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MATHEMATICAL MODEL OF RECALCINATION OF LIME SLUDGE WITH FLUIDIZED BED REACTORS

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MATHEMATICAL MODEL OF RECALCINATION OF LIME SLUDGE WITH FLUIDIZED BED REACTORS

by

G. J. Ducar
P. Levin

General American Transportation Corporation
Niles, Illinois 60648

for the

FEDERAL WATER QUALITY ADMINISTRATION

DEPARTMENT OF THE INTERIOR

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TABLE OF CONTENTS

TABLE OF CONTENTS (CONT'D)

ABSTRACT

This report describes the development of a computer program to evaluate lime sludge recalcination systems. Data for the program was collected from a literature survey and field trips to operating installations of pulp mills and a water softening plant.

Some of the design relationships were found to have the same correlating variables for the recalciner as for the fluidized bed incinerators. Equations were developed for the least square curves which fitted the data best. These equations were used as the basis for the computer program developed to size some of the major components and to estimate capital, operating and maintenance costs for fluidized bed lime recalcination systems.

The fluidized bed reactor recalcining process has not yet been installed at any sewage treatment plant. Therefore, no data on the process was available. However, the pulp and paper plants and water softening plants presently recalcine in a manner which would be applicable to tertiary treatment recalcination requirements. Therefore the models developed herein represent the recalcination process as applied to tertiary treatment plants.

The computer program developed will be used (by FWQA) as a subroutine for the executive computer program entitled "Preliminary Design of Wastewater Treatment Systems." The executive program will be used to evaluate and optimize new wastewater treatment systems which are to be funded by FWQA.

This report was submitted in fulfillment of Contract #14-12-415, Program #17090 EHQ, between the Federal Water Quality Administration and General American Transportation Corporation.

INTRODUCTION

Lime (calcium hydroxide, $Ca(OH)_2$) is extensively used in both water and waste treatment processes. In combination with soda ash, it is used to reduce the hardness of water. Present indications are that it can be used economically to reduce the nitrogen and phosphorus content of wastewater.

The disposal of lime sludges produced in these water and wastewater treatment processes is becoming a serious problem due to the limited capacity of our environment to absorb such wastes. These sludges, once considered innocuous, now are considered as pollutants adversely affecting the environment. The requirements for higher degrees of treatment will require increased quantities of lime at wastewater treatment plants, presenting increased disposal problems for the increasing volumes of sludges produced.

The most ideal solution available for disposal of this sludge is to convert it economically into a usable product. This can readily be accomplished by heating it to about 1600°F and decomposing the sludge to calcium oxide and rehydrating or slaking the calcium oxide for recycling to the process. This processing is commonly referred to as recalcining.

Various types of reactors are available to accomplish this, including rotary kilns, multiple hearth and fluidized bed incinerators.

The purpose of this program was to develop a mathematical model of a fluidized bed reactor for recalcining calcium carbonate sludge. The model is composed of equations which represent the calcination equipment and process. This model should be of assistance to FWPCA in sizing reaction vessels and related equipment, the amount of fuel and power required, and estimates of the capital, operating, and maintenance costs for fluidized bed reactors used for lime recalcination. The necessary equations were developed from data obtained from the literature and from field surveys of plants which recalcine calcium carbonate produced in the manufacture of wood pulp and the lime-soda ash water softening process. No plants are currently in operation which recalcine lime from primary or tertiary treatment processes in sewage treatment plants. These systems closely resemble the equipment which could be used in sewage treatment applications in the near future.

Origin of Lime Sludges

The addition of lime to raw sewage or secondary effluents is an effective procedure for reducing phosphorus concentrations. The residual dissolved phosphate level declines as a function of the pH to which the wastewater is treated, with about 0.15 mg/l P remaining at a pH of 11. The elevation to pH 11 also results in conversion of the NH_4^+ to NH_3 which can be removed in an air stripping tower. The sludge precipitated either in the primary tank or tertiary plant contains calcium carbonate sludge with both organic and chemical impurities.

Similarly, in a typical water softening process, sufficient lime is added to elevate the pH to 11 and precipitate the $Mg(OH)^2$ and $CaC\emptyset_3$.

Subsequent recarbonation of the effluent or softened water at a pH of 11 depresses the pH to approximately pH 9.5, precipitating an essentially pure (99%+) $CaC\emptyset_3$ which can be effectively recalcined.

DESCRIPTION OF SYSTEMS ANALYZED

Terminology

A schematic of the reactor or combustion chamber, which is the heart of the fluidized bed recalcining system, is shown in Figure 1. The parameters related directly to the reactor are shown and the various regions of the reactor are labeled on Figure 1. The lower section of the reactor consists of a fluidized air plenum, an air distribution grid and a pellet cooling section called the cooler. The center section of the reactor is the calcining section containing the fluidized CaØ pellet bed. Above the calcining section is the freeboard. The terminology shown on Figure 1 is specifically related to the reactor vessel. However, auxiliary equipment was considered on this program. Therefore additional terminology was required and this terminology is given in Appendix A.

System Components and Operation

The system components included in this study are shown in Figure 2. The recalcined lime handling system (not shown in Figure 2) and the components shown in Figure 2 are included in the capital costs developed on this program. The layout shown is the complete fluidized bed recalcining system which is usually purchased as a package. This system was used as the basis for including equipment in the study because cost data collected included all of the components shown in Figure 2 plus the lime handling system. Since it was not possible to obtain costs for the individual components it was necessary to develop cost information for the total system. The costs developed on this program do not include land costs for recalcining equipment or ash disposal or land costs for storage of wasted $CaCØ_3$ for a non-recalcining installation. Slaking lime loss rates were not included in this program.

