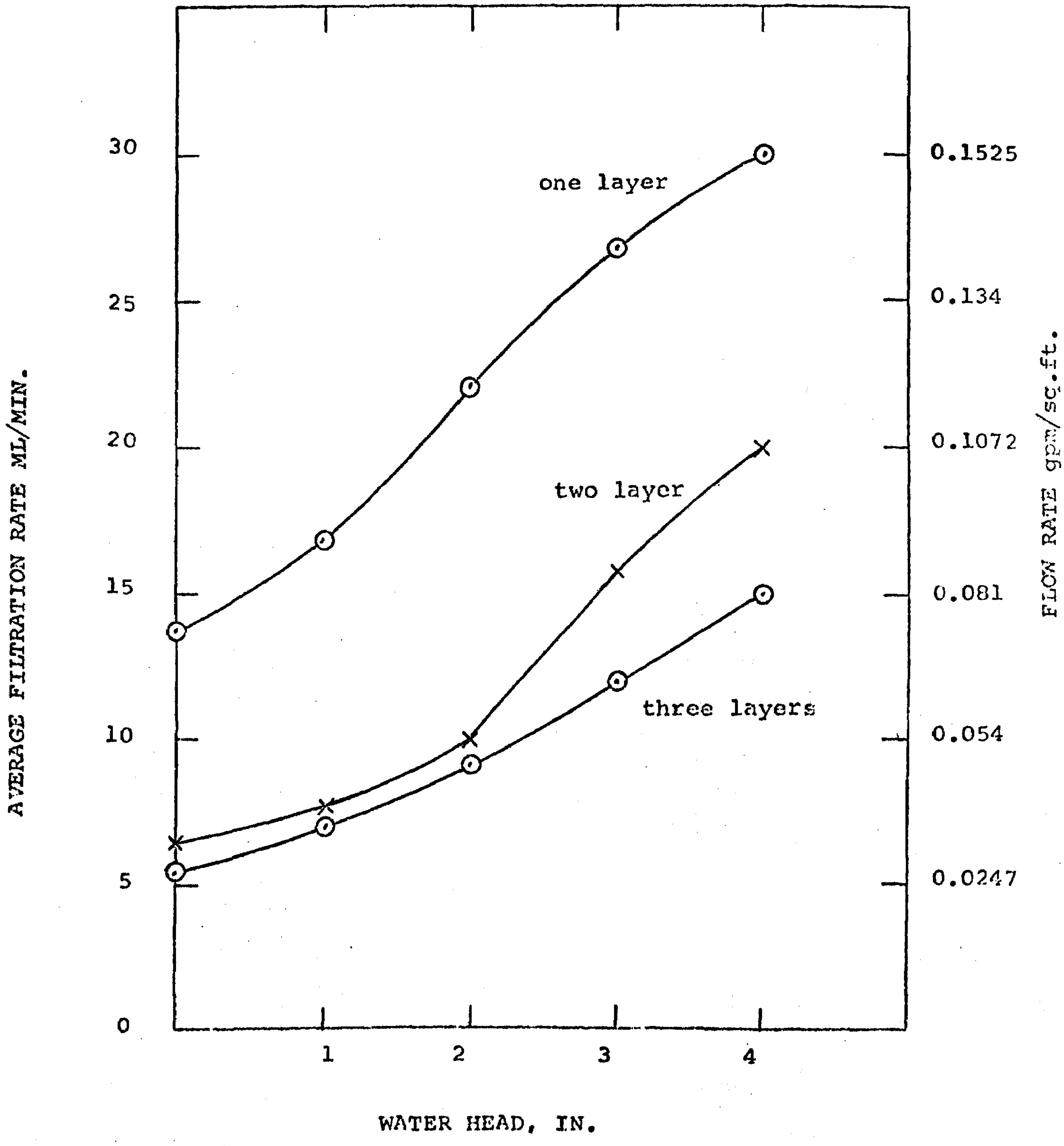


FIGURE 6. FILTRATION THROUGH SAND AND MANURE IN LAYERS

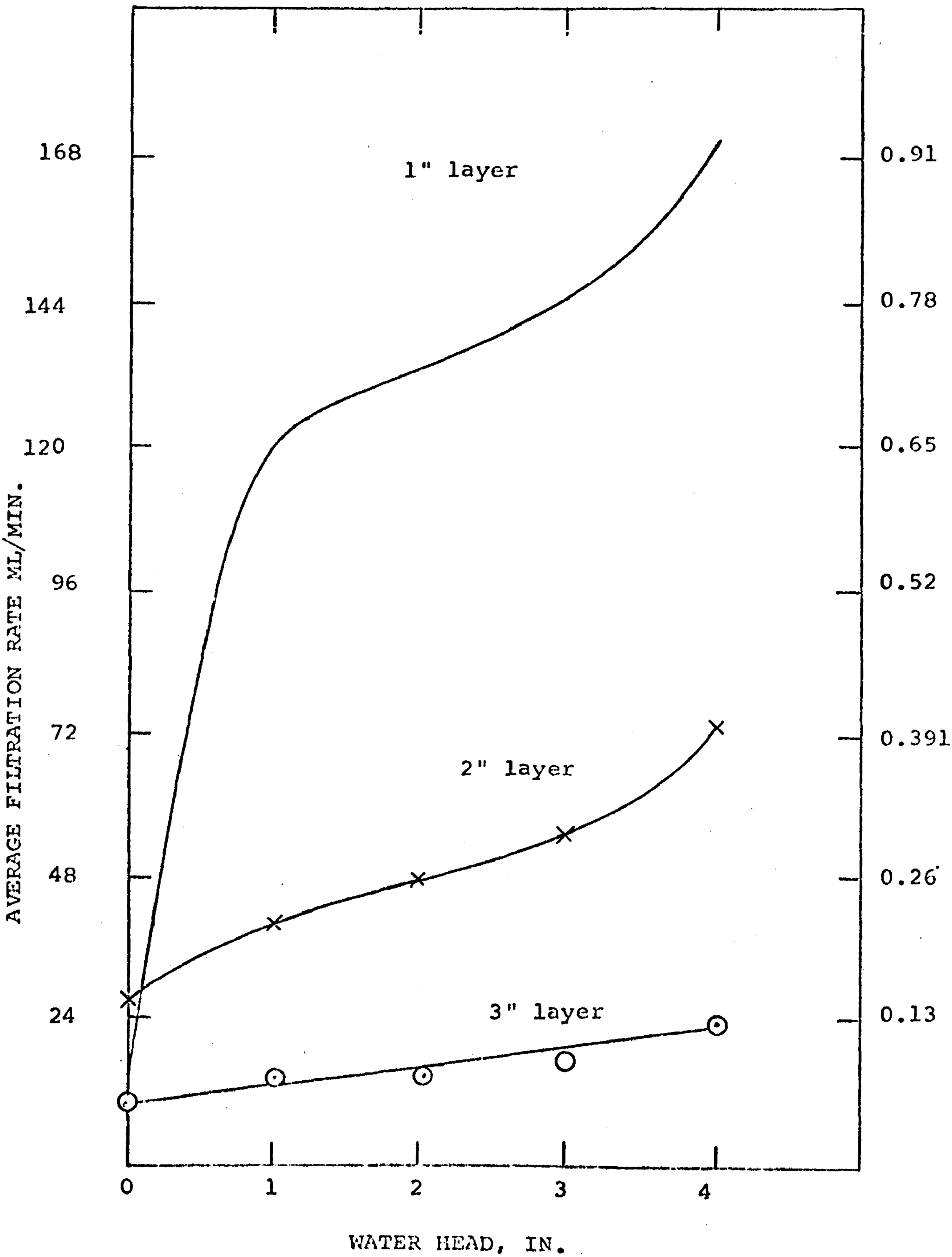


FIGURE 7. FILTRATION THROUGH SAND AND MANURE MIXTURE

Approximate screen size distribution of these sands is as follows:

	Sand MG (%)	Sand P (%)
+ 30 m.	1	25
-30 + 50 m.	35	60
-50 + 80 "	30	10
- 80 "	34	5

The coarse sand P is somewhat purer and is more stable as a recycle sand when quenched in the filter circuit. It was the sand used for the subsequent filtration experiments. In order to further investigate the filtration properties of paunch manure-sand mixes, a third program of a group of 15 fixed bed experiments were performed. These experiments were performed in 3" and 8" cylindrical chambers made of both steel and plexiglass depending on the need for either pressure or visibility in each experiment.
As noted previously, various samples of paunch manure were used throughout the entire experimental program and these varied in concentration from 14-18%. Materials used in Experiments 1 and 2 were from a 14% lot. Materials used for the other 13 experiments were from a 17.1% lot. Sand to manure ratios were held at 4:1, a value which permits a reasonably economical operation of the sand recycle system. The material itself was first filtered on a Buchner funnel

using laboratory filter paper (repeating a procedure developed during the first group of experiments) in order to establish a basis for comparison of results when using sand as a filter medium. Results of those tests are shown in Table III.

TABLE III

FILTRATION TEST RESULTS

Expt. No.	Technique	Filtrate Results	Other Comments
A-1	Filtration through Laboratory Buchner	TS(Total Sol.)=5360 mg/liter TOC(tot. Organic Carbon) = 3000 ppm	Eaton-Dikeman Filter Paper grade 512
B-1	Same as A-1 except sand on an 80 mesh screen is the filter bed-3" 0 column, 3" deep bed.	TS(1st. 15 sec.)=1300 TS(after 15 sec.)= 7330 TOC " " " = 5900	Avg. filtration rate 0.91 gpm/sqf Rapid blinding after 1st 15 sec. used to wet bed 84% moisture in cake.
B-5	Simulation of deep bed multilayer concept-3"0; containing three, 4" deep sand manure layers. Filtrate removed by free draining	TS - 6910 TOC - 5540	Avg. filtration rate- .115 gpm/ sq.ft.
B-6	Same as B-5 except 5 psig pressure use on filter	TS-8870 TOC-6900	Avg. filtration rate- .837 gpm/sq.ft.
B-7	Same as B-5 except each 4" layer of manure sand was well mixed before adding to filter. Pressure was increased to 10 psig.	TS-7350 TOC-6060	Filtration rate average 1.5 gpm/sq.ft.

TABLE III (Cont.)

FILTRATION TEST RESULTS

Expt. No.	Technique	Filtrate Results	Other Comments
B-8	Same as B-1 except the 80 mesh screen is replaced by a sample of porous stainless steel belt.	TS-7310 TOC-5530	
B-12	Same as B-5 except one manure-sand layer followed by one sand layer; no pressure	TS-7230	Rate-.363 gpm/sq.ft.
B-13	Same as B-12 except 10 psig pressure	TS-7650	Rate-1.74 gpm/sq.ft. % moisture in cake-81%

The following conclusions were drawn from the above series of experiments:--

(a) The basic filtration quality through lab filter paper on a total solids basis (TS), was 5360 mg/liter and total organic carbon (TOC) was 3000 ppm. This establishes that there will be present in the feed stock to the process, a fraction containing certain minimum dissolved or colloidal volatile solids. These cannot be removed by normal filtration techniques. The concentration of these dissolved and colloidal solids will vary depending on several factors e.g. diet of the cow, length of time in slaughtering process, age of paunch manure before processing etc.

(b) Although several different kinds of experiments were made using sand as a filter media it was noted that total solids did not increase much beyond 7000 to 8000 mg/liter for any of these experiments.

(c) Batch filtration rates are improved to some extent with pressure (or vacuum) from under 0.2 gpm/sq. ft. filter area to approx. 1.75 gpm/sq. ft. at pressures of 10 psig.

(d) Cake moisture content could not be decreased to below 81% (or 19% solids) on a sand free basis during this series of experiments. This value is on the low side for antogenic incineration when compared with an acceptable level of 25-30%.

(e) The most general conclusion reached was that the filtration rates for paunch manure were disappointing; the dewatering characteristics were likewise. These results confirmed the work done on compressibility during the earlier testing program.

At this point in the process development program it became necessary to change the process which had a single dewatering stream to one which would have two:-

(1) A major dewatering stream-which would process the fibrous and larger particles of paunch manure to higher than 35% moisture levels.

(2) A minor dewatering stream consisting of the finer paunch manure particles which would be processed through the moving bed sand filter.

SECTION III

PROCESS DEVELOPMENT

DEWATERING

Preliminary work in the laboratory was directed towards the investigation to the extent of dewatering that could be attained by the following treatments:

1) Simple dewatering
2) Pressure dewatering

The preliminary filtration studies indicated that only the unbound water could be removed. These tests were carried out using a simple screen and in later cases vacuum filtration. The maximum attainable solids content was 18.5%. Pressure dewatering tests were conducted using hand operated rolls, of the type used for rolling metal in metal fabrication shops. Three tests were run:

1- Paunch Manure containing 18% solids was put on a screen and rolled. The solids concentration of the rolled paunch manure was increased to 23%. The solids concentration of the liquid extracted was 4%.

2- Paunch Manure (18% solids) was again put on a screen, but this time was rolled and recycled continuously through the rollers until no further liquid could be extracted. The solids content of the material from this test was 27%.

3- The third experiment was conducted to test the effect of sand as a dewatering aid. 1/2 part of sand was added to 1 part paunch manure and processed through the rollers. The solids concentration of the sand-manure mix was 54%. Correcting for the sand, the solids concentration of the paunch manure was 23%.

These tests (1 & 2) indicate that the solids content of the paunch manure can be raised to approach the autogenic point and thus economic operation.

Primary Dewatering

In the development of the flow sheet for this process the need for simple (primary) dewatering became apparent. This step was required to:

(a) Remove free draining liquid to reduce the load on main dewatering equipment.

(b) Provide a constant feed stream to the main dewatering equipment i.e. to eliminate any effect of large water inputs upstream of the dewatering system.

Preliminary investigation into equipment for this service indicated that it could be broken down into two broad categories a) simple screens and b) vibrating screens. The use of either of these two approaches for this duty is classical in the chemical and mineral processing industries. Simple screens are not normally used in dilute slurry processing service. Their use is normally restricted to the handling of dry or near-dry solids. However, recent developments in this equipment have been made by companies active in the meat packing, domestic sewage, and pulp and paper industries. These screens, set at predetermined varying angles, using patented screen configurations, have the ability to dewater low solid content slurries and discharge the dewatered solids without clogging the screens. Screens of this type are the Bauer Bros. Inc. "Hydrosieve" and the Dorr-Oliver "DSM Screen".

The second category of screen available is the vibrating type. There are many of this type available, the vibration to the screen being induced electro-magnetically, by unbalanced fly-wheels or eccentric shafts. In this type of machine, the vibration induced to the screen ensures discharge of the solids deposited on it.

In an effort to provide a plant whose operation would be economic as well as simple, and based on the experience of screen manufacturers in dewatering paunch manure, the angle screen described above was given primary consideration. The Hatfield Packing Co., Hatfield, Pa. and Wilson Packing Co., Cedar Rapids, Iowa were visited to obser this equipment in operation. Those operations supported manufacturers claims that paunch manure could be dewatered to 18% solids on this equipment.

Mechanical Dewatering

As described elsewhere in this report, the solids in paunch manure are a mixture of hay, straw, grain and corn. These materials are fibrous and cellular in nature and contain, when dewatered of all surface moisture, approximately 80% water.

The laboratory investigation conducted on this material has shown that mechanical dewatering would be necessary in order to reduce the quantity of water in this material.

The investigation of the types of equipment available to dewater this material uncovered the following:

1- The substantial work done previously in the dewatering of this material was discouraging in that only one device could be found to dewater this material to the extent required.

2- No information was available as to the quality of the liquid stream exiting from this dewatering device. The solids concentration in this stream was not available from previous tests.

3- The advancements made in dewatering equipment spurred by recent interest in ecology would require investigation for applicability to paunch manure dewatering.

4- Developmental work would be required to provide a machine suitable for paunch manure dewatering.

5- Due to the inexperience of some equipment manufacturers on paunch manure, actual operating tests would be required to judge the suitability of the various machines offered.

Equipment Investigated

The following types of dewatering equipment were investigated:

1- Screw Presses

2- Disc Presses

3- Rollers

4- Miscellaneous

Screw Presses

a) Conventional Design

This press is essentially a screw conveyor with the cross-sectional area decreasing with the length of the conveyor. This reduction is accomplished by either changing pitch of the screw or shaft diameter. The exit is usually restricted so that the material to be discharged is further subjected to compression, usually by means of a pressurized cone.

Manufacturers of this type of equipment were contacted and invited to quote on this application. Two declined to quote and those quotations that were received indicated that this type of equipment is generally more expensive than the other types of equipment available.

It has been reported(1) that paunch manure particles, after compression, tend to swell. This action in a screw press tends to overload the drive, causing the machine to stall.

A field test was conducted on a screw press and the machine failed before any dewatered paunch manure was discharged. Failure was due to an overloaded motor.

Most screw presses are of the conventional design described above. The quotations received showed them to be expensive, no doubt due to the heavy duty drives and motors required to accomplish dewatering of this material.

Improved Screw Press

An improved design of screw press has recently been made available. It improves on the basic design of a conventional screw press in that it incorporates an expansion zone to compensate for swelling after compression. This reduces the drive and motor requirement.

A device of this type was tested on paunch manure (18% solids) The test proceeded smoothly with the paunch manure dewatered to 38 weight percent solids. The liquid underflow stream contained 3.5% solids.

Disc Presses

During the search for suitable equipment for the dewatering of paunch manure, it was found that extensive testing had been done with this type of equipment.

This press consists of a pair of inclined screened discs that rotate very slowly. The paunch manure is fed in the top and is squeezed as it passes through the reducing area caused by the incline of the discs.

The manufacturer reported that paunch manure could be dewatered to 40% solids. Subsequent investigation in an effort to arrange for field testing showed that 1) presses were not being used for paunch manure dewatering at the present time 2) presses were not available on which to test the material 3) data for the solids content of the liquid stream were not available.

Rollers

The application of rollers for dewatering operations is old. It is the classical method for the removal of syrup from sugar cane. Despite this, most roller manufacturers are not inclined to quote on, or are disinterested in the application of roller machinery to dewatering applications. Although five manufacturers of roller equipment were contacted, only one was interested in this application.

A developmental program was initiated and executed for four months. After a discouraging start, machine development has progressed to the point where the solids stream from the rollers is 37%. This has been confirmed in two tests.

The liquid discharge from this equipment contains 6% solids, the highest concentration of solids in the liquid stream of any of the machines tested. Further development no doubt could decrease this concentration of solids.

Miscellaneous

Investigation into the various types of dewatering equipment available uncovered several miscellaneous machines available for dewatering service.

a) Roller Type Hydraulic Press

This machine consisted of two rolls separated by a floating ring. It had been used primarily in the wood and pulp industry. Its main disadvantage was the high hydraulic pressures required for its operation. This machine has been out of production for some time.

b) Multiple Roller Presses

These units are a fairly new development finding application in the sanitary field. It consists of a porous belt that travels through several stages of compression and shear rolls. The material to be dewatered is placed on the belt and travels through the rolls. Sufficient information was not available nor units available for field testing with paunch manure.

Summary

The actual operating tests run in the field on paunch manure showed that this material could be dewatered to a solids content approaching 40%. This was accomplished on both the three roll mill and the newer design of screw presses.

Manufacturers of other equipment, namely the disc press and conventional design of screw press, state their equipment is also capable of attaining this solids composition.

In addition to the solids concentration of the solids stream exiting from the dewatering equipment, it is important that the solids in the liquid stream be held to a minimum. This is a necessary requirement for the process design of the disposal plant.

Thus, in the selection of dewatering equipment, the conservation of solids is a prime requirement. Those solids discharged in the liquid stream must be removed in additional processing steps, and it is these streams which dictate the quality of water leaving the process.

FILTRATION

It was found that liquid streams from the mechanical dewatering devices could contain up to 4-5% solids in the form of fines, colloidal materials and solubles. A fourth group of filtration experiments were then conducted using feed streams having concentrations ranging from 0.7% to 4.85% solids. During the course of these experiments, it was noted that the fines could be concentrated to some extent by settling. Work on settling is described in a later section of this report.

In addition to the use of classical techniques (Buchner funnel & filter leaf) the Procedyne bench scale regenerating filter was also used. This filter and its auxiliaries are described schematically in Figure 8. Typical results from that program are shown in Table IV. From this last series of experiments it was concluded that filtration rates up to 0.5 gpm/sq.ft. could be obtained. In all of the batch or semi-continuous experiments during Part I and Part II rapid blinding of the filtration area took place with the formation of a thin manure cake. This of course resulted in a rapid increase in pressure drop and decrease in filter rate.

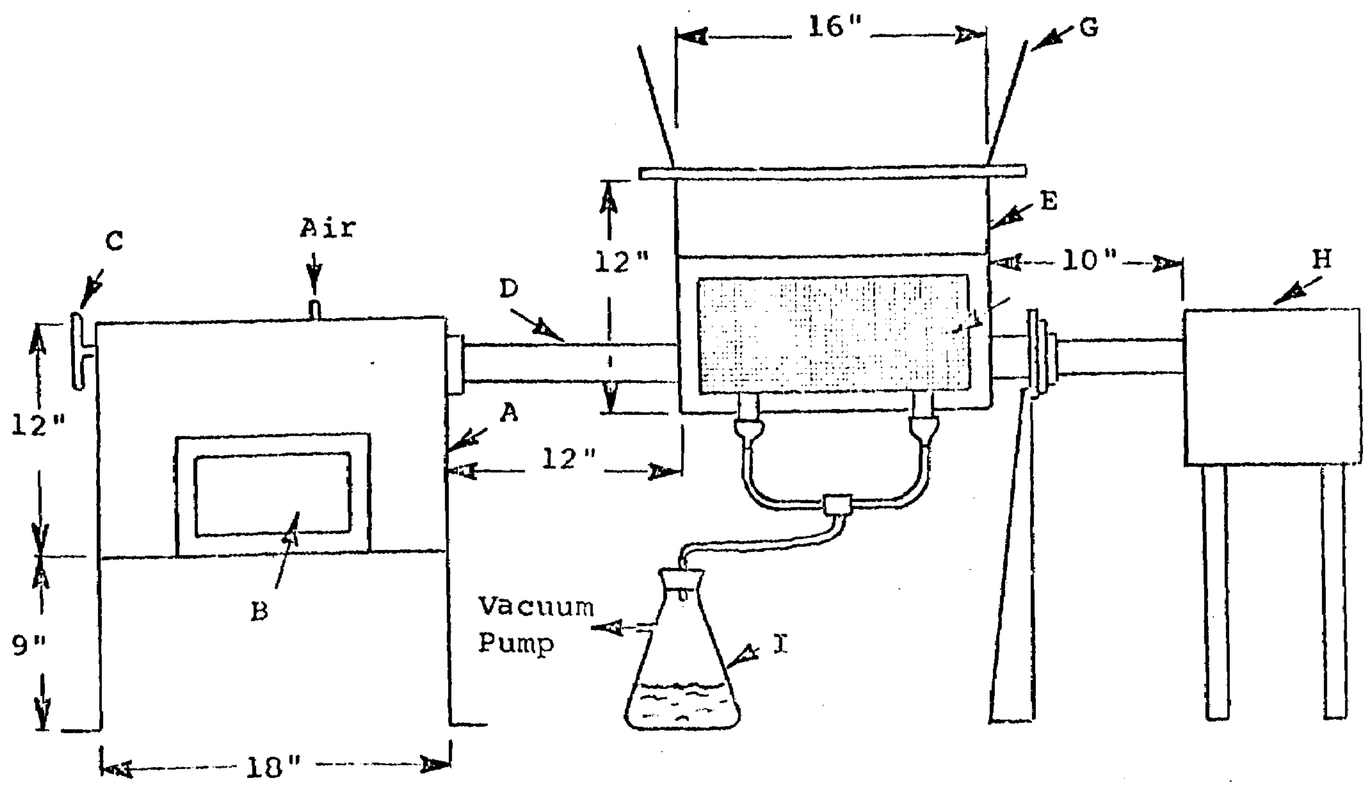
Only a continuous removal of sand and the manure layers prevents blinding. This removal is accomplished while dewatering takes place in a moving bed filter. Filtrate qualities varied from 0.35 to 0.48% solids compared with .31% solids . in soluble or colloidal form and represents material which cannot be removed without additional treatment of the water using tertiary treatment devices.

TABLE IV CONDITIONS AND RESULTS OF EXPERIMENTS

EXPT. NO.	EXPERIMENT DESCRIPTION	FILTRATION RATE GPM/SQ.FT.	FILTRATE QUALITY	SAND QUALITY		CAKE QUALITY		
				%H_2O	%SOL.	%SAND	%SOL.	%H_2O
1	Vacuum filtration with sand, 0.7% solution (initial) Buchner Lab filter paper	0.215	0.41%	---	---	---	23	---
3	Vacuum filtration through Buchner funnel without any sand through lab filter paper, feed concentration 0.7%, 11" Hg vacuum.	0.51	0.48%	---	---	---	21.7	---
4	Vacuum filtration through 170 mesh screen, 11" Hg., 0.7% initial solution, Buchner Funnel.	0.595	0.464	---	---	---	17.3	---
5	Hot vacuum filtration through 170 mesh screen, temp. 165°F, 27" Hg., No sand, initial conc. 0.7%.	1.35	0.342 initial	26.3% in the sand resev.	1.67%	---	14.65	---
6	Procedyne truncated filter, P=10 psi sample is collected on the screen, 0.7% sol. initial	0.5	---	---	---	37	15	---

TABLE IV CONDITIONS AND RESULTS OF EXPERIMENTS (contd.)

EXPT. NO.	EXPERIMENT DESCRIPTION	FILTRATION RATE GPM/SQ.FT.	FILTRATE QUALITY	SAND QUALITY %H_2O	%SOL.	CAKE QUALITY %SAND	%SOL.	%H_2O
7	Filtration through truncated filter, continuous feed of 0.7% sol. and sand Sand rate 450 gms./ 5 min.	0.44	0.494% solid	---	---	38.7	9.3	---
	Mixed Sand	---	---	---	---	86.73	1.84	11.4
	Sand from screw	---	---	---	---	87.6	0.31	12.1
8	Truncated Procedyne Filter, vacuum 14" w.g. feed, 6400 gms of 6.8% sol., 3,200 gms sand, 0.5 psi back pressure.	0.5	0.9%	24%	3.7%	---	---	---



A. Sand and manure receiver

B. Sight glass

C. Pressure gauge

D. Pipe containing 1 3/4" diameter screw feeder

E. Filter chamber

F. Two 6" x 14" screens on both sides

G. Feed Hopper

H. Screw drive system

I. Filtrate receiver

FIGURE 8. REGENERATING FILTER

SEDIMENTATION

As in the case with all commonly used dewatering devices in which the solid stream is concentrated; the liquid stream contains water soluble and colloidal solids and a quantity of small particles in the form of fines. The production of fines in the two streams from the two dewatering devices has been described above.

The fines from the primary dewatering device were principally small strawlike particles and tended to settle rapidly. That stream was also fairly dilute. The fines from the secondary device were somewhat more concentrated and tended to settle more poorly.

Two specific sets of settling experiments were carried out with the two types of particles produced. The results of these are described in the graphs presented in figures 9 thru 12.

As shown in the flowsheet developed for the process, the concentrated, more difficult to settle fines are taken through the filter directly and the easily settlable materials from the primary dewatering device are settled first before feeding to the filter. Filtrate from the filter is recycled through the settler before sewering to take advantage of the settler's capacity and residence time.