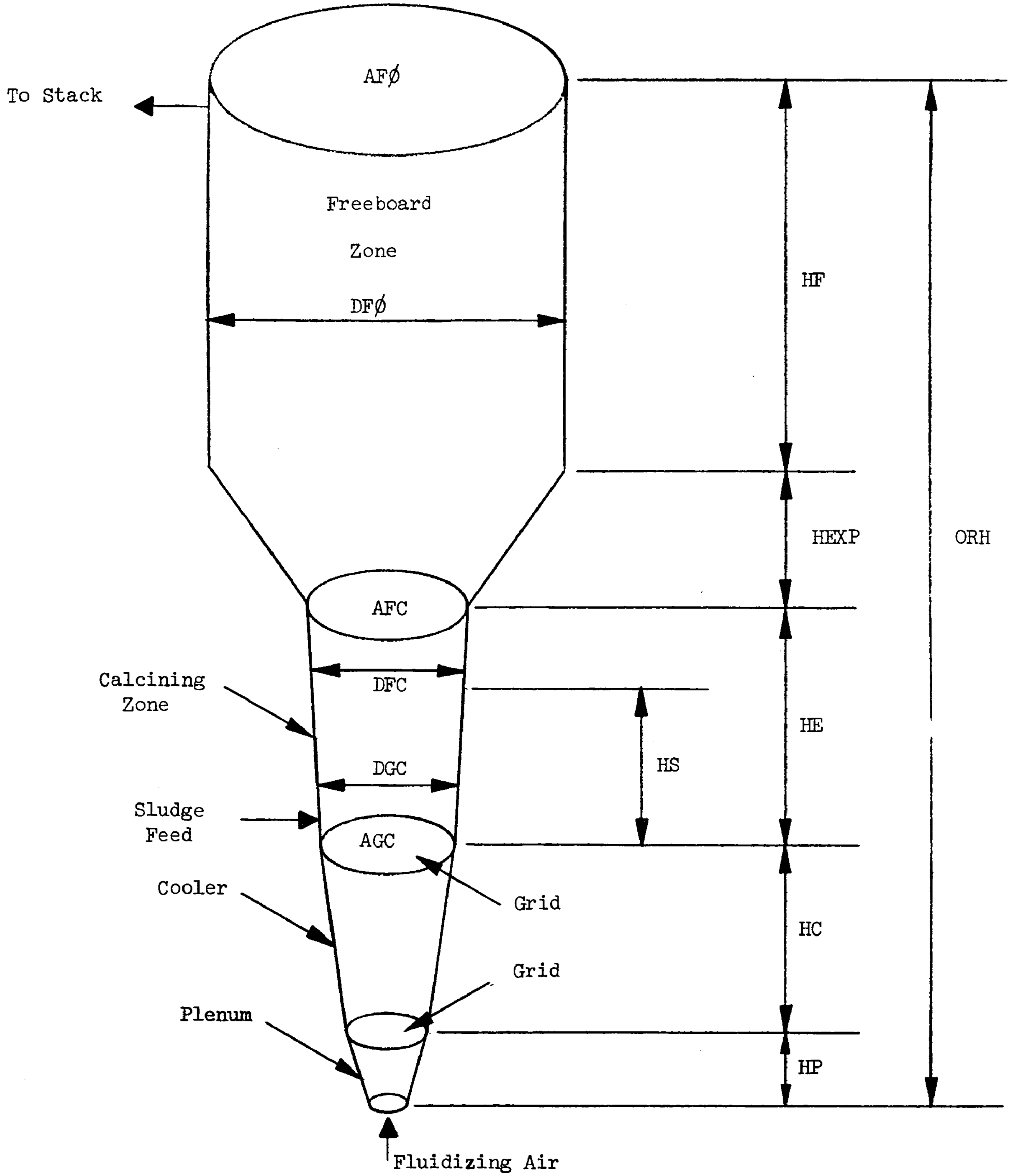

Figure 1 REACTOR TERMINOLOGY

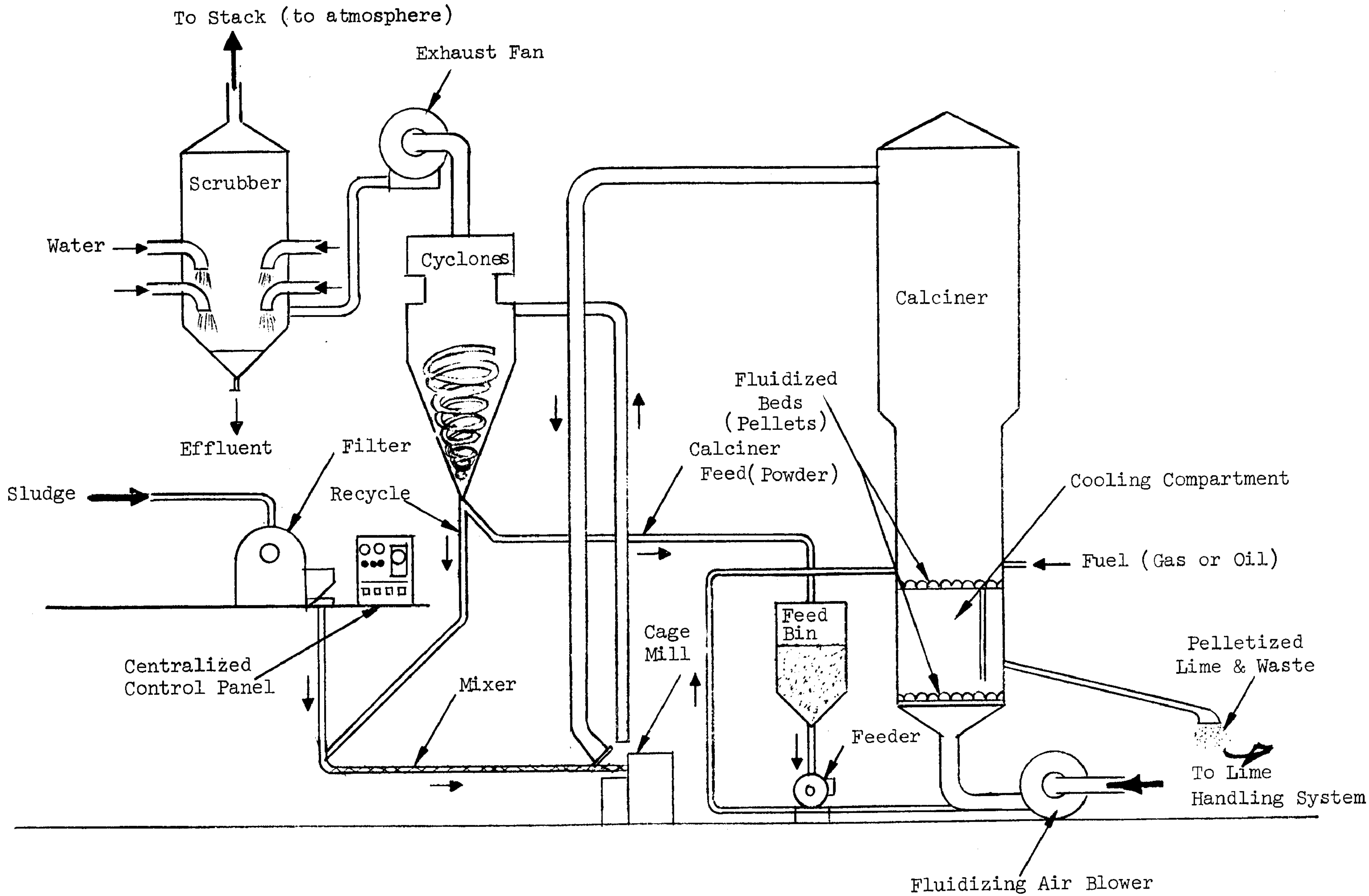

Figure 2 SCHEMATIC OF RECALCINATION SYSTEM

Calcining Process Description

Since no primary or tertiary lime recalcination-sludge combustion systems were in operation at the time of this report the process was not available for analysis. However, the pulp and paper and water softening recalcining processes closely resemble the tertiary and primary recalcining process which could be used in the future. The calcium carbonate sludge is initially dewatered either by a centrifuge or vacuum filter to a cake with a solid content of 60 to 65 percent. This cake is then passed through a mixer and cage mill where it is mixed with dried recycle $CaC\emptyset_3$ and hot stack gases containing fine $CaC\emptyset_3$ particles. The stack gases are usually diluted with air or quenched with water to reduce the gas temperature entering the cage mill to a level which will not affect the cage mill. The mixture is ground and dried in the cage mill and the particulate matter is carried with the gases to the primary cyclone.

Very fine material $CaC\emptyset_3$ passes through the primary cyclone to the secondary cyclone. The large particles are either sent to the reactor and mixed with soda ash for recalcination and particle growth or recycled through the cage mill (75% are recycled). The finer particles entering the secondary cyclone pass through and are collected in the scrubber. The heavier particles entering the secondary cyclone are sent to the feed bin and then to the reactor for recalcination or particle growth.

Soda ash is added to the feed stream to promote pelletization of $Ca\emptyset$ particles in the reactor. It is a necessary additive which melts to a sticky material in the reactor and coats the surface of $Ca\emptyset$ particles causing them to stick together. The cost of soda ash is $33/ton (1969 price less freight) and the normal feed rate for pulp and paper plants is approximately 0.5 to 1.0% by weight of the sludge leaving the dewatering device.

The dried $CaCO_3$ sludge is fed into the fluidized CaO bed in the calcining section of the reactor with fuel to make the endothermic reaction proceed. The material remains in the reactor at a temperature of approximately 1620°F. (recalcination occurs at approximately 1520°F.) until it is blown up and out through the freeboard zone by the fluidizing gases (fine particles) or it falls through a pneumatic level sensitive valve into the cooler zone. The freeboard zone prevents larger particles from leaving the reactor with the exhaust bases. The particles which have grown to pellet size, from continued agglomeration accelerated by the soda ash, enter the cooler section at 1620°F. and are cooled to approximately 400°F. by the fluidizing air which passes up through the cooler. The fluidizing air is preheated to approximately 400°F. in the cooler.