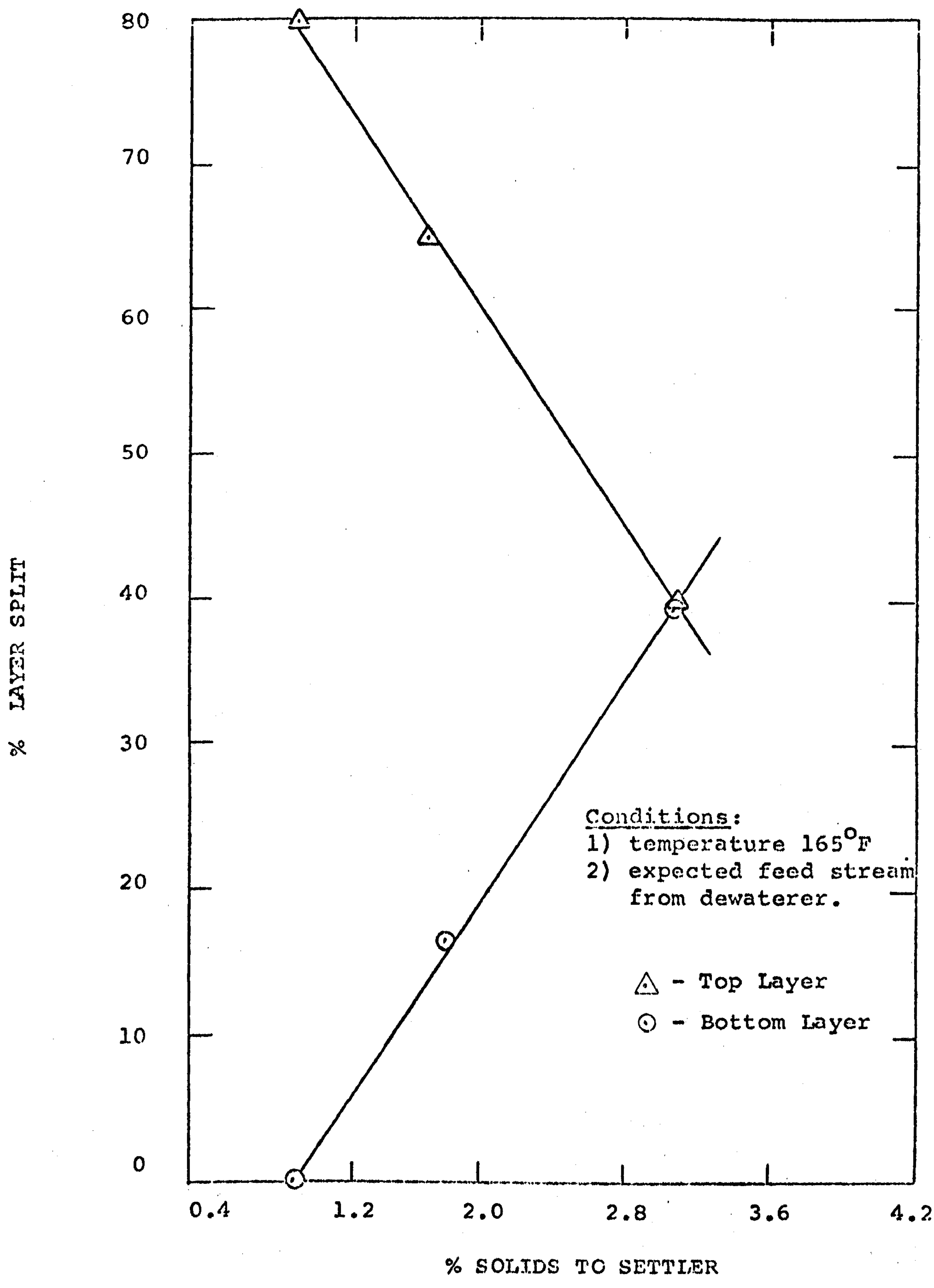


FIG. 9 GRAPH OF % SOLIDS TO SETTLER vs. % OF LAYER SPLIT

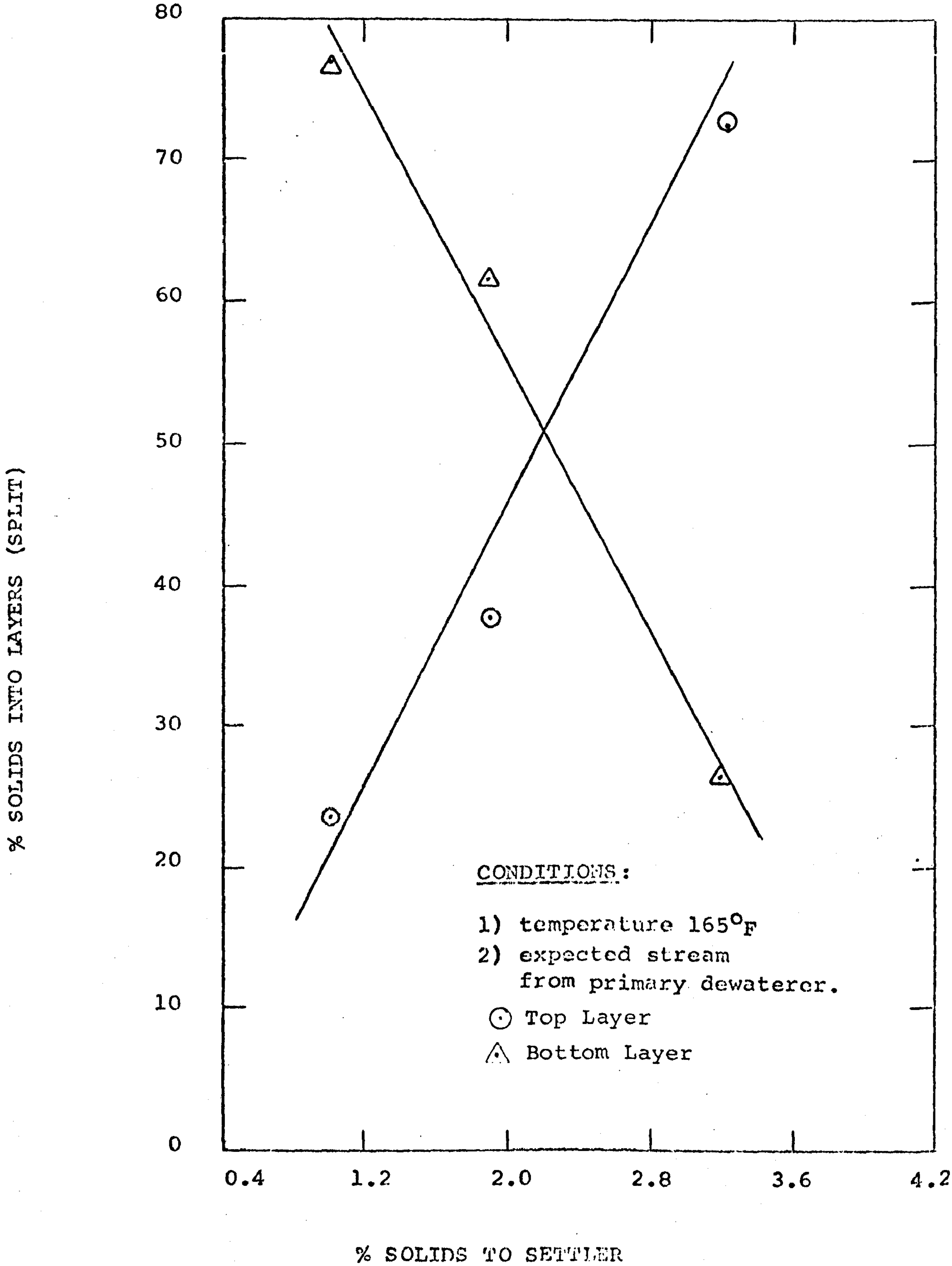


FIG. 10 GRAPH OF % SOLIDS TO SETTLER vs. % SOLID SPLIT INTO LAYERS

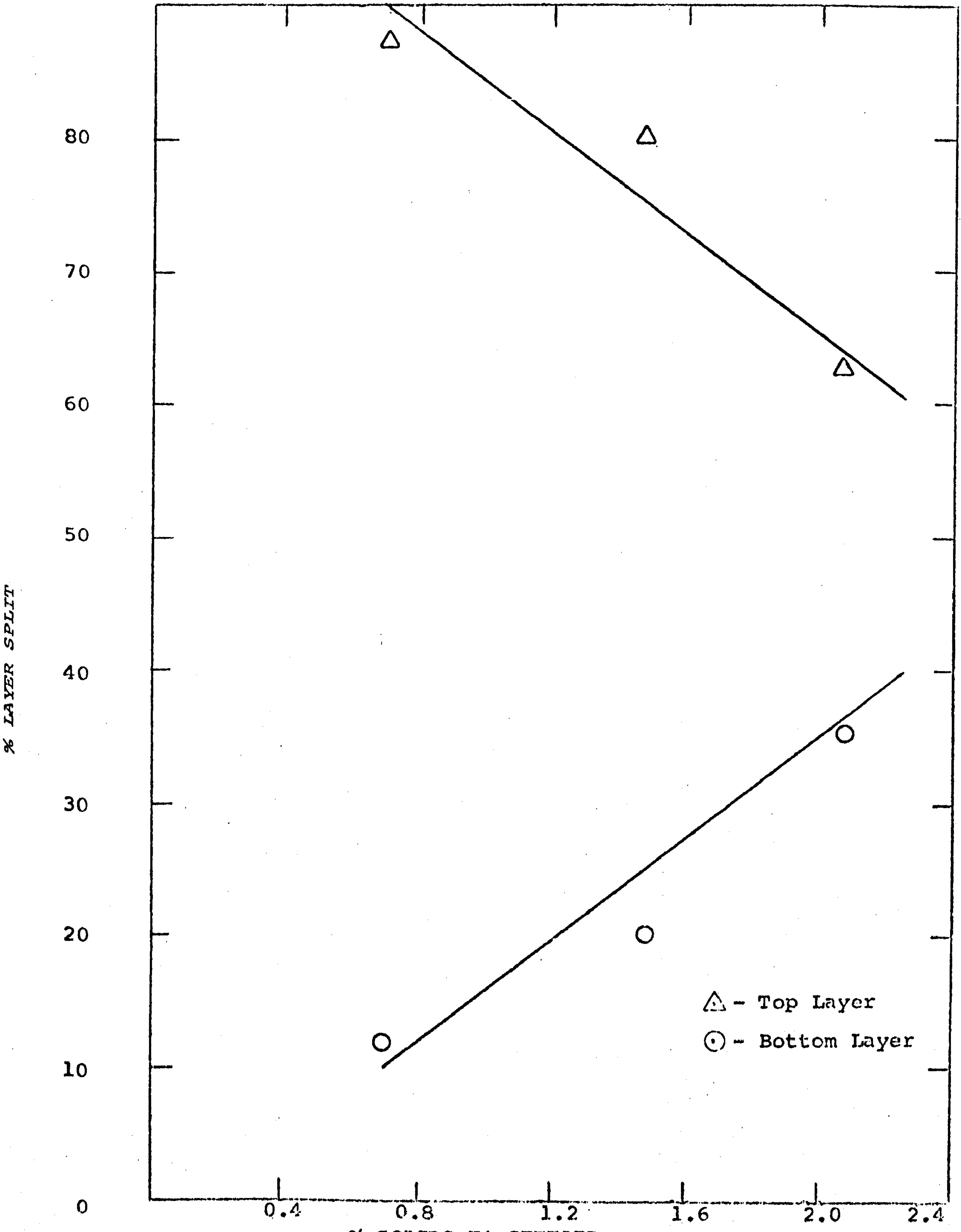


FIG. 11 GRAPH OF % SOLIDS TO SETTLER vs. % LAYER SPLIT

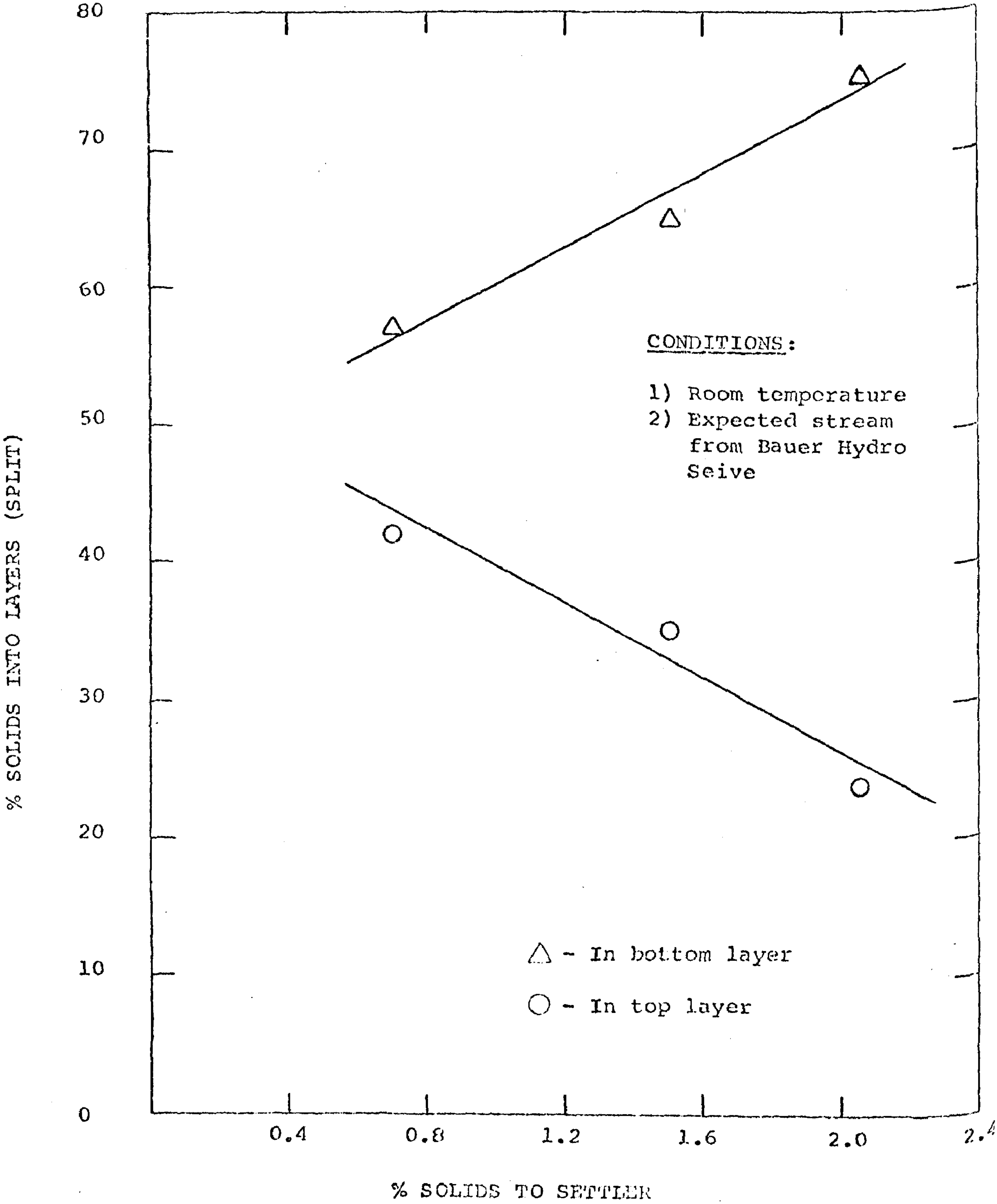


FIG. 12 GRAPH OF % SOLIDS TO SETTLER vs. % SOLIDS IN LAYERS

INCINERATION

A recently developed technique for disposing of sludge from municipal waste treatment plants involves the use of fluidized bed reactors. In this device, sludge containing 25-30% organic solids is fed into a fluidized bed of inert material at approximately 1400°F. The reactor is capable of completely oxidizing (incinerating) all of the organic material. This operating technique is based on the principle long used for processing in the chemical industry, that when solids are suspended in an upward moving stream of gases, the mixture posseses the characteristics of a liquid. The properties of this fluidized bed, in terms of mixing, result in good heat transfer which can be utilized effectively to incinerate manure. Incineration of paunch manure in a fluid bed reactor has not been attempted commercially. Incineration studies were performed on a 6" bench scale fluid bed reactor in order to establish the incineration characteristics of paunch manure.

Description of Laboratory Apparatus

The experimental unit used for these studies consists of a 6" diameter reactor made up of four sections. Each section is made up of type 330 Stainless Steel. The bottom section A (Refer Fig. 13) of the unit is called the plenum chamber where the plenum burner is mounted.

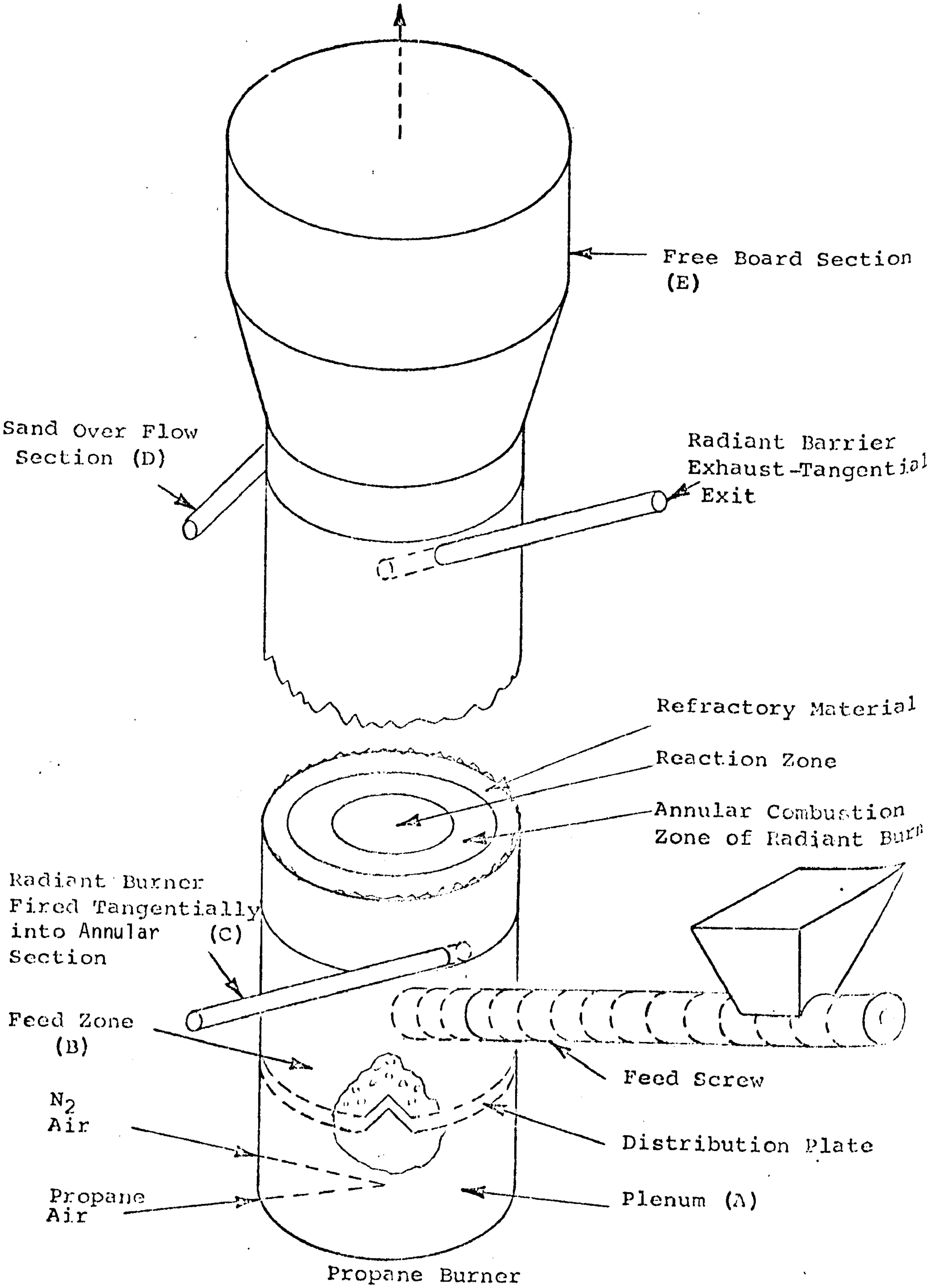


FIGURE 13. Conceptual Drawing - Procedyne Fluid Bed Reactor

There are provisions for a relief valve and for excess air in the plenum. The plenum chamber is lined with 3" of castable refractory. A metal distributor is placed above the plenum chamber. This plate is similar to a bubble cap distributor plate.

The section B above this distributor plate is called the feed section into which the feed is conveyed by screw. A radiant section C of 8" height is placed over this 4" feed section. The radiant section consists of an 8" reactor section, surrounding which is an annular 2" space around this 6" by 8" cylinder. Gas is fired through this annular space tangentially by means of a North American Burner. Castable refractory of 1" thickness is used around the 2" annular section to protect the metal jacket surrounding the radiant section. This metal jacket holds the inside 1" thick refractory. The next higher reactor section is a 4" refractory lined section which is accessible for refractory evaluation purposes.

The sand overflow section sits on top of this, and is also a 4" section. The free board and expanded section (E) completes the fluidized bed reactor. Detailed schematic diagram are shown in Fig. 14.

General Test Procedure

The bed is initially heated to 1400°F by means of both the plenum burner and the radiant burner. When the steady state temperature is attained, propane gas to the plenum

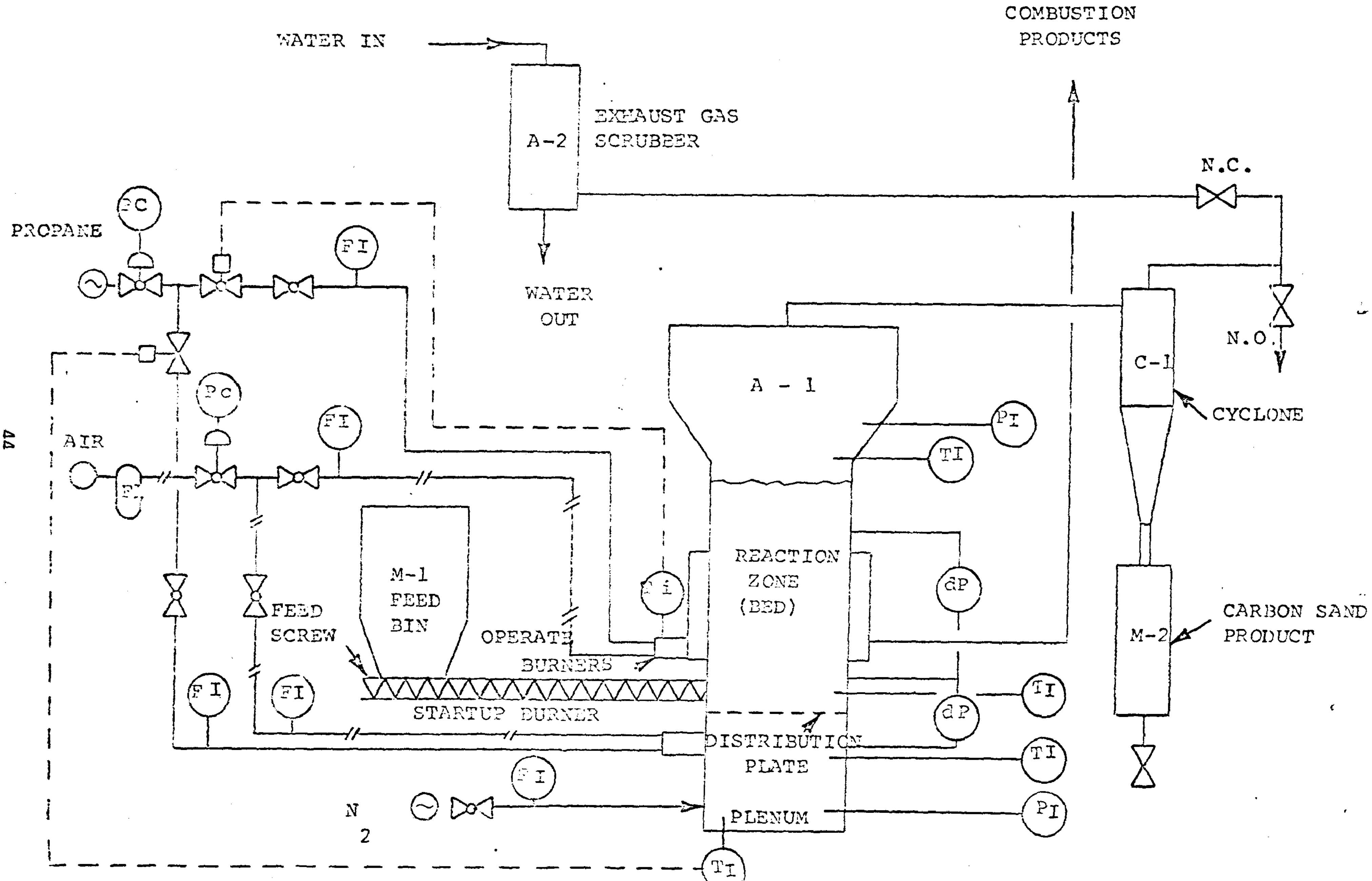


Figure 14 - PROCESS SCHEMATIC FOR PILOT PLANT INCINERATOR

and radiant burner and the air requirements are adjusted according to the heat requirement of the material being burned. Combustion air for paunch manure combustion was supplied at 4% excess oxygen in the exit gases. When the flow rates of air and gas are adjusted, sand in the feed screw hopper is blown initially with air and the screw is started. Sand, which is initially present in the hopper, is fed to the incinerator slowly and then the actual feed is started. The speed of the screw is adjusted to give precalculated feed rates. Gas samples, temperature and pressure measurements at various points are taken when the feed rate and off-gas rate were uniform. Several types of experiments were run during the early part of the work on paunch manure. From a feedstock point of view these could be divided into three general catagories:

(1) Paunch manure at 18% dry solids plus sand in ratios of 1 part manure to 4 parts sand.

(2) Paunch manure in the 28-31% dry solids range as produced from squeezing in Procedyne's bench scale rollers.

(3) Paunch manure-sand mixtures from early bench scale filtration studies.

Results from this experimental program were used in the incineration scaleup calculations described elsewhere in this report.

Data from one each of the three experimental catagories are shown in the Table V.

TABLE V

PAUNCH MANURE INCINERATION EXPERIMENTS

Expt. Number	1	2	3
Feed Stock	Paunch Manure 31% Solids	Paunch Manure 18% Solids + Sand	Filter Mix #5 3.36% solids 12.0% H_2O 84.64% Sand
Feed Rate	16.5 lbs/hr	11.0 lbs/hr	11.5 lbs/hr
Plenum Temp.	1500°F	1430°F	1500°F
Bed Temp.	1300-1410°F	1465-1510	1590-1260
Fluid Bed Height	12"	12"	12"
Main Air Flow	3.5 Scfm	3.0	3.0
Heat input	23,200 Btu/hr	27,200 Btu/hr	18,200
Calc. Heat Loss	15,000 "	-	-
% CO in exit	.02%	tr.	tr.

On the basis of the low levels of CO present in the off-gas, it was concluded that the residence time in the reactor was too short. It was decided to increase the height of the bed and another 4" high x 6" diameter section was added to the laboratory fluid bed reactor.

The experimental program was then continued. Feedstocks for the program were chosen to simulate feeds from the dewatering system being developed in the other preconstruction studies connected with the project. As before, one set of results from each of the three types of feedstock experiments are described in Table VI.

TABLE VI

Expt. Number	1	2	3
Feed	Paunch Manure 37% Solids	% Sol. 4.46 % H_2O 22.66 % Sand 72.70	% Sol. 13.15 % H_2O 31.40 % Sand 55.45
Rate	6.75 lbs/hr	12.6 lbs/hr	12.4 lbs/hr
Air for Combustion	3.9 Scfm	3.75 Scfm	2.89 Scfm
Plenum Temperature	1540-1480	1500	1400°F
Bed Temperature	1332	1300	1400°F
Heat Supplied	17,300 But/hr	15,900 But/hr	17,390 Btu/hr
Radiation Losses	18,200 "	12,000 "	14,000 "
CO in exit gas	.04%	0.0%	0.05%

The general conclusions reached were that feeding of the various feeds was very satisfactory, no major mechanical problems were encountered and that paunch manure could be burned satisfactorily, although 16" bed heights at this scale would be considered an absolute minimum.

Incinerator Design Considerations

The steps used in the basic design of the incinerator are as follows:

(a) Calculate minimum fluidization velocity for the sand to be used.

(b) Determine an operating velocity (based in part on experimental results).

(c) Perform a mass balance on the system. Inputs are; paunch manure + moisture, sand, fuel and air. Outputs are products of combustion, excess air and ash.

(d) Perform a heat balance on the system. Heat inputs are heat of combustion from paunch manure and heat of combustion from fuel. Outputs are heat loss by radiation and in exhaust gases.

(e) Gas flows in the fluidized system are the sum of fluidizing and oxidizing gas required plus products of combustion.

(f) The diameter of the reactor is determined from (b) and (e)

(g) The height of the fluidized bed is determined empirically from experimental results.

Fluid bed incinerators contain a distributer plate through which fluidizing gas passes and which holds up the fluid bed material.

In designing the plate and its holes, calculations are based on the following considerations:

(a) Hole spacing must permit uniform distribution of the fluidizing gas.

(b) Bubble coalescence should not take place right at the plate, otherwise channeling and spouting takes place.

(c) Slugging of the bed should be avoided. This places a restriction on maximum bubble size.

(d) Pressure drop across the bed cannot exceed 10% of the total pressure drop.

(e) Limits are placed on the ratio of orifice diameter to particle diameter.

One then proceeds with the following generalized approach:

(a) List fluidization velocity, reactor diameter, height and inert material size.

(b) Assume a number of holes n and calculate bubble flow per hole, hole spacing L_1, bubble volume and bubble diameter.

(c) Calculate maximum bubble diameter d_{max} and new bubble diameter d_1 resulting from the coalescence between two adjacent bubbles.

(d) $d_1 < \frac{L_1}{2}$ estimated bubble volume is acceptable.

If not, a new diameter d_2 is calculated and a distance L_2 between adjacent bubble tracks is taken as equal to $2 \cdot L_1$. That diameter d_2 is then compared with L_2 and the procedure repeated until d is less than L/2.

(e) That bubble diameter d is the estimated maximum diameter which must agree with d_{max} in section (c).

(f) For the particular number of holes n and the spacing, calculate pressure drop across the place considering each opening as an orifice. This is compared with an empirically determined acceptable number.

(g) Each hole in the plate in an orifice and the diameter of each orifice d_{or} can be determined, number of holes n and mass flow.

(h) The ratio of the orifice diameter to particle diameter is compared and must fall into an acceptable range with pressure drop and weepage of particles through the plate controlling parameters.

(i) This generally implies that several calculations are made for various n's (number of holes) until a satisfactory value of d_{or} is found.

(j) Thickness and detailed construction of the plate will depend on weight of bed and temperature considerations.

The free board in fluidized bed systems permits disengaging of the solids from the fluidizing gas stream.

The gas, in the form of bubbles, erupts on the surface of the fluidized bed. These erupting gas bubbles can and do intermittently splash solids into the free board region above the surface of the bed. This intermittent bursting action of bubbles causes velocity fluctuations and these fluctuations smooth to an average velocity at a certain height. If the gas exit is situated immediately above the top of the bed, a considerable amount of solids will be entrained by gas. With higher gas exit, the amount of entrainment is smaller, and finally a level is reached above which entrainment becomes approximately constant and this height is called transport disengaging height (T D H).

It has been found that the entrainment from a bed closely sized solids is not significant until a superficial velocity U_o considerably in excess of terminal velocity, U_t is reached. Under these conditions, free board acts like a penumatic conveying tube. Thus, according to Zenz and Weil, the solids that are conveyed become constant and is termed as saturation carrying capacity of the gas stream under pneumatic transport conditions initially present.