The pelletized lime is removed from the cooler through a pneumatic level control valve. The lime is pelletized because it is more convenient to handle in a pellet form, rather than a powder. Also, since the particles forming the pellet, for the most part, are recycled through the reactor several times more complete recalcination is assured than if the finer particles were used as the final product. For proper fluidication in the recalcining section a particular particle size range must be maintained.

Since the CaO particles continually grow in the reactor it is necessary to replenish the fine particle supply in the reactor periodically. This is done by crushing some of the stored pelletized lime and feeding it to the reactor.

This particle addition is also necessary when the recalcining section becomes loaded with impurities which must be dumped to waste.

The input stream to be used for the computer program developed is the feed material $CaCO_3$ in a dry (zero water content) state. This is equivalent to the composition of the material leaving the cage mill in a dried state. All of the water in the feed is evaporated in the cage mill by the hot exhaust gases.

Method of Approach

The method of approach used for this study was to:

1) Review the literature to establish the availability of published data concerning recalcination by fluidized bed reactors;
2) Visit representative pulp mills and water softening plants to obtain data on fluidized bed reactors in use today;
3) Develop a computerized procedure for evaluating lime sludge recalcination systems assuming the systems at the pulp mills and water softening plants closely resemble the equipment which could be used in future sewage treatment applications.

By necessity, the evaluation procedures are limited in their application to the range of the parameters disclosed by the literature search and facility visits. Also, the available funding limited the amount of facility visits.

The facility visits were conducted at:

1) A pulp mill of Kimberly-Clark at Reading, California which was not visited directly, but data on it were obtained by a personal visit to the corporate engineering offices located in Neenah, Wisconsin,

2) A water softening plant at Ann Arbor, Michigan,

3) A pulp mill of P. H. Glattfelter at Spring Grove, Pennsylvania,

4) A pulp mill of S. D. Warren Company at Muskegon, Michigan, and

5) A water softening plant at Lansing, Michigan.

A review of the literature found no recalcination units treating sludge from wastewater treatment plants. There were recalcination units utilized to recalcine $CaCO_3$ sludge produced in the manufacture of wood pulp and the lime-soda ash water softening process.

One commercial process which reportedly employs addition of lime to the primary tank suggested that this lime can be recovered by dewatering and incinerating the primary sludge and slaking the resulting ash to recover the lime; however, our literature study did not reveal any such plants even in the design stage. We were also unable to find any plants calcining sludge containing calcium phosphate. The recalcination of this material would result in a build-up of non-volatile impurities in the CaO produced.

In this study, considerable data were obtained in the recalcination of essentially pure calcium carbonate by visiting paper plants which produce this type of sludge. Our model is therefore based on the recalcination of essentially pure $CaCO_3$. At present, this appears to be a realistic approach since at the Advanced Waste Treatment Plant in operation at Blue Plains we were advised that none of the lime produced from impure $CaCO_3$ was recycled to the system. It was the operator's belief that this material would be utilized more effectively as agricultural lime because of its phosphate content. Only the essentially pure $CaCO_3$ produced in the settling basin following recarbonation would be recycled to the multiple hearth furnace they were utilizing for recalcination.

RESULTS

Data Collected

Table I is a summary of the data collected on the recalcination systems studied during this program. This data was obtained by visits to installations and a survey of literature. The costs presented in Table I are the costs actually paid for the equipment. These costs can be compared by applying a cost index relating purchase dates.

The equipment sizing and capital cost data are reliable since they were obtained from visual inspection and available drawings. However, the data for operating costs (fuel, power and operator), though fairly accurate, were not readily available. The gross costs ($/ton of recalcined lime) were available. These costs are functions of operating efficiencies which were in various stages of improvement at the various plants. Note that there is no reason that the fluidized bed calciner cannot be used in smaller sizes.

Since the plants surveyed were purchased at different times, a standard cost reference date of February, 1968 was chosen for this program. All local unit prices were corrected to the national average of February, 1968 prices, using the cost index of the Department of the Interior as found in Engineering News Record magazine.

In Table I, Location 4, the design capacity rating 45/70 means that the unit has a design capacity of either 45 or 70 TPD depending on the diameter of the firebrick. The unit capacity can be upgraded to 70 TPD by removing firebrick, thus increasing the bed diameter.

Equations Developed

Equations were developed based on data gathered from the field and several useful publications. The data were plotted using various correlations and least square curves were plotted. Equations were developed for the best correlations found relating the variables.

TABLE I

FIELD DATA

ITEM	LOCATION 1 Kimberly Clark	2 Ann Arbor, Mich.	3 P. H. Glatfelter	4 S. D. Warren	5 Lansing, Mich.
Purchase Date	Jan 63	1965	Nov 65	Feb 63	1953
Startup Date	Oct 64	Feb 67	Nov 66	Dec 63	1954
Design Tons Dry CaØ/Day	50	24	128	45/70	30
% CaØ in Reactor Product	88	-	85	85	88-91.5
Fuel Type	No. 6 Fuel Oil	No. 4 Fuel Oil	No. 6 Fuel Oil	No. 6 Fuel Oil	No. 6 Fuel Oil
Number of Sludge Feed Points	9	6	24	12	-
Type of Feed	Pneumatic	Pneumatic	-	Pneumatic	-
Lime Recovered % of Feed	-	-	90	94	90
Total System Cost, $	$200,000	-	$475,206	$575,000	-
Number of Bed Burners	10	12	12	-	8
Fluidizing Blower, HP	125	100	200	300	-
Fluidizing Blower, SCFM	4790	2100	3800	10,000	-
Fluidizing Blower, inches $H_2Ø$	194	194	166	-	-
Bed Temperatures, °F	1600	1580	1600	1650	-
Freeboard Temperatures, °F	1650	1495	1550	1600	-
Sludge Feed Point Distance above Grid, ft	1.5-2	1.5	-	-	-
Cooler Section Included	No	Yes	Yes	Yes	Yes
Bed Warmup Time, hrs	24	-	-	8	-

TABLE I (CONT'D)

FIELD DATA

ITEM	LOCATION 1	2	3	4	5
Bed Cooling Rate	30°F/hr	25°F/hr	-	-	-
Major Problems	Slaking	Slaking	Pellet transfer to cooler. Reactor scaling	Pellet transfer to cooler. Reactor scaling	-
Inside Diameter of Calcining Section, ft	8	6	8/10	11	6
Inside Diameter of Freeboard Section, ft	14	8.5	16.5	-	-
Static Bed Depth, ft	6	6	7	7	7
Expanded Bed Depth, ft	9.8	9.0	10.5	10.0	10.5
Expanded Bed Freeboard, ft	15	16.5	16.7	16.7	-
Total Installed HP	416	345	465	1048	460

Sizing the Reactor

A recalcination system is rated on its output design capacity of CaØ. For a given output design capacity (CAP), the total dry solids which must be handled by the system is

$$PDS = CAP\ (2000.)/\ (x)(y) \qquad (1)$$

Note that there are 2000. pounds per ton. The weight of CaØ contained in one pound of $CaCØ_3$ is termed x, and x = 56/100 = 0.56. The weight of $CaCØ_3$ contained in one pound of dry solids entering the reactor is defined as y and this must be determined for the particular sludge being recalcined. Field data obtained were for systems as small as PDS = 4000. Therefore, because of the uncertainty of curves such as for ØHP and IHP, it was decided to limit the lower limit of reliability of the data to PDS = 4000. Similarly, since no fluidized bed reactor was found with a bed grid diameter (DGC) in excess of 16 feet (Kansas City fluidized bed incinerator), this was taken as the upper limit of structural achievement using standard construction materials.*

Figure 3 shows a plot of the net heat input to the reactor required in excess of the endothermic heat of the $CaCØ_3 \rightarrow CaØ + CØ_2$ reaction.