Several investigations (4,5,6) have shown that elutrication rates increase sharply with rising superficial gas velocity decrease with increasing diameter of fines, and decrease with increasing free board up to a limiting value of the free board beyond which no further entrainment occurs. However, T D H is not as sensitive to gas velocity; it increases by about 70% for a doubling in gas velocity. Hence, it is advisable to design the free board of a fluidized bed very near to T D H.

In spite of the importance of T D H in fluid bed design, there is very little information available on T D H except for the work of Zenz and Weil on FCC Catalyst.

Virtually all the reported work on entrainment from fluidized beds has been carried out with either of two simplifications: Singly, closely sized fractions or a mixture of two such fractions in a small laboratory units. It was felt that these results may not establish a correct criterion for elutriation since that column diameter, particle diameter, particle density, density of gas, viscosity of gas are effecting entrainment eitherway, depending on the specific conditions.

Hence, for engineering design of free board for Procedyne's fluidized bed reactor, T D H is calculated from the basic concept of bubble theory and checked by using the correlation developed by Zenz and Weil and as reported in a recent publication. (7)

As already pointed out the bursting action of the bubbles

on the surface of a fluidized bed projects agglomerates of particles into the space above the bed. These particles (depending on size) will be carried by the gas into the free board. Practically all the gas that is in excess of minimum fluidizing conditions passes through the bed in the form of bubbles. These bubbles grow in size as they pass through the bed with a velocity that is higher than superficial velocity of gas. While the bubbles travel through the bed, they collide with other bubbles and trap some solids in the bubble and wake. These solids that are in the wake are also carried along with the bubble. When these bubbles reach the surface of the fluidized bed they burst, thus throwing the solids in the bubble in the free board. At this stage, some of the solids will be having a velocity equal to bubble velocity. When this phenomenon occurs, some particles will be carried away in the free board by gas depending on the terminal velocity of gases. Some of them will fall back into the bed.
The following generalized approach was therefore used in calculating T D H.

(a) Small bubbles form at the distributor, coalesce, grow and speed up as they rise through the bed.

(b) The velocity of rise of a crowd of bubbles is related to the velocity of rise of a single bubble in a bed. The absolute rise velocity of bubbles in bubbling bed is calculated by Davidson's (8) model.

(c) Bubbles are assumed to be spherical and their diameter are calculated by Davidson's (8) model.

(d) When a bubble bursts, the particles in the wake and bubble are thrown up. The particles whose $U_t < U_b$, are carried in the free board.

(e) The frequency of bubbles at the surface of bed is calculated from [4] $n = 1.5 \frac{(U_o - U_{mf})}{d_b}$

(f) The voidage of the wake in the bubble is equal to the voidage @ U_{mf}.

(g) Ratio of volume of wake to bubble is taken to be 0.2, for irregular sand from the data of Rowe and Partridge [9].

(h) The basic eqauation for entrainment is same as that used by Lewis et al (10).

(i) The entrainment rate is maximum 1% of the total sand in the bed per day.

IV PROCESS CALCULATIONS

Basis for Design: (1)

$$95 \frac{\text{cattle}}{\text{hour}} \times 55\# \frac{\text{paunch manure}}{\text{cattle}} = 5225\# \frac{\text{paunch manure}}{\text{hour}}$$

Now, paunch manure as removed from the sack is 14% (wt) dry solids, this is equivalent to:

$$5225\# \frac{\text{paunch manure}}{\text{hour}} \times \frac{14}{100} = 731\# \frac{\text{Dry solids}}{\text{hour}}$$

Paunch Table Operation

The paunch manure, after removal from the sack is free drained, draining to 17% Dry solids, the water drained containing dissolved and very fine solids.

$$\frac{731}{.17} = 4310\# \frac{\text{paunch manure}}{\text{hour}} \text{ (after draining)}$$

5225 #/hour - 4310 #/hour = 915#/Hr. Total material drained away.

Solids in drainage is 3000 ppm

915 #/hr. x .003 = 3#/Hr. solids in drainings.

Based on the calculations above, stream OA is composed of the following.

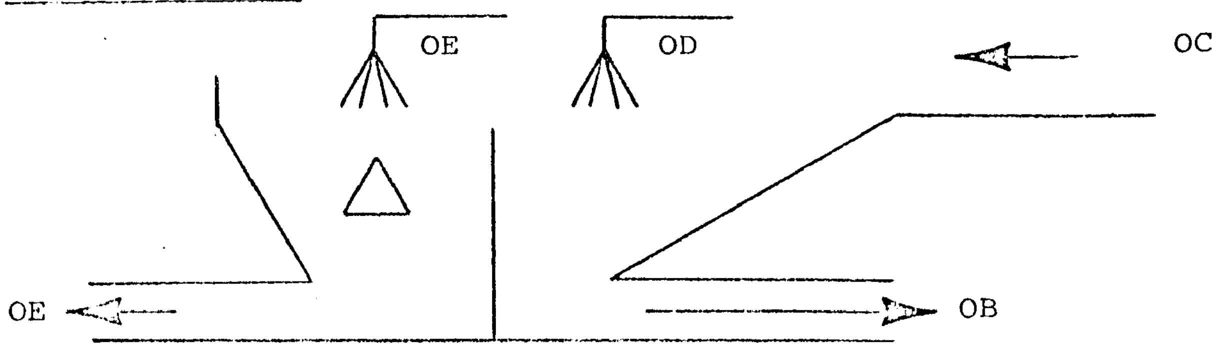
Paunch Manure (As Rec'd)	Total 5225	Dry solids 731	Water 4494	%Solids 14
Paunch Manure (drained)	4310	728	3582	17
Paunch Manure Drainings	915	3	912	0.3

Water is added for the following:

STREAM	FUNCTION
OC	Table Rinse
OD	Sack Wash
OE	Sack Rinse

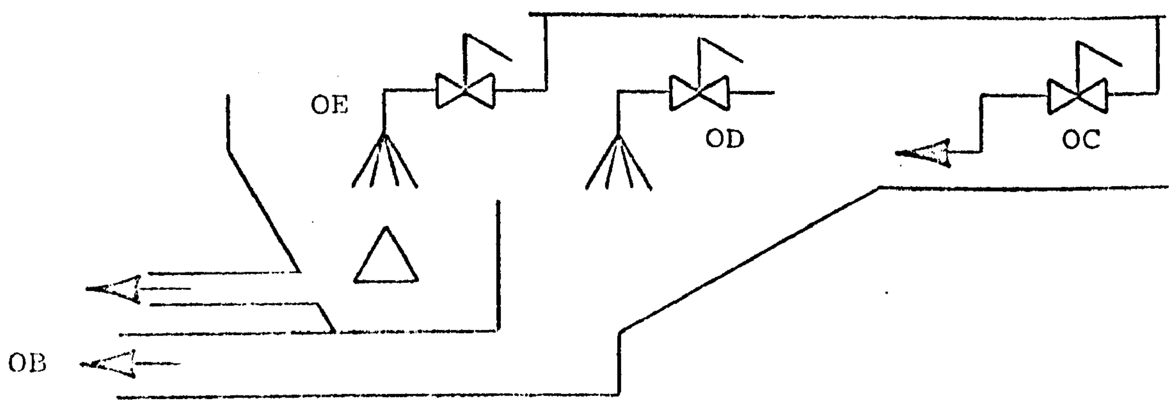
At present, at Illinois Packing Co., these three streams are continuous, each approximately 20 gpm, or a total of 60 gpm. Significant improvements in the disposal plant efficiency will be realized by installing foot pedal valves OC and OD. This will result in significant water savings, estimated at 2/3 of the present water usage. The new water rate is assumed to be 20 gpm. This figure is designated stream OB. Stream OE, the sack rinse, will be sewered directly.

Present System



Present Water Flow - 60 gpm (20 gpm each flow)

Pedal Controlled System



Under this system, OB will flow at rate of 20 gpm.

$$20 \frac{\text{gallons}}{\text{minutes}} \times 60 \frac{\text{minutes}}{\text{hour}} \times \frac{8.33\#}{\text{gallon}} = 10{,}000 \frac{\#}{\text{hr}}$$

It is assumed that this water will extract fine particles of paunch manure in range of 3000 ppm.

$$10{,}000\# \times 0.003 = 30\# \frac{\text{fine solids}}{\text{hour}}$$

Stream 1, exiting from the paunch table is composed of streams OA and OB and is as follows:

	OA	OB	Total Stream 1
Free Water	912	10,000	10,912
Bound Water	3582	--	3,582
Dry solids	731*	--	731*
			15,225

*of this 731# of Dry solids, 33# is estimated to be dissolved or very fine solids.

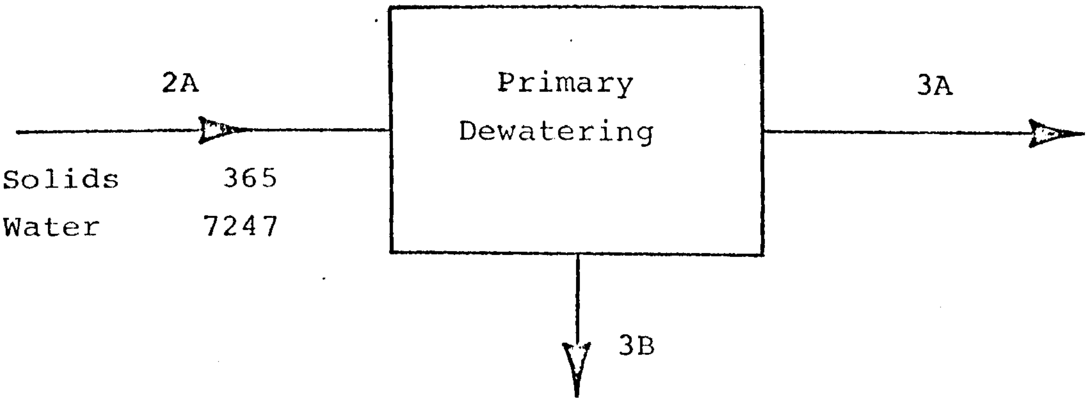
PAUNCH BIN

This stream (1), is fed to the paunch bin which functions as both surge bin and feed tank to the remainder of the process. The plant design is based on a 16-hour per day operation, or an hourly rate of 365.5 of paunch manure on a dry basis.

Feed to process: (Stream 2A)

Free water	5456.0
Bound water	1791.0
Dry solids	365.5
	7612.5

Primary Dewatering



1. The expected yeilds in streams 3A and 3B are as follows:
   a. 3A = .8 of total solids in 2A = 292 #/hr.
   b. 3B = .2 of total solids in 2A = 73 #/hr.
   c. Solids in stream 3A = 18%

2. Water in Stream 3A:

   $\frac{292.4}{.18}$ = 1625 #/hr. - total weight stream 3A

   1625 - 292 = 1333 #/hr. water in stream 3A

3. Water in Stream 3B:

   Total water - water in 3A = water in 3B

   7247 - 1333 = 5914 #/hr. (8)

4. Determination of Bound Water - Stream 3B only:

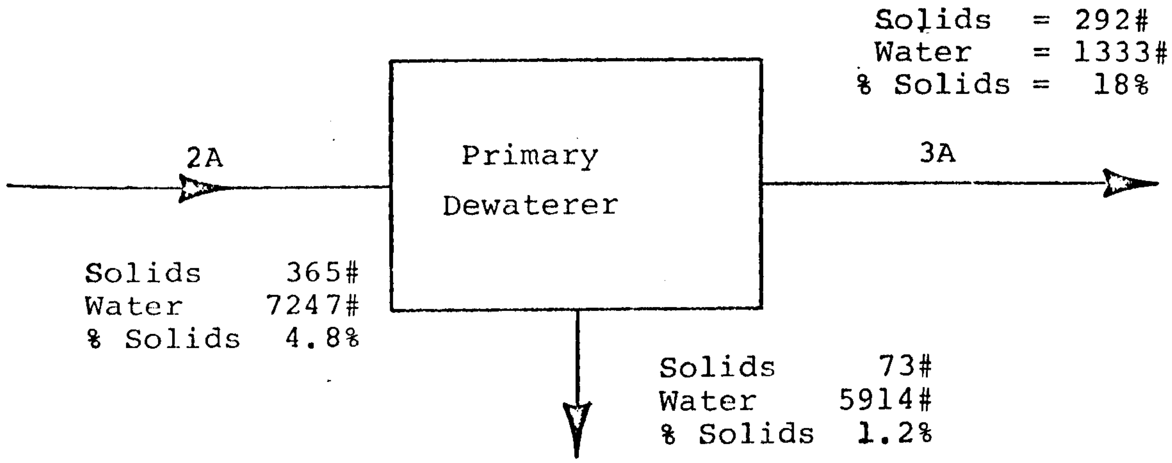
   Total water - free water = bound water

   5914 - 5456.0 = 458 #/hr.

5. Solids Concentration in 3B:

   $\frac{73}{73 + 5914} \times 100 = 1.22\%$

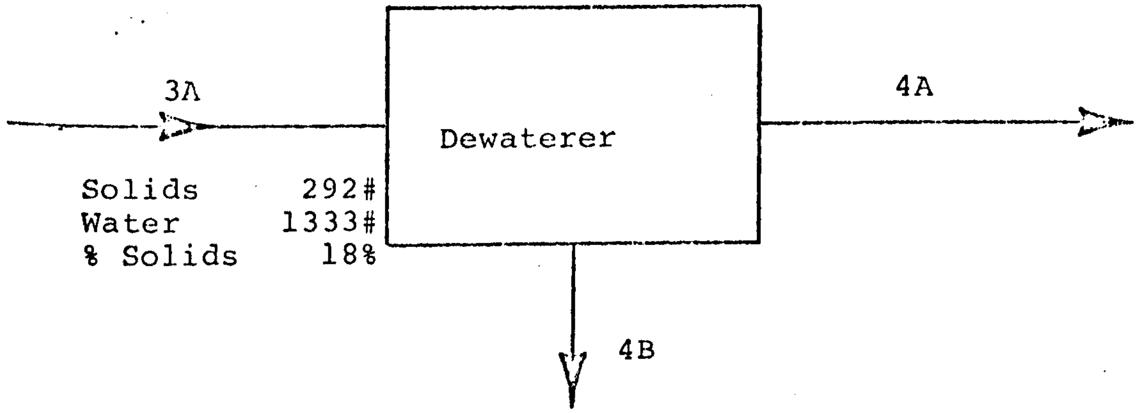
6. Thus, balance about the primary dewaterer:



Stream 3A, the solids rich stream, goes forward to the dewaterer, stream 3B, relatively weak in solids, goes to the settler for thickening.

Dewaterer

Stream 3A is fed directly to the dewaterer, yielding 2 streams; 4A, high in solids content and 4B, a liquid rich stream.



1. The expected compositions of 4A and 4B are as follows:
   a. 4A has a composition of 40% dry solids
   b. 4B has a composition of 5% dry solids
   c. 82.5% of entering solids in stream 3A will be recovered in stream 4A

Let stream 4A = X and 4B = Y

Total weight of stream 3A = 1625 #/hr.