To obtain the net heat input (QCAØ = Btu/ton CaØ), the endothermic heat of the reaction (2.69×10^6 Btu/ton of CaØ) was subtracted from the total gross heat input to the reactor (fuel, dried sludge and air preheat in cooler). The result shown in Figure 3 is

$$QCAØ = 5.0 \times 10^6 \qquad (2)$$

To obtain the total net heat input to the reactor, the net heat input per ton of CaØ is multiplied by the design CaØ capacity of the reactor.

$$WUMQN = QCAØ\ (CAP) \qquad (3)$$

*Dorr-Oliver has informed the authors that they have industrial calciner units in operation with diameters of 45 feet.

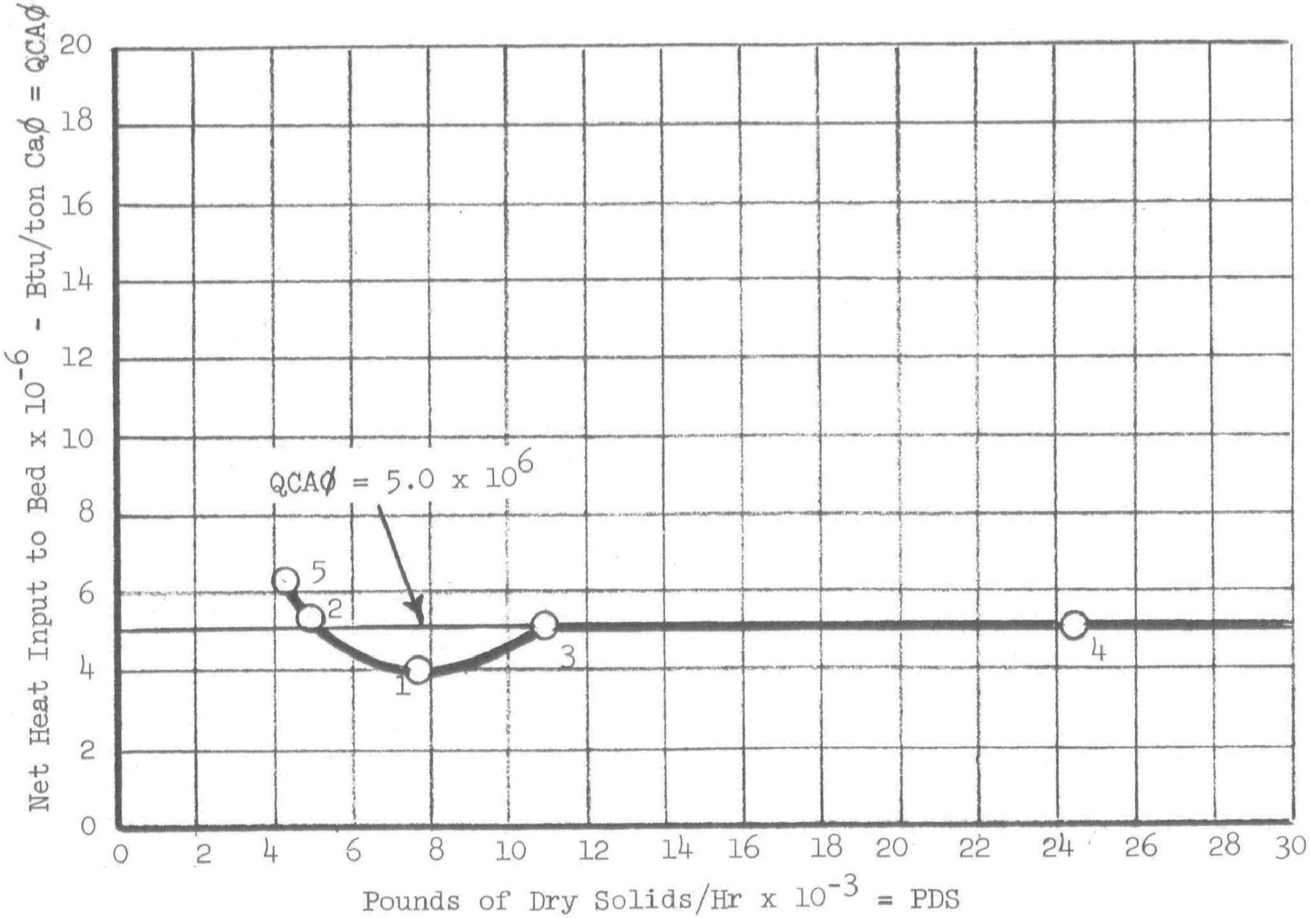

Figure 3 NET BTU/TON CaØ VS POUNDS DRY SOLIDS/HR

Then similarly

$$\text{END}\emptyset \;=\; 2.69 \times 10^{6}\ (\text{CAP}) \tag{4}$$

and the gross heat input to the reactor is

$$\text{SUMQ} \;=\; \text{SUMQN} + \text{END}\emptyset \tag{5}$$

It is possible that more than one reactor may be needed for a given installation. This is because of the maximum sizing limit of 16 feet diameter in the calcining grid section (DGC). When the required recalcining capacity exceeds the capacity of a reactor with DGC = 16 feet more than one reactor is required for the installation. This is handled by specifying a sufficient number of reactors (ZII) of equal calcining grid diameter (DGC), to handle the required capacity. Thus the net heat input to each reactor is

$$\text{SUMQI} \;=\; \text{SUMQN}/\text{ZII} \tag{6}$$

Each reactor handling SUMQI will have a calcining grid cross-sectional area (Figure 4) of

$$\text{AGC} \;=\; \text{SUMQI}/241{,}000 \tag{7}$$

The diameter of the calcining grid is

$$\text{DGC} \;=\; \left[4.0\ (\text{AGC})/\pi\right]^{1/2} \tag{8}$$

with a maximum limit of DGC = 16.0 feet.

The pounds of dry solids per hour which must be handled by each reactor in the system is

$$\text{ZI} \;=\; \text{PDS}/\text{ZII} \tag{9}$$

Since the feed material enters the reactor dry, because of the evaporation process in the cage mill, the reactor calcining section cross-sectional area does not vary vertically as does the fludized bed incinerator handling wet