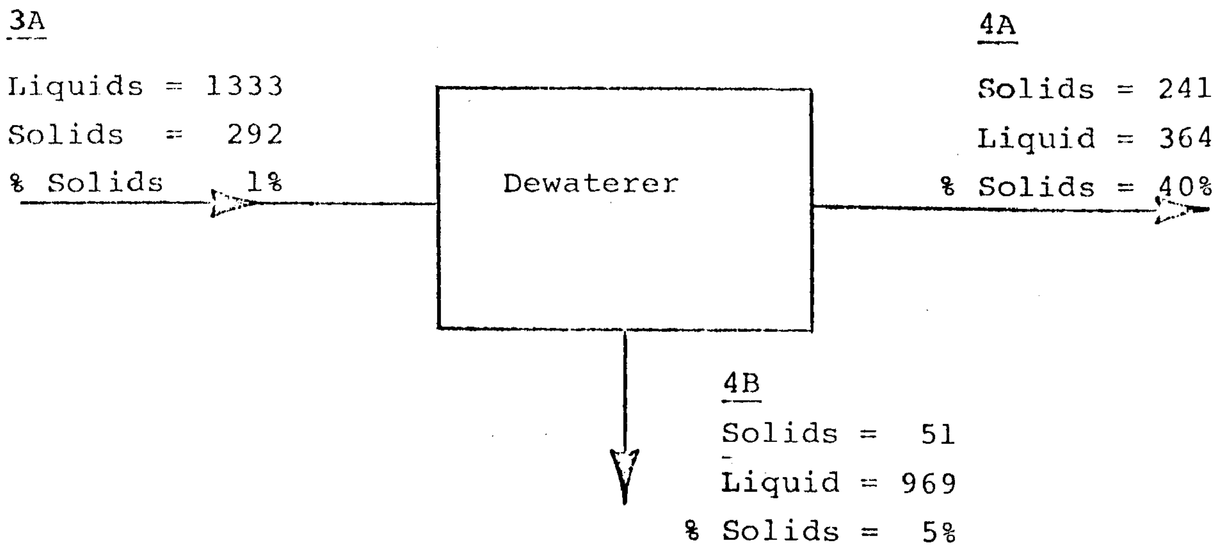
Therefore X + Y = 1625#

and

Simultaneous solution of these equations yields the following:

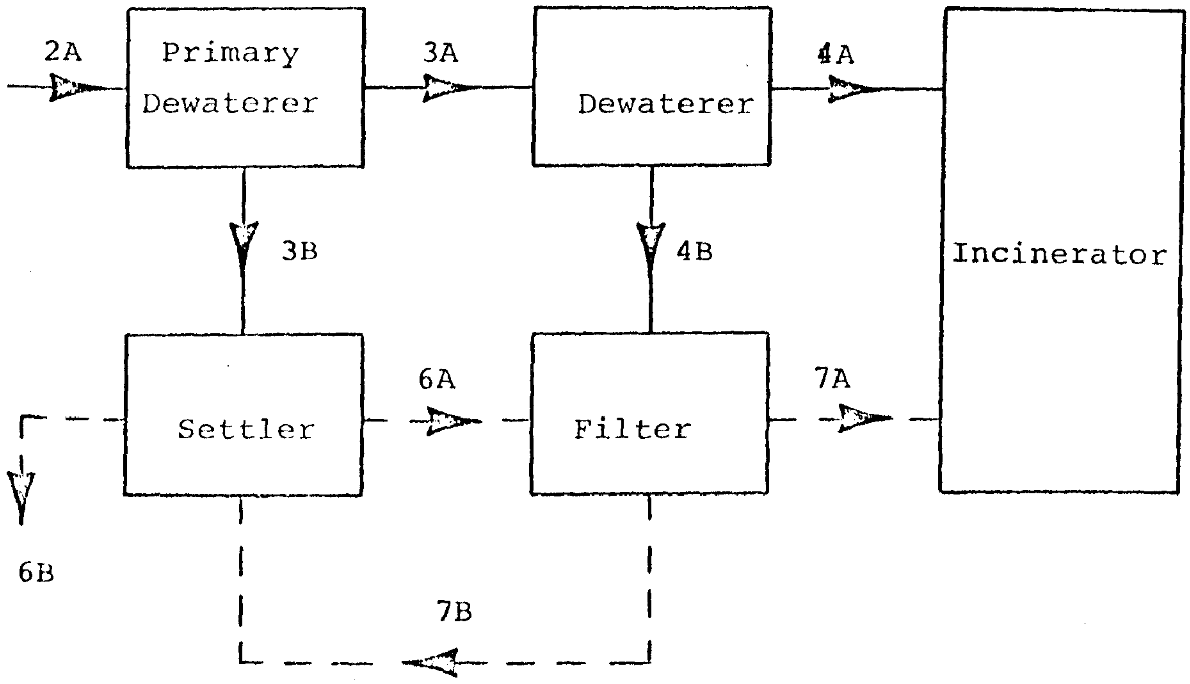
Stream	4A	+	4B	=	3A
Water	364		969		1333
Solids	241		51		292
Total	605 #/hr.		1020 #/hr.		1625 #/hr.

Thus, balance about the dewaterer:



Settler and Filter

Stream 4A is fed directly to the incinerator, Stream 4B is sent directly to the sand filter where combining with Stream 6A (from the settler), it is filtered, the solids Stream 7A going to the incinerator, and the filtrate 7B going back to the settler for further treatment. This is shown schematically on the following page.



It can be shown that those streams indicated by the dashed lines (6A, 6B, 7A, 7B) are dependent on the settler and filter performance. Stream 7B is recycled, effecting both stream 6A and stream 6B which in turn effects stream 7A and 7B, and so forth. By a series of consecutive, iterative calculations, these streams are determined to be:

Stream	6A	6B	7A	7B
Dry Solids	69	37	87	33
Water	1625	6453	430	2164
% Solids	4.1%	0.58%	0.17%	1.57%

Stream 4B, the required sand for filtration is also arrived at by these calculations. This value is 1352# sand/hour.

Incinerator

The feed to the incinerator has been determined to be:

Stream	4A	7A	Total
Dry solids (#/hr)	241.	87	328
Water (#/hr)	364	430	794
Sand (#/hr)	-	1352	1352

This is 90% of the total dry solids entering the process, being fed to the incinerator as 29.4% solids (sand free basis). The following process design criteria have been determined:

a. % ash in dry solids - 8%

b. Heat of combustion of paunch manure - 7200 Btu/hr.

c. Excess Air (dry basis) 4%

d. Paunch manure taken to be cellulose.

Material and Heat Balance Calculations:

1. Feed to Incinerator

   a. Pounds of combustibles present:

   329.4 (.92) = 303 #/hr

   b. Pounds-moles of combustibles present (as cellulose):

   (M.W. Cellulose = 162 #/hr-mole)

   $$\frac{303}{162} = 1.87 \ \frac{\text{\#-moles}}{\text{hour}}$$

   c. Combustion equation:

   $$C_6H_{10}O_5 + 6O_2 - - - - 6CO_2 + 5\ H_2O$$

   c.1. Stochiometric quantity of Oxygen required for complete combustion:

   (1.87) (6) = 11.22 # - moles/hr.

c.2. Amount of Nitrogen, (a)-above provided by air.

$$\frac{11.22 \text{ \#-moles } O_2}{HR.} \times \frac{3.76 \text{ \# moles } N_2}{\text{\# mole } O_2} = \frac{42.2 \text{ \# moles } N_2}{HR.}$$

d. Water present in the feed.

$$\frac{793.6 \text{ \# } H_2O}{Hr.} \times \frac{1 \text{ \# mole } H_2O}{18 \text{ \# } H_2O} = \frac{44.1 \text{ \# moles } H_2O}{Hr.}$$

2. From combustion of paunch manure

a. CO_2 from combustion of paunch, from combustion equation:

1.87 x 6 = 11.22 # moles CO_2

b. H_2O from combustion of paunch, from combustion equation:

$$1.87 \times 5 = \frac{9.35 \text{ \# moles } H_2O}{HR.}$$

c. Nitrogen present (1-c-2) above

$$42.2 \frac{\text{\# moles } N_2}{hr.}$$

d. Water present in feed (1-d) above

$$44.1 \frac{\text{\# moles } H_2O}{hr.}$$

3. Heat from Combustion of manure

303 #/hr x 7200 Btu/# = 2,181,600 Btu/hr.

4. Heat Output (All Discharged at 1400°F)

a. Heat required to heat water in feed: to 1400°F

793.6 x 1700 = 1,345,000 Btu/hr.

b. Heat required by CO_2

$NCP\Delta T = Q$

$11.22 \times 1328 \times 11.48 \text{ Btu} = 170,800 \frac{\text{Btu}}{\text{hr.}}$

c. Heat required by H_2O formed.

(9.35) (18) (1700) = 286,000 Btu/hr

d. Heat required by N_2 present

(42.2) (7.35) (1328) = 412,000 Btu/hr

e. Heat required by excess O_2 present:

Let X represent # moles of excess O_2

X (7175) (1328) = 10,280 X Btu/hr

f. Heat required by excess N_2 present:

excess N_2 = 3.76 x

3.76 (x) (7.35) (1328) = 36,700x Btu/hr

5. Heat losses from System

a. Incinerator - 9.5' O.D. x 23.75' high

b. Area of shell - II D H

(3.14) (9.5) (23.75) = 710 ft^2

c. Area of Heads = $\frac{II\ D^2}{4}$

$\frac{3.14}{4}$ (90) (2) = 142 ft^2

d. Total exposed area of incinerator = 852 ft^2

e. Total heat losses;

852 ft^2 x $\frac{440\ Btu}{ft^2}$ = 375,000 Btu/hr

6. Heat Loss From Recirculating Sand:

1362 x .25 x 1328 = 450,000 $\frac{Btu}{hr}$

7. Summation of Heat Losses

Water in Feed	1,345,000
CO_2 formed	170,000

H_2O formed	286,000	
N_2 present	412,000	
Excess O_2		10,280x
Excess N_2		36,700X
Surface Heat losses	375,000	
Recirculating Sand	450,000	
Total Heat Losses (Btu/hr)	3,038,800 +	46,980x

8. Total heat requirement

Heat losses - Heat supplied by combustion of paunch = heat required.

3,038,800 + 46,980x - 2,184,600 = 857,200 + 46,980x(Btu/hr)

x ← 18

Calculation of Fuel and Air Requirements:

The calculations are based on a fuel oil composition of 87% Carbon, 12% Hydrogen and 1% Sulfur, and exit gases from the incinerator at 1500°F. The basic equations used are common to the combustion literature (11)and calculations are developed as follows.

1. Heat available; $\frac{.87 \text{ \#Carbon}}{1\text{\# combustible}} = \frac{66 \text{ BTU}}{\text{Scf air}}$

2. Air required (Scf) $= \frac{\text{Btu/hr required}}{66}$

3. Air required (lb.-moles) $= \frac{\text{BTU/hr. required}}{23.694}$

4. O_2 in this air (lb-moles) $= \frac{\text{BTU/hr. required}}{113,000}$

5. N_2 in this air (lb-moles) $= \frac{\text{BTU/hr required}}{30,000}$

Air required for combustion:

6. $\frac{\text{Air required (Scf)}}{\text{\#Fuel}} = \%\,C(1.514)+\%H(4.54)+\%S(.568)$

$= 188 \frac{\text{Scf}}{\text{\#Fuel}}$ or $.525 \frac{\text{lb.-moles}}{\text{\#Fuel}}$

7. $\frac{CO_2 \text{ formed (Scf)}}{\text{\#Fuel}} = .315(\%C) = \frac{27.4 \text{ Scf}}{\text{\# Fuel}}$

or $.076 \frac{\text{lb-moles } CO_2}{\text{\# Fuel}}$

8. $\frac{H_2O \text{ formed (Scf)}}{\text{\#Fuel}} = 1.89\ (\%H) = \frac{22.68 \text{ Scf}}{\text{\# Fuel}}$

or $.063 \frac{\text{lb.-moles } H_2O}{\text{\# Fuel}}$

From equation (3) and (6) above:

$$9. \quad \frac{.525 \text{ lb.-moles Air}}{\#\text{Fuel}} \times N\# \text{ Fuel} = \frac{\text{BTU/hr required}}{23,694}$$

$$\text{or: } N\text{=Fuel required} = \frac{\text{BTU required}}{12.500}$$

Rewriting equations (7) and (8) above:

$$10. \quad \text{lb.-moles } CO_2 \text{ formed} = .076 \left(\frac{\text{BTU required}}{12,500}\right)$$

$$= 5.2 + .284x$$

$$11. \quad \text{lb.-moles } H_2O \text{ formed} = .063 \left(\frac{\text{BTU Req'd}}{12,500}\right)$$

$$= 4.32 + .236x$$

Summation of Products

	CO_2	H_2O	N_2	O_2
A	11.22	9.35	42.2	--
B	--	44.1	--	--
C	(5.2 + .284x)	(4.32 + .236X	(28.5 + 1.56X	--
D	--	--	3.76x	x

Where:

A = Products of combustion of paunch manure.

B = Water in paunch manure.

C =Products of combustion of additional fuel.

D = Excess Air

Thus:

$CO_2 = 16.42 + .284X$

$H_2O = 57.77 + .236X$

$N_2 = 70.70 + 5.32\ X$

$O_2 = X$

Calculation of excess oxygen on a dry basis:

Total moles of dry gas = 87.12 + 6.60x

$$\frac{4}{100} = \frac{x}{87.12+6.604x} \qquad X = 4.76\ \frac{\text{lb.-mole}}{\text{Hr}}$$

Composition of Flue Gases (lb.-moles/hr.)

CO_2	17.77
H_2O	58.89
N_2	95.90
O_2	4.75
	177.31 lb.-moles/hr.

Additional heat required:

$$\frac{\text{BTU}}{\text{hr}}\ \text{(required)} = 857{,}200 + 46.980\ x$$

$$= 1{,}080{,}000\ \frac{\text{BTU}}{\text{Hr.}}$$

Fuel oil required:

$$N = \frac{\text{BTU/hr required}}{12{,}500} = 86.5\ \text{\#/hr.}$$

Air requirement (lb.-moles/hr.):

For manure combustion	53.4
For fuel combustion	45.5
For excess air requirement	22.7
	121.6 lb.-moles/hr.

or:

$$121.6\ \frac{\text{lb.-moles}}{\text{Hr.}} \times \frac{359}{60} = 727\ \text{SCFM}$$

Effect of sulfur content of fuel; sulfur content is assumed at 1%.

$$\frac{86.5\ (.01)}{32} = .027 \text{ lb.-moles sulfur}$$

$$S + O_2 \longrightarrow SO_2$$

O_2 requirement for sulfur = .027 lb.-moles.

SO_2 in stack = 0.027 lb.-moles/hr.

Composition of stack gases:

CO_2	17.77
H_2O	58.89
N_2	95.90
O_2	4.75
SO_2	0.027
	177.337 lb.-moles/hr. (1064 SCFM)

Solids content of incinerator exit gases. This stream is based on the following 1) solids discharge is composed of ash and sand, 2) all ash is carried over, and 3) sand carry over is equal to 1% of bed capacity per 24 hrs. of operation.