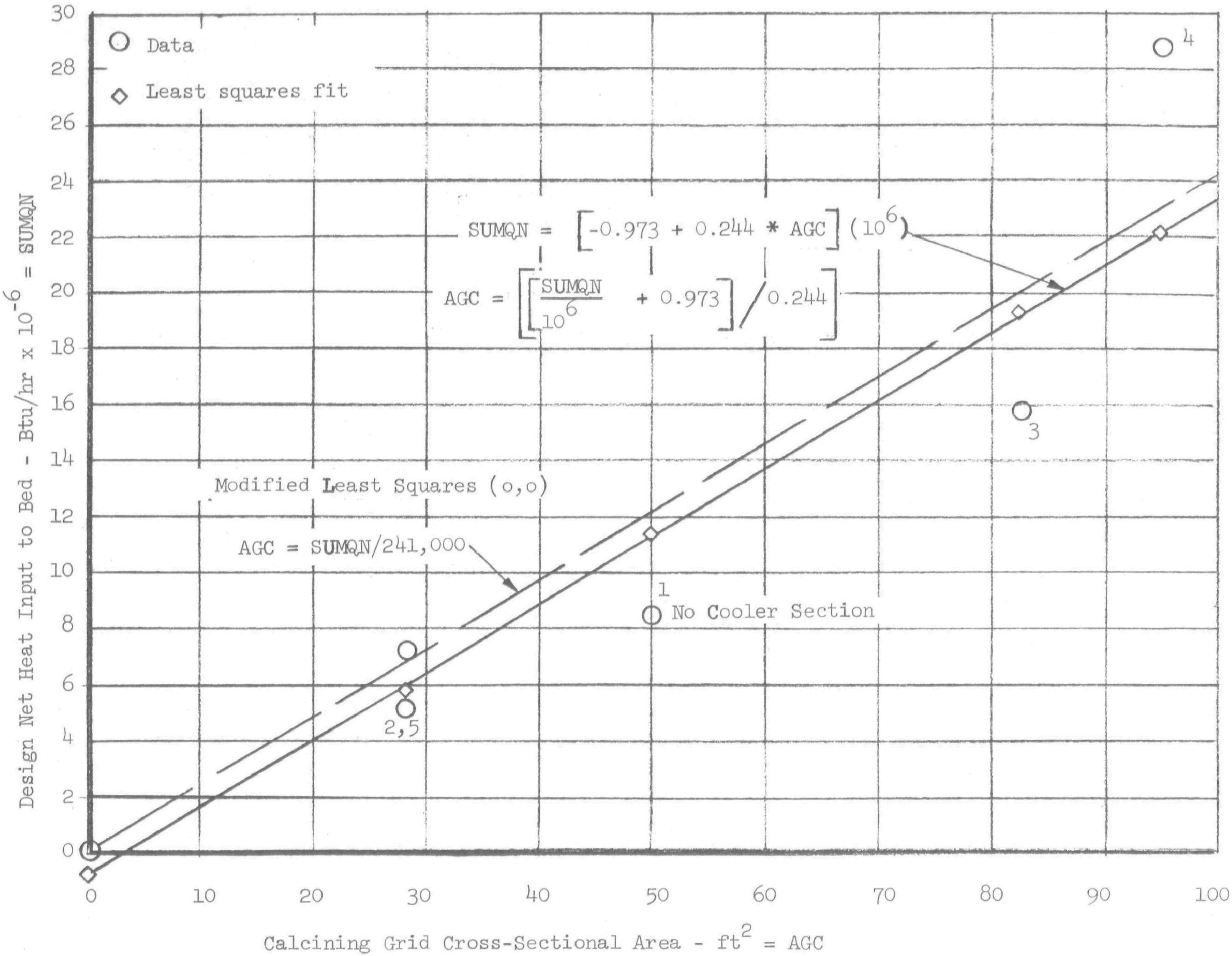

Figure 4 DESIGN NET HEAT INPUT TO BED VS CALCINING GRID CROSS-SECTIONAL AREA

sewage sludge. Therefore

$$DFC = DGC \tag{10}$$

for the recalciner.

Figure 5 shows the relationship determined for the freeboard diameter located above the cylindrical recalcining zone

$$DF\emptyset = (DGC - 1.5)/0.49 \tag{11}$$

The cross-sectional area of the freeboard zone is

$$AF\emptyset = \pi\ (DF\emptyset)^2/4.0 \tag{12}$$

The fluidized bed in a recalciner is composed of Ca∅ pellets too large to be carried into the freeboard by the exhaust gases and too small to be dropped into the cooler section.

The expanded volume (VE) of the fluidized bed in the recalcining zone of each reactor is (Figure 6)

$$VE = SUMQN/23{,}700 \tag{13}$$

where SUMQN is changed to SUMQI when more than one reactor is used in a system. Note that the least squares fit curve was not used for this relationship but a slightly modified parallel line was used passing through the origin (o,o). This eliminated a negative net heat input at zero bed volume which is physically unreasonable.

It has been found that the expanded fluidized bed volume in reactors is about 50% greater than the static bed volume so

$$VS = VE/1.5 \tag{14}$$

Field data obtained indicated an expanded bed depth (HE) of from 9 to 10.5 feet. Most of the data showed a bed depth of 10 feet

$$HE = 10.0 \tag{15}$$

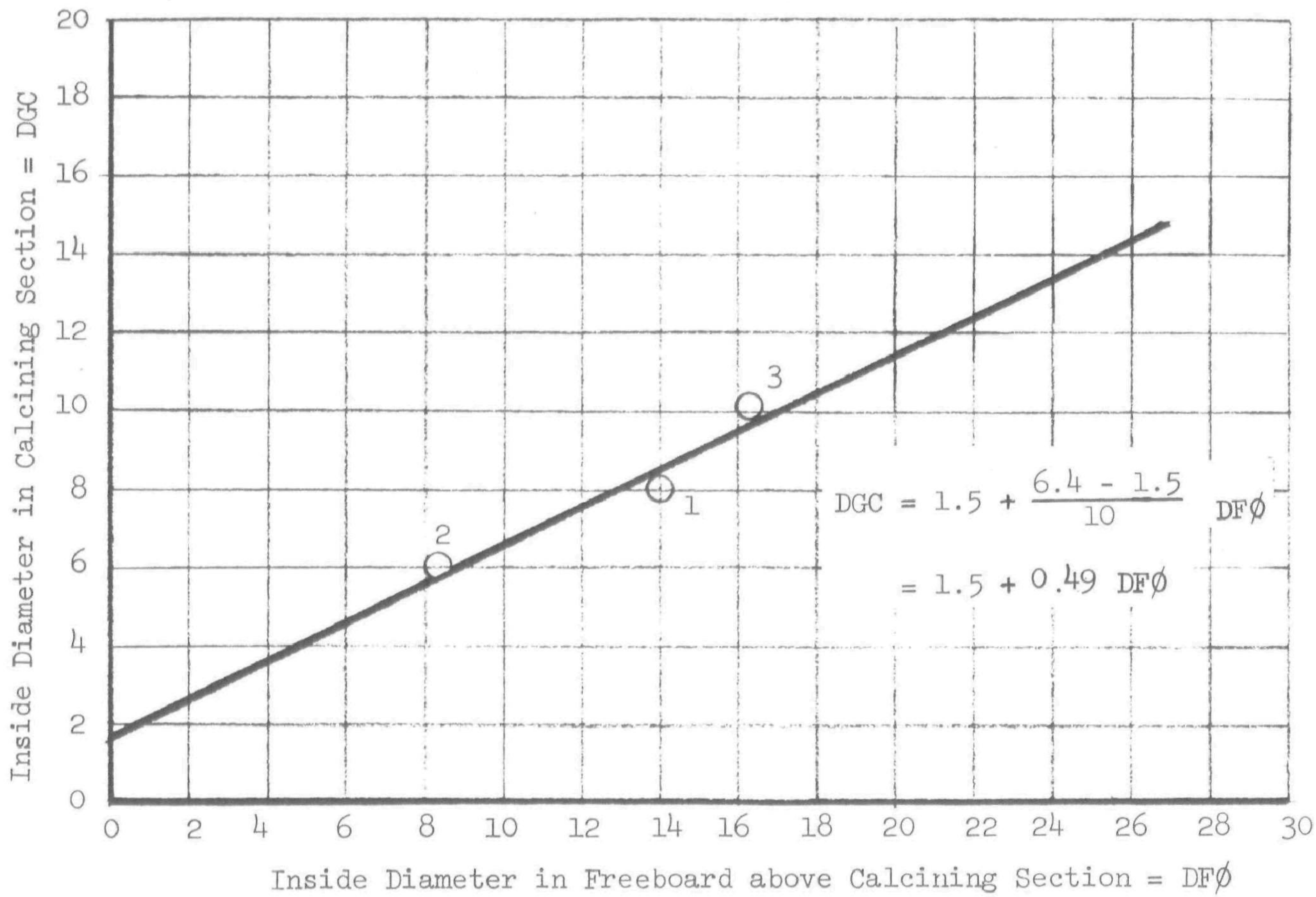

Figure 5 INSIDE DIAMETER IN CALCINING SECTION VS INSIDE DIAMETER IN FREEBOARD ABOVE CALCINING SECTION

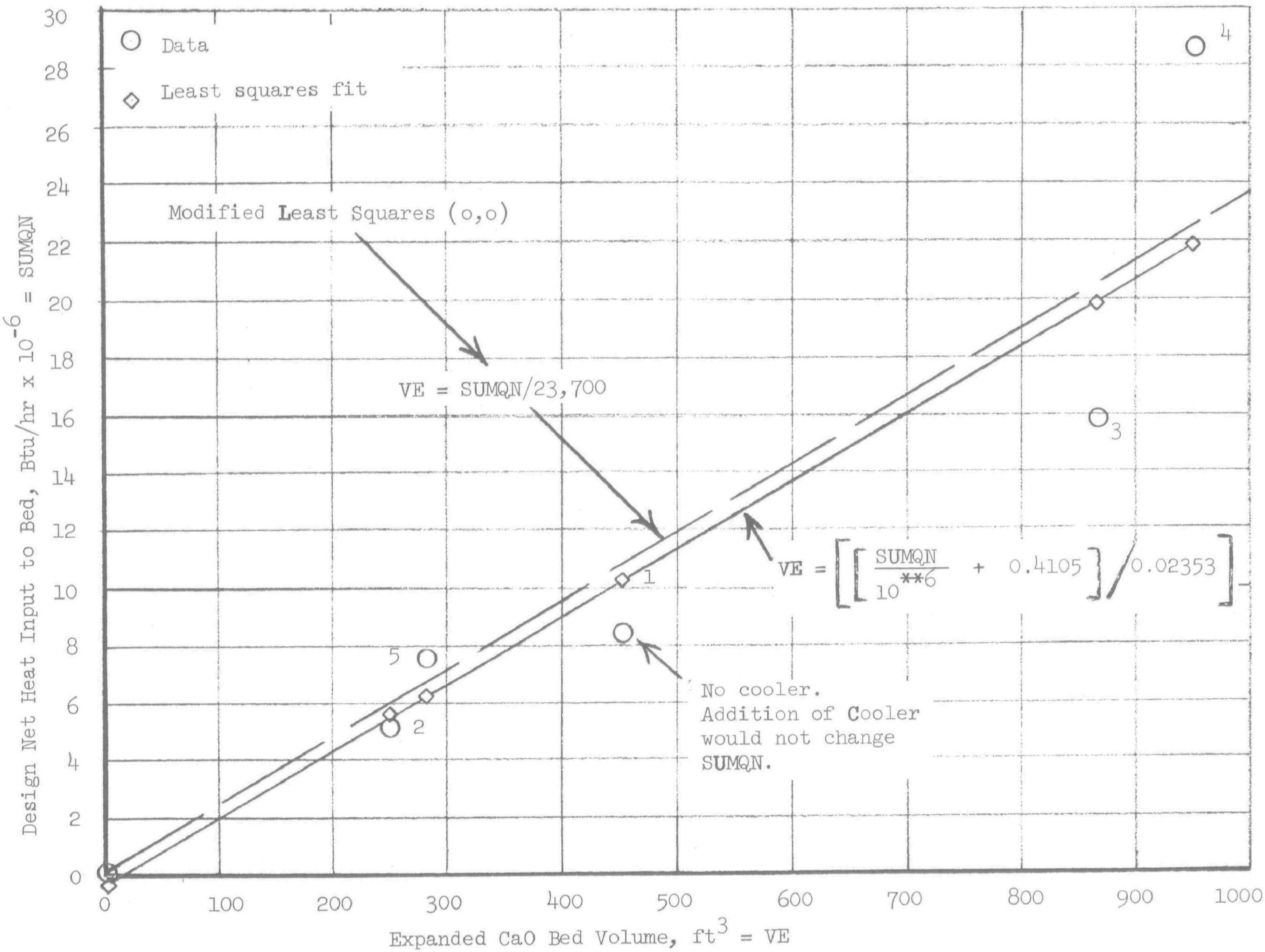

Figure 6 DESIGN NET HEAT INPUT TO BED VS EXPANDED BED VOLUME

For 50% bed expansion on fluidization the static bed depth is

$$HS = HE/1.5 \tag{16}$$

The following physical dimensions were estimated as

$$HEXP = 0.5 \tag{17}$$

$$HP = 3.0 \tag{18}$$

The height of the cooler section (HC) did not vary significantly with unit capacity and the relation

$$HC = 8.5 \tag{19}$$

was found to approximate most of the data.

Figure 7 shows the freeboard height to be closely approximated by

$$HF = 16 \tag{20}$$

Summing the heights of the individual sections of the reactor, the overall reactor height can be closely approximated by

$$ØRH = HP + HC + HE + 0.5 + HF + HEXP \tag{21}$$

Note that the 0.5 feet added to the ØRH term represents the approximate distance between the height of the expanded fludized bed (HE) and the beginning of the transition section between the recalcining and freeboard zones.

Figure 8 shows the relationship between the number of feed points in the reactor and the design dry solids feed rate to be

$$NFP = 2.3 + 0.00087\ (PDS) \tag{22}$$

where PDS is changed to ZI when more than one reactor is required in the system. This number must obviously be rounded-off to the nearest integer value.