Ash content	26	#/hr.
Sand content	10	#/hr

Loading to cyclone $= \frac{233 \text{ grains}}{\text{Scf.}}$

Cyclone

Gases to cyclone (lb.-moles/hr.)

CO_2	17.77
O_2	4.75
N_2	95.90
H_2O	58.89
SO_2	0.03
	177.34 lb.-moles/hr.

The solids to the cyclone as shown above are 26 lb./hr of ash and 10 lb./hr. of sand.Based on a solids collection efficiency of 90% of all particles above the 5 micron size and a particle size analysis as follows:

Sand - All above 5 micron size

Ash -	32%	0-5 micron
	36%	$>$ 5-10 micron
	23%	$>$10-20 micron
	8%	$>$20-40 micron
	1%	$>$40-80 micron

On this basis it is assumed that 100% of the sand and 60% of the ash entering the cyclone will be collected.

Collected in cyclone:

Sand	10#
Ash (26)(.6)	15#

Solids loading to scrubber:

Ash - 11# or 72 grains/ scf.

<u>Scrubber</u>

The stream fed to the scrubber is a s follows:

Gases (lb.-moles/hr)

CO_2	17.77
O_2	4.75
N_2	95.90
H_2O	58.89
SO_2	0.03
	177.34 lb.-moles/hr.

Solids

Ash 11#/hr (72 grains/scf.) This stream entering the scrubber will be at 1400°F. The gases exiting from the scrubber will be saturated and leaving at 185°F. The water evaporated in the scrubber is 1585 #/hr.

Composition of gases leaving scrubber (lb.-moles/hr.):

CO_2	17.77
O_2	4.75
N_2	95.90
H_2O	146.90
SO_2	0.03
	265.35 lb.-moles/hr.

Solids from Scrubber

Scrubber is to be sized on 1/2 the maximum allowable particulate emission (City of Chicago Environmental Control Ordinance, Chapter 17)of 0.2 grains/scf. at 50% excess air. The allowable emission under this code is 4.1 #/hr, the design criteria is 2.0 #/hr. or 0.1 grains/scf.

Solids

Ash 11#/hr (72 grains/scf.) This stream entering the scrubber will be at 1400°F. The gases exiting from the scrubber will be saturated and leaving at 185°F. The water evaporated in the scrubber is 1585 #/hr.

Composition of gases leaving scrubber (lb.-moles/hr.):

CO_2	17.77
O_2	4.75
N_2	95.90
H_2O	146.90
SO_2	0.03
	265.35 lb.-moles/hr.

Solids from Scrubber

Scrubber is to be sized on 1/2 the maximum allowable particulate emission (City of Chicago Environmental Control Ordinance, Chapter 17)of 0.2 grains/scf. at 50% excess air. The allowable emission under this code is 4.1 #/hr, the design criteria is 2.0 #/hr. or 0.1 grains/scf.

ECONOMICS OF AN AIR PREHEATER INSTALLATION

The temperature which the air is heated to will be 1200°F. This provides a maximum amount of heat transferred to the air, while not pinching in the warm end temperature difference in the heat exchanger to too low a value (200°F).

Let X = lb. moles/hr of excess O_2

then $53.4 + \frac{100X}{21}$ is the lb moles/hr of air, following the calculation procedure already established. This does not include air for combustion of supplementary fuel. The total heat losses with no air preheating, as previously computed are:

$$857{,}200 + 46{,}980\ X\ \text{Btu/hr}$$

Heat added to the system by preheating the air with flue gases is:

$$\left(53.4 + \frac{100X}{21}\right)(7.3)(1200-72) = 439{,}000 + 39{,}200X\ \text{Btu/hr}$$

Therefore the heat losses are reduced to:

$$418{,}200 + 7{,}780X$$

For supplementary fuel with flue gases leaving at 1400°F. 66 Btu of heat are available for every scf of air used in supplying supplementary heat.

Using the same equation numbers as in the master calculation:

3. Air required $\frac{\text{(lb moles)}}{\text{hr}} = \frac{\text{Btu/hr required}}{23{,}694}$
4. O_2 in this air $= \frac{\text{Btu/hr required}}{113{,}000}$
5. N_2 in this air $= \frac{\text{Btu/hr required}}{30{,}000}$
6. Air required for combustion of 1 lb. of fuel oil is 0.525 lb. moles as computed previously
7. CO_2 formed = 0.076 lb moles/lb fuel
8. H_2O formed = 0.063 lb moles/lb. fuel

   From (3) and (6)

   N = no. of lbs. of fuel oil/hr = $\frac{\text{Btu/hr required}}{12{,}500}$

   Rewriting equation (7) and (8) above =

10. $\frac{\text{lb moles}}{\text{hr}}$ CO_2 formed = $0.076 \frac{\left(\frac{\text{Btu}}{\text{hr}}\text{ required}\right)}{12{,}500}$

$= 2.543 + 0.0473X$

11. lb moles H_2O formed = $0.063 \frac{\left(\frac{\text{Btu}}{\text{hr}}\text{ required}\right)}{12{,}500}$

$= 2.11 + 0.0392X$

SUMMATION OF PRODUCTS

	CO_2	H_2O	N_2	O_2
A.	11.22	9.35	42.2	----
B.	----	44.1	----	----
C.	(2.543+0.0473X)	(2.11+0.0392X)	(13.94+0.259X)	---
D.	----	----	3.76X	X

Where:

A. = Products of combustion of paunch manure

B. = Water in paunch manure

C. = Products of combustion of additional fuel

D. = Excess Air

$CO_2 = 13.76 + 0.0473X$

$H_2O = 55.6 + 0.0392X$

$N_2 = 56.1 + 4.02X$

$O_2 = X$

Calculation of excess O_2 on a dry basis = 4%

$$\frac{4}{100} = \frac{X}{69.86 + 5.067X}$$

$X = 3.50$ lb moles O_2/hr

Composition of flue gases

CO_2	13.96 lb moles/hr
H_2O	55.74
N_2	70.2
O_2	3.50
	143.40

Additional [illegible] required = 4[illegible],200 + 7,780X = 445,400 $\frac{\text{Btu}}{\text{hr}}$

Fuel oil [illegible] = $\frac{[illegible]00}{[illegible]}$ = 35.[illegible]/hr.

Air requirement (lb mole/hr)

For manure combustion	53.4
For fuel combustion	18.7
For excess air	16.7
	88.8 or 531 SCFM

	Fuel Oil Required lb/hr	Air Required SCFM
No preheater	86.5	727
Preheater	35.7	531

Annual savings in fuel (4000 hrs/yr operation)

$$50.8 \frac{lb}{hr} \times \frac{gal}{7.44\ lb} \times \frac{\$0.10}{gal} \times 4000 \frac{hr}{yr} = \$2730.$$

Annual savings in blower electricity (based on centrifugal blower performance)

$$\frac{14.4\ BHP}{600\ SCFM} \times (727-531)\ SCFM = 4.7\ BHP$$

$$4.7\ BHP \times \frac{kw}{1.34\ HP} \times \frac{1}{0.85} \times 4000 \frac{hr}{yr} \times \frac{\$0.02}{kwh} = \$331.$$

Capital expense of additional blower capacity ÷ $300.

Btu/hr transferred in preheater

Air goes from 130°F to 1200°F

$$\frac{531\ SCFM}{359} \times 60 \frac{Min}{hr} \times \frac{7.3\ BTU}{lb\ mole\ °F} \times (1200-130)\ °F = 694,000 \frac{BTU}{hr}$$

Heat released by cooling products to 500°F (computed for determination of heat capacity of this stream)

CO_2	13.96 [(1400-77) 11.45 - (500-77) 9.9]	=	153,100 BTU/hr
H_2O	55.74 [(1400-77) 8.92 - (500-77) 8.2]	=	465,000
N_2	70.2 [(1400-77) 7.35 - (500-77) 7.1]	=	472,500
O_2	3.50 [(1400-77) 7.75 - (500-77) 7.2]	=	25,300
Total Heat Released			1,115 ,900 BTU/hr

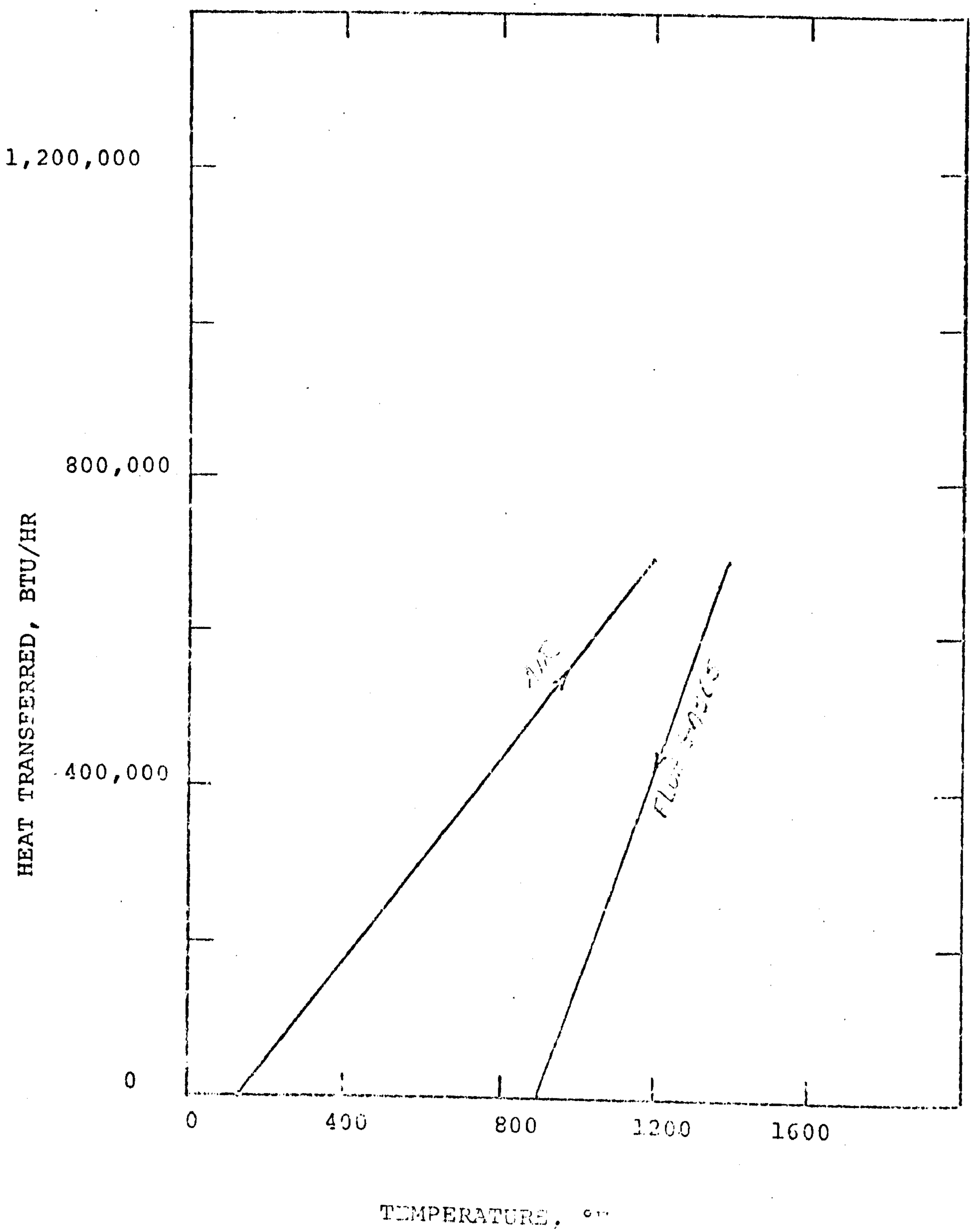


FIGURE [illegible] HEATER TEMPER[illegible] PROFILES

Using the attached graph, a stream with this heat capacity would only be cooled to 900°F.

Based on a previous quotation for a heat exchanger in similar service, the purchase price of this heat exchanger is estimated at $16,400 plus $3600 for installation. A cyclone would have to be added to protect the preheater from erosion. Because the incinerator itself could be smaller due to lesser air requirements, the savings in incinerator cost is estimated at $3000; however the plenum chamber and distributor would have to be designs for high temperature. The extra cost incurred for this would be about $6000. Net increase in incinerator cost is then $3000.

Net change in operating cost after preheater addition =

Fuel	$2730/yr
Electricity	331
	$3061/yr savings

Net change in investment after preheater addition =

Cyclone (refractory lined)	$ 5,000
Preheater	20,000
Blower	- 300
Incinerator	3,000
	$ 27,000 increase

Return on investment before taxes = $\frac{3061}{\$\ 27{,}000} \times 100 = 11\%$

This is the minimum ROI most companies would consider. The fact that the corrosion and erosion problems have been severe with known operating units would dissuade all companies from this investment for this particular plant.

ECONOMICS OF A WASTE HEAT BOILER INSTALLATION

Heat available when flue gases, as shown in the master calculation, are cooled to 500°F.

CO_2	17.77 $\frac{\text{lb moles}}{\text{hr}}$	195,000 BTU/hr
H_2O	58.89	491,000
N_2	95.90	645,000
O_2	4.75	34,000
		1,365,000 $\frac{\text{BTU}}{\text{hr}}$

This would be equivalent to

$$\frac{1,365,000}{118,000} = 11.6 \text{ gal/hr fuel for steam generation}$$

or $11.6 \frac{\text{gal}}{\text{hr}} \times \frac{\$0.10}{\text{gal}} \times 4000 \frac{\text{hr}}{\text{yr}}$ as a credit

A conventional industrial waste heat boiler could be purchased for about \$10,000. The erosion problem resulting from the carryover from fluid bed incinerators would require installation of a hot, refractory-lined cyclone at about \$5000.

Special corrosion problems known to exist from trace components with off gas from incinerators may raise this to \$25,000. Piping and installation, tying into the existing steam system, and instrumentation could bring this to \$35,000.

Operation of a boiler in accordance with local codes usually requires an operator or part of an operator if the boiler is located in the powerhouse. Assume the effective labor rate for boiler attendance is \$0.50/hour - This is a very optimistic rate and depends on the waste heat boiler being part of an installation of several standard boilers.

The estimated return on investment before taxes would then be (4000 hr/yr operation) :

$$\frac{\$4640/\text{yr} - \$2000/\text{yr}}{\$35,000} \times 100 = 7.5\%$$ which is extremely unattractive

V SUMMARY OF THE PROCESS DESIGN

Description of the Process

The process equipment flowsheet for the processing of paunch manure is shown in Figure 14 (Procedyne Corporation Drawing D-05149).

Paunch manure and fluid from the paunch sack (OA), plus water from the table and from sack washing operations (OB), currently runs by gravity to an existing paunch storage bin. A new screw conveyor (2700) will be installed in the bin to feed paunch to the slurry transfer pump (1100).