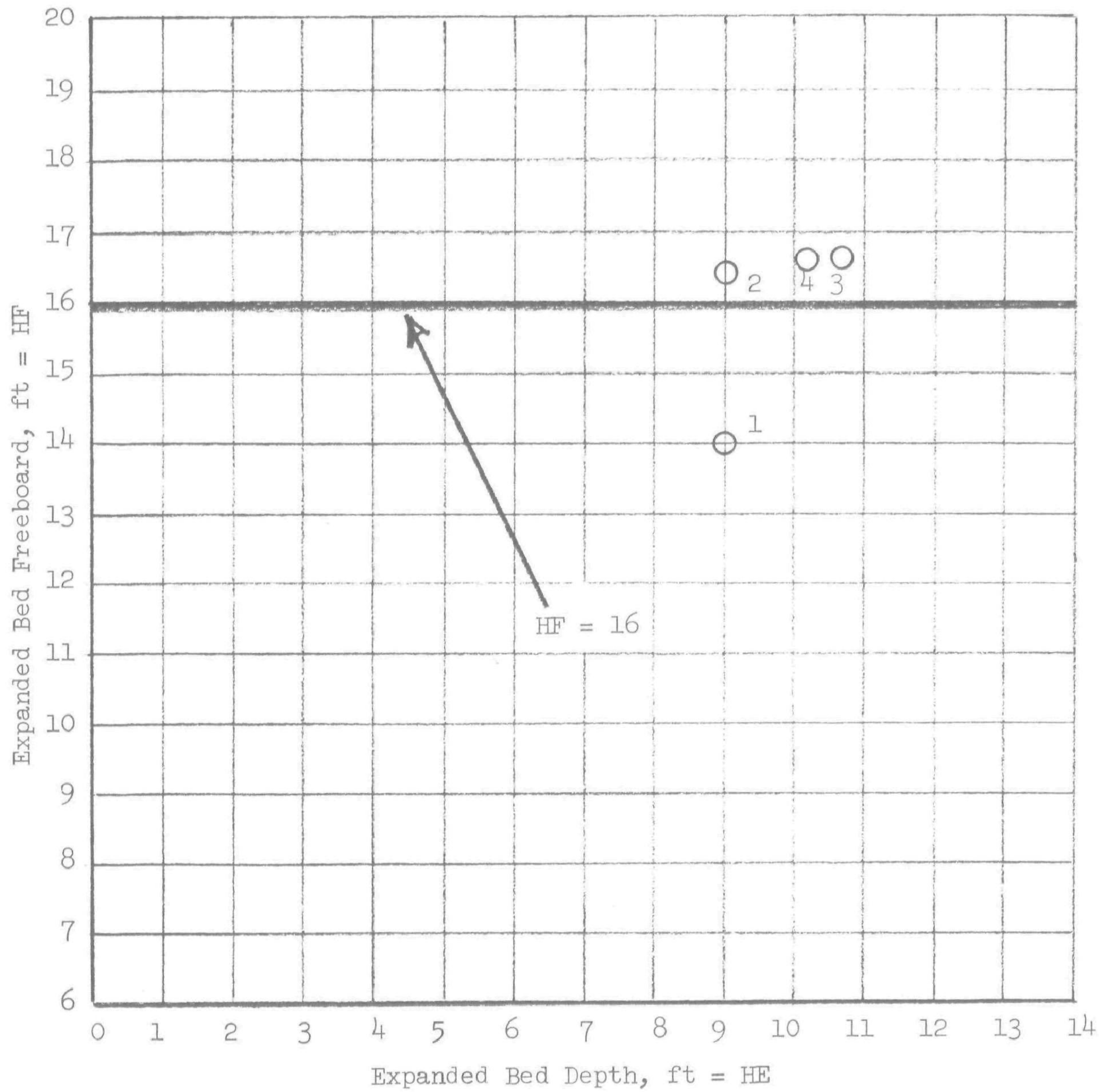

Figure 7 EXPANDED BED FREEBOARD VS EXPANDED BED DEPTH

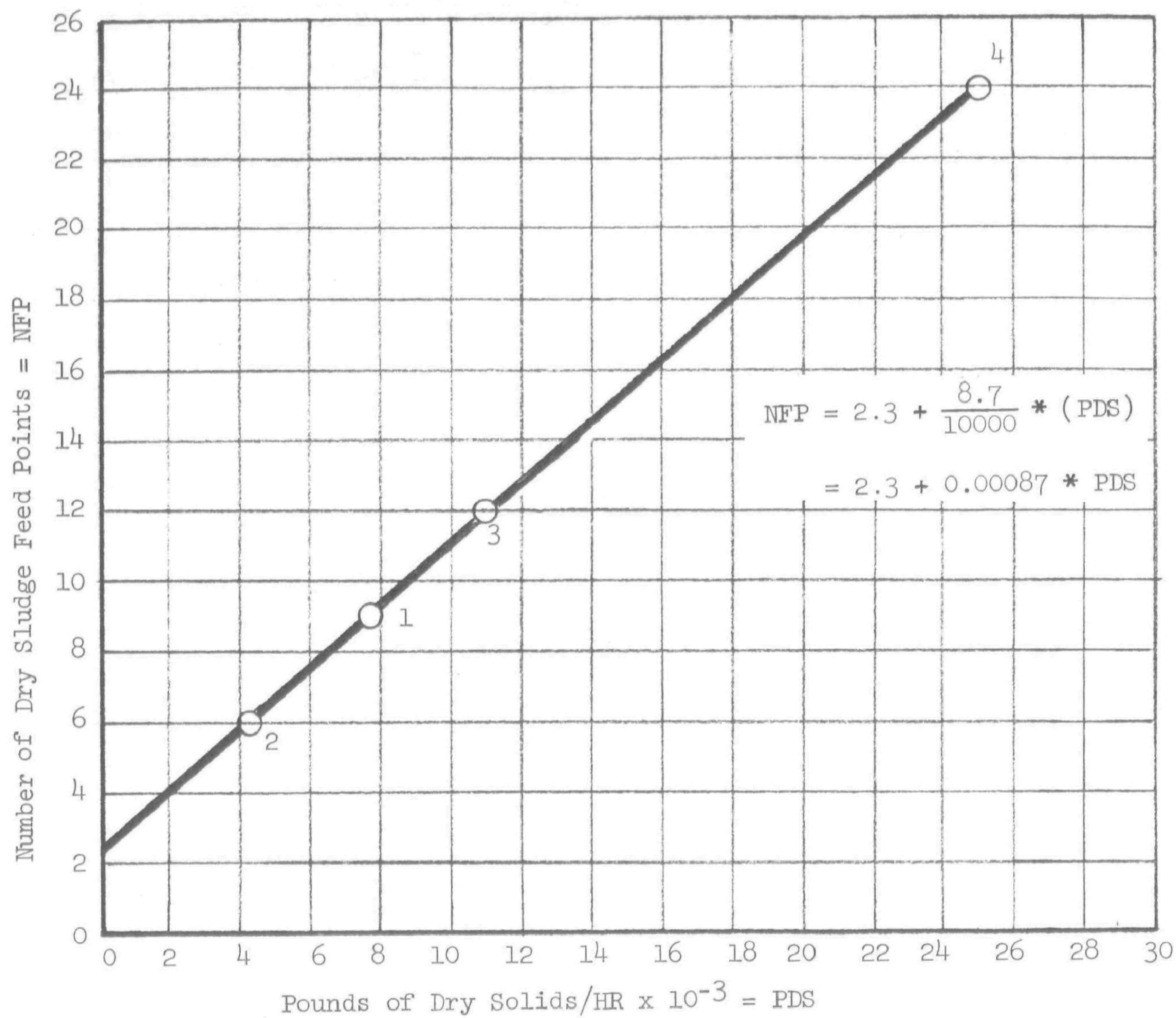

Figure 8 NUMBER OF DRY SLUDGE FEED POINTS VS POUNDS OF DRY SOLIDS PER HOUR

System Electrical Power

The installed and operating horsepower for each reactor in the system are plotted in Figure 9. Least squares equations fitting the data almost perfectly are

$$HPIZI = 320.0 + 2.44\ (ZI/1000) + 1.067\ ((ZI/1000)^2) \qquad (23)$$

$$\emptyset HPZI = 321.55 + 3.39\ (ZI/1000) + 0.523\ ((ZI/1000)^2) \qquad (24)$$

The total system installed and operating horsepower are given by

$$HPI = HPIZI(ZII) \qquad (25)$$

$$\emptyset HP = \emptyset HPZI(ZII) \qquad (26)$$

Electrical power costs are given by

$$CP = \emptyset HP\ (EPC)\ (0.746)\ (PCTY)\ (8760) \qquad (27)$$

where 0.746 is the conversion factor for converting horsepower to kilowatts and 8760 is the number of hours in 365 days or one year.

Lime Loss Rates

Lime can be lost from the recalcining process by incomplete recovery in the centrifuge, blowdown of the recalciner CaO bed because of impurities or stack losses. The sum of these losses is

$$SLIM = CAP \left[\left| (PERW + DEWAL)/100 \right| + W \right] \qquad (28)$$

The reactor bed blowdown losses (PERW) of Ca$\emptyset$ are required when inert impurities build up in the system and affect fluidization of the Ca$\emptyset$ bed in the reactor and system performance. No data was available on blowdown losses, however, these losses should be very small if essentially pure $CaC\emptyset_3$ (99%+ is fed).