Stream 2A, from the transfer pump (1100), is fed to the primary dewaterer (1200), a screening device which separates the coarse solids in the paunch manure stream (stream 3A) from the free water and fines (stream 3B). That stream is fed into a dewaterer (1300) which discharges material with solids content raised to approximately 37% (stream 4A). Stream 4A is fed into the fluid bed incinerator (0100) via screw conveyor system (1400). Stream 4A contains approximately 2/3 of the solids to be incinerated.

Returning to the screening device (1200), the liquid stream 3B is fed into a settler (1500) where it is settled before becoming part of a feed stream to sand filter (1700). That feed stream (6A) is combined with the liquid stream (4B) from the dewaterer and transferred via transfer pump 1600 to filter 1700. Solids content of stream 6A + 4B is estimated to be approximately 4.5% and contains approximately 1/3 of the solids to be incinerated by the fluid bed incinerator 0100.

Filter 1700 is also fed with clean recycle sand overflowing from incinerator 0100 (stream 5B). Product stream 7A from the filter, containing approximately 15% solids (on a sand free basis) is fed to incinerator 0100 using screw conveyor system 1800. Makeup sand (stream 5A) is also added via conveyor 1800. Filtrate from the filter (stream 7B) is recycled back to settler 1500 and combines with other streams from the dewatering part of the process before entering the sewer system as overflow from the settler 1500 (stream 6B).

Fluidizing air (stream 5C), which also supplies oxygen for combustion, is fed to incinerator 0100 from blower 2400. Combined feed streams 4A and 7A are expected to contain 25-30% solids on a sand free basis, as well as the recycling sand used for filtration.

The system is expected to be self incinerating at approximately 30% solids content. Burner system 1000 is used to burn a small amount of oil (5D) to maintain the incinerator temperature at approximately 1500°F during periods of high moisture feed. A burner-fuel system is also required to start up the system, bringing the temperature of the incinerator to to at least 1300°F before starting a feed stream.

Stream 5A contains the products of combustion, CO_2 and water, and excess air required by the process, as well as a trace of fluidizing sand carried away from the incinerator by the combustion gases. That stream flows through a cyclone separator (1900) which removes a high proportion of the solids from the combustion gases, dropping them into portable ash bin (2000) via stream 8B.

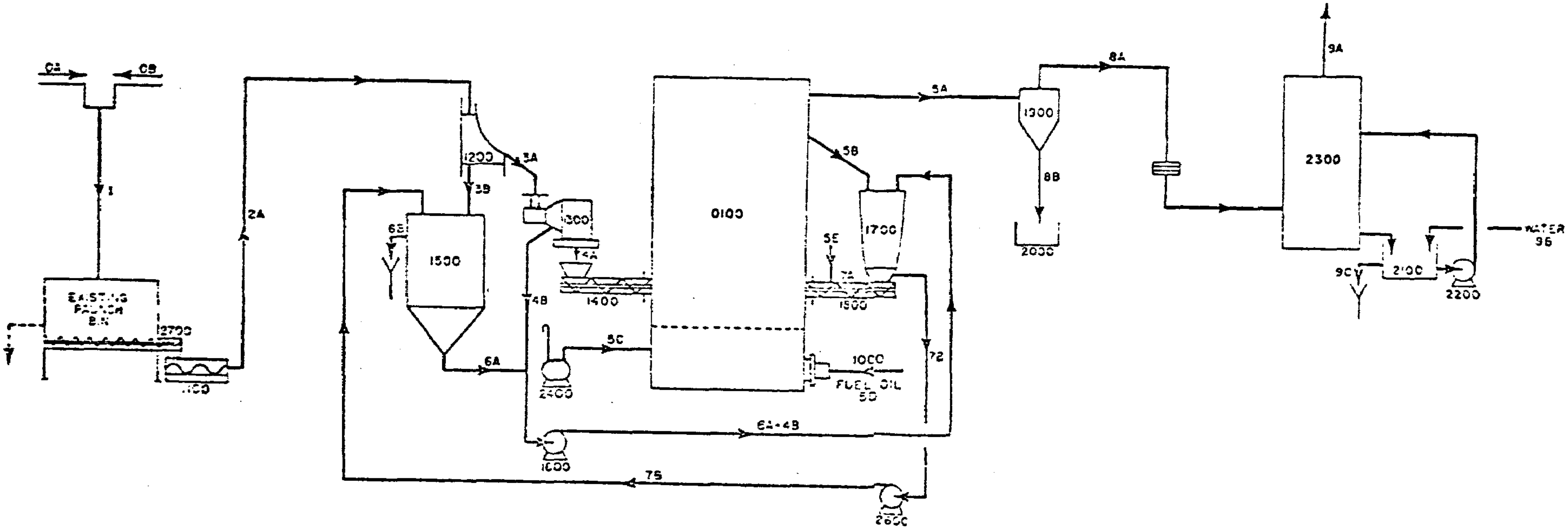
Exit gases leave the cyclone (1900) as stream 8A and are fed to a water scrubber (2300) for cooling and final removal of particulates before entering the atmosphere via stack 9A. Scrubber (2300) is part of a system which includes tank (2100) to which makeup water is added via stream 9B. Circulating pump 2200 circulates scrubbing water over scrubber (2300). Overflow water from tank (2300) enters the sewer via stream 9C. Solids, which accumulate by settling in tank (2100), are removed periodically as a wet ash.

Major Equipment List & Material Balance

The major equipment list for the process is as follows:

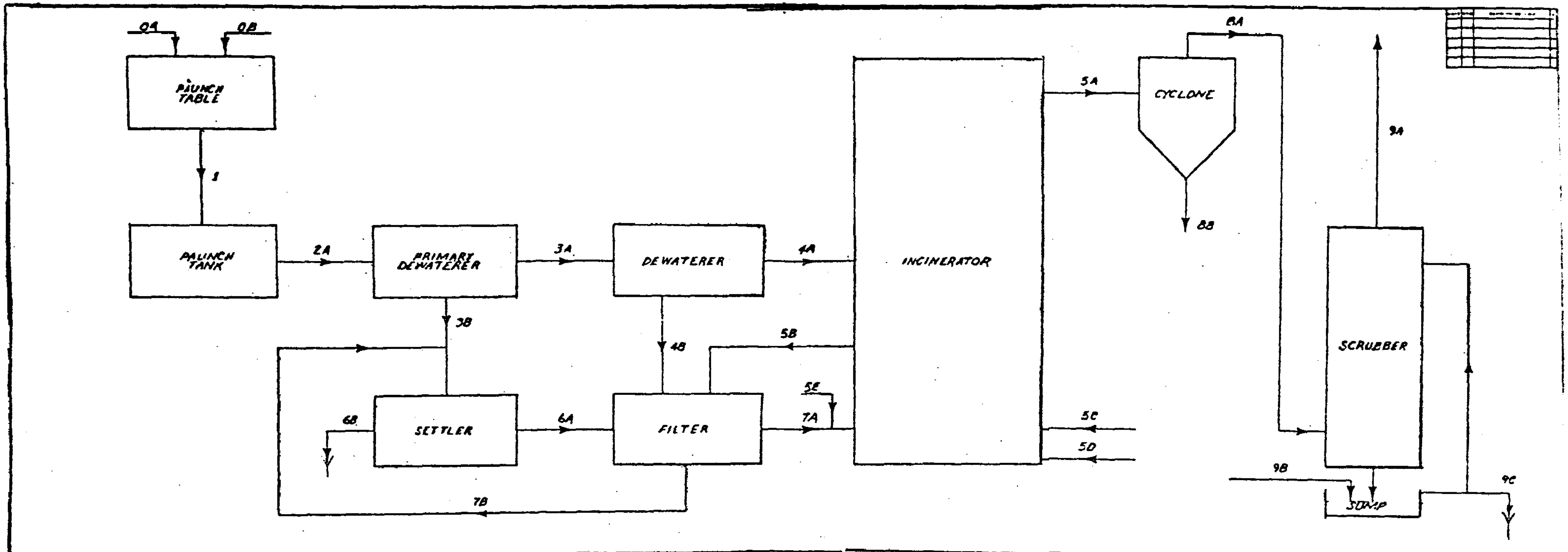
ITEM	DESCRIPTION
0100	Procedyne Fluid Bed Incinerator
1000	Burner
1100	Slurry Transfer Pump
1200	Primary Dewaterer
1300	Dewaterer
1400	Paunch Conveyor
1500	Settling Tank (Procedyne)
1600	Filter Feed Pump
1700	Procedyne Sand Filter
1800	Sand Conveyor
1900	Cyclone
2000	Ash Bin (portable)
2100	Scrubber Feed Tank
2200	Scrubber Feed Pump
2300	Scrubber
2400	Blower
2500	Panel Board (Procedyne)
2600	Settler Feed Pump
2700	Paunch Bin Conveyor

Figure 16 (Procedyne Drawing D-05141) gives the results of all calculations made on process flows. Streams described above and illustrated in Figure 15 are all defined in terms of flow rate. Data used for the calculations are described in Section II and calculations presented in Section IV.



EQUIPMENT LIST

0100	INCINERATOR
1000	BURNER
1100	SLURRY TRANSFER PUMP
1200	PRIMARY DEWATERER
1300	DEWATERER
1400	PAUNCH CONVEYOR
1500	SETTLING TANK
1600	FILTER FEED PUMP
1700	SAND FILTER
1800	SAND CONVEYOR
1900	CYCLONE
2000	ASH BIN (PORTABLE)
2100	SCRUBBER FEED TANK
2200	SCRUBBER FEED PUMP
2300	SCRUBBER
2400	BLOWER
2500	PANEL BOARD
2600	SETTLER FEED PUMP
2700	PAUNCH BIN CONVEYOR



STREAM	UNITS	OA	OB	1	2A	3A	3B	4A	4B	5A	5B	5C	5D	5E	6A	6B	7A	7B	8A	8B	9A	9B	9C
1 PAUNCH MANURE B.D.S.	#/HR	731(2)	–	731	365	292	73	241	51	–	–	–	–	–	69	37	67	33	–	–	–	–	–
2 WATER (BOUND)	#/HR	3582	–	3582	1791	1333	458	364	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
3 WATER (FREE)	#/HR	912	10,000	10,912	5456	–	5456	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
4 WATER (TOTAL)	#/HR	4494	10,000	14,494	7247	1333	5914	364	969	–	–	–	–	–	1625	6453	430	2164	–	–	–	3750	2165
5 SAND	#/HR	–	–	–	–	–	–	–	–	10	1352	–	–	10	–	–	1352	–	–	10	–	–	–
6 ASH	#/HR	–	–	–	–	–	–	–	–	26	–	–	–	–	–	–	–	–	11	15	4	–	7
7 TOTAL STREAM	#/HR	5225	10,000	15,225	7612	1625	5987	605	1020	36	1352	–	–	10	1694	6490	1849	2197	11	25	4	3750	2172
8 FUEL OIL	#/HR	–	–	–	–	–	–	–	–	–	–	–	87	–	–	–	–	–	–	–	–	–	–
	GALLONS/HR	–	–	–	–	–	–	–	–	–	–	–	11	–	–	–	–	–	–	–	–	–	–
9 AIR	#MOLES/HR	–	–	–	–	–	–	–	–	–	–	121.6	–	–	–	–	–	–	–	–	–	–	–
	SCFM	–	–	–	–	–	–	–	–	–	–	727	–	–	–	–	–	–	–	–	–	–	–
10 FLUE GASES	#MOLES/HR									177.34(4)											265.35		
11 CO_2	#MOLES/HR	–	–	–	–	–	–	–	–	17.77	–	–	–	–	–	–	–	–	–	–	17.77	–	–
12 O_2	#MOLES/HR	–	–	–	–	–	–	–	–	4.75	–	–	–	–	–	–	–	–	–	–	4.75	–	–
13 N_2	#MOLES/HR	–	–	–	–	–	–	–	–	95.90	–	–	–	–	–	–	–	–	–	–	95.90	–	–
14 H_2O	#MOLES/HR	–	–	–	–	–	–	–	–	58.89	–	–	–	–	–	–	–	–	–	–	146.90	–	–
15 SO_2	#MOLES/HR	–	–	–	–	–	–	–	–	0.03	–	–	–	–	–	–	–	–	–	–	0.03	–	–
16 TOTAL	SCFM	–	–	–	–	–	–	–	–	1040	–	–	–	–	–	–	–	–	–	–	1590	–	–

NOTES

1- FUEL OIL IS ASSUMED TO BE #2 FUEL OIL-1% MAX. SULFUR PER CHICAGO ENVIRONMENTAL CONTROL ORDINANCE (CHAPTER 17) - CITY OF CHICAGO.

2- BASED ON MAXIMUM FEED RATE OF 95 CATTLE / HR.

3- BASED ON ALLOWABLE PARTICULATE EMISSION - ENVIRONMENTAL CONTROL ORDINANCE - (CHAPTER 17) - CITY OF CHICAGO.

ILLINOIS PACKING CO.

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VI CONCLUSIONS

The results of the preliminary studies have led to a process design more complex than that originally proposed. The principal departure from the proposed process being the breakdown of the dewatering operation into two distinct steps. This split in the dewatering operation was necessitated by the radically differing properties of the course and fine fractions of the paunch material.

The project is presently in the engineering design phase for which the material presented herein forms the design basis. Results of the design phase will be presented in a subsequent interim technical report.

SECTION VII

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TECHNICAL REPORT DATA

(Please read Instructions on the reverse before completing)

1. REPORT NO.	2.	3. RECIPIENT'S ACCESSION NO.
4. TITLE AND SUBTITLE A METHOD OF MANURE DISPOSAL FOR A BEEF PACKING OPERATION		5. REPORT DATE 06/75 6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) Anon		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS Procedyne Corporation 221 Somerset Street New Brunswick, NJ 08903		10. PROGRAM ELEMENT NO. 1BB037 11. CONTRACT/GRANT NO. 12060 EOF
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15. SUPPLEMENTARY NOTES

16. ABSTRACT

The report contains the preliminary studies, process development, process calculations, and process design for a system to successfully handle the paunch manure in a beef slaughtering operation.

These studies resulted in a system in which the paunch manure is collected from the slaughtering operation and is fed to a screening device which separates the coarse solids. The screenings are dewatered to a solids content of 37 percent. This dewatered material is then sent to a fluid bed incinerator via a screw conveyor. The liquid stream from the screening is fed to a settler and is combined with the liquid stream from the dewaterer prior to sand filtering. The filter cake is fed to the incinerator using a screw conveyor. The filtrate is recycled back to the settler.

Work was completed as of February 1971.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Industrial Wastes, Industrial Waste Treatment, Incinerators, Dewatering Filtration, Meat	Meat Packing Industry, Paunch Manure Disposal, Fluid Bed Incineration	13B

18. DISTRIBUTION STATEMENT Unlimited; Available from NTIS, U.S. Dept. of Commerce, 5285 Port Royal Rd. Springfield, VA 22151	19. SECURITY CLASS *(This Report)* 20. SECURITY CLASS *(This page)*	21. NO. OF PAGES 91 22. PRICE

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