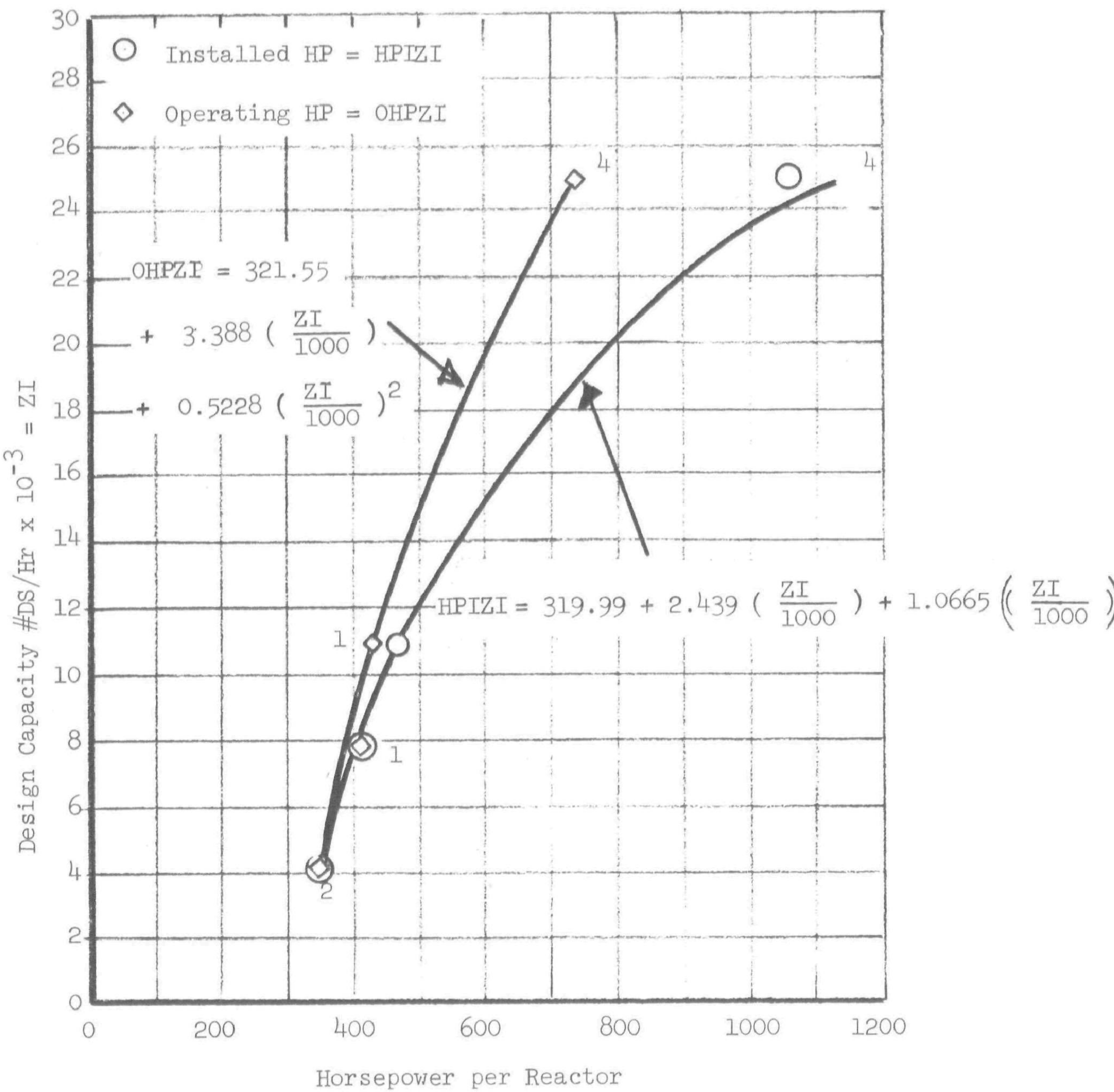

Figure 9 DESIGN CAPACITY VS HORSEPOWER

Dewatering losses (DEWAL) of $Ca\emptyset$ run in the range of 20% (9). This is a function of the dewatering device used.

Stack losses (W) of $Ca\emptyset$ are very low in a recalcining system because of the multiple cyclone arrangement and high efficiency scrubber used. Stack losses at the one installation with data were 25 pounds of $Ca\emptyset$ per day or 0.0001786 pounds $Ca\emptyset$ lost per pound $Ca\emptyset$ recovered. Based on this it is reasonable to assume that the stack losses on a recalciner are negligible. For example, translating the above loss to yearly cost for lime at $20/ton gives $91.25/year. Slaking losses do occur, but such data was not available.

<u>Ash</u>

The recalcination process will produce some inert ash material which must be disposed of. This will include $Ca_3(PO_4)$ and other non-volatile materials. For a reactor feed rate of PDS (pounds of dry solids/hr) the reduction in solids through the process will be caused by $C\emptyset_2$ generated from the reaction ($CaC\emptyset_3 \longrightarrow Ca\emptyset + C\emptyset_2$) and by $Ca\emptyset$ recovery from the other solids by slaking. The $C\emptyset_2$ generation rate from $CaC\emptyset_3$ is

$$C\emptyset_2 = 2000\ (CAP)\ (44/56) \tag{29}$$

where the fraction (44/56) accounts for the fact that there are 44 pounds of $C\emptyset_2$ per 56 pounds of $Ca\emptyset$.

Ash formed from the fuel oil or natural gas used to supply heat to the reactor is negligible. Therefore the total ash which must be handled and disposed of is

$$\begin{aligned} PAPH = PDS - CAP(2000) - C\emptyset_2 + (W + PERW/100)(2000)(CAP) \\ - PDS(PVS)/100 \end{aligned} \tag{30}$$

where CAP (2000) is the $Ca\emptyset$ recovery rate and $C\emptyset_2$ is the gas generation rate from the $CaC\emptyset_3$ reaction to $Ca\emptyset + C\emptyset_2$. The fourth term in Equation (30) is the rate of $Ca\emptyset$ loss from the system which must be treated as ash and the fifth term is the rate of combustion of the volatile solids fed to the reactor in the dry solids.

Reactor Heat Balance

The heat transferred to the fluidizing air from the pelletized lime in the cooler section is

$$QAPR = CAP(2000)(.217)(1650 - 400) \tag{31}$$

This equation states that the $Ca\emptyset$ discharged through the reactor has a specific heat of 0.217 Btu/pound °F and is cooled from 1650°F to 400°F by the fludizing air. The fluidizing air is heated to approximately 400°F in the cooler.

Each reactor requires heat from a fuel (usually number 6 fuel oil) to recalcine the lime. The heat from a fuel required by each reactor in the system is

$$QF\emptyset = \left[SUMQ - QSL(PDS) - QAPR \right] /ZII \tag{32}$$

Thus the fuel must supply the gross heat required by each reactor in the calcining zone less the heat rate supplied by the volatiles in the feed to the reactor less the heat recovered by the fluidizing air in the cooler.

System Costs

Electrical

Electrical power costs are given by Equation (27).

Fuel

Fuel costs for each reactor in the system is given by

$$CF\O = WF\O(PCTY)(8760)(FC/10^6) \tag{33}$$

and fuel costs for the system of ZII reactors is

$$CF\O I = CF\O\ (ZII) \tag{34}$$

Operating and Maintenance

The yearly cost of operating labor per year is

$$CL = 0.25(CLR)(PCTY)(ZII)(8760) \tag{35}$$

for the system. This is based on observations of the operating procedure used for both fluidized bed incinerators and recalciners. The average operator must be available at all times in case of emergencies. However, the operation of a fluidized bed system is almost completely automated. Thus an operator need only spend about one hour in every four operating hours checking instrumentation and making minor adjustments. The remainder of the time the operator is available for other duty. Equation (35) was developed with the assumption that one operator per reactor is needed for 25% of the operating time. It is entirely possible that one operator could handle four reactors in a system unless an emergency arose in which case another man would be required. The optimum labor force for this type of equipment should be one operator for every two to four reactors. The operator should be used for routine preventive maintenance in his spare time. At least one maintenance personnel should be used for each 4 to 8 reactors. Required yearly maintenance labor costs for the system (ZII reactors) are estimated as

$$CLM = CLMR(PCTY)(8760)(ZII)/8 \tag{36}$$

Note that Equations (35) and (36) account for only the labor required to keep the reactors running properly. No attempt is made to estimate how efficiently personnel are used. For example, if an operator is required for two hours per day it is assumed that he is doing other useful work during the remaining six hours and not charging idle time to the operating cost. If this is not true for a particular installation then these equations must be modified to reflect the true labor situation.

Maintenance on specialized equipment such as instrumentation will usually be done by a service organization rather than plant personnel. It is estimated that one man hour of service labor is required for every 150 hours of reactor operation. Therefore the yearly system cost of maintenance service labor is estimated as

$$CLMA = CLMS(PCTY)(8760)(ZII)/8 \tag{37}$$

The cost of operating a recalcining system for one year is

$$C\emptyset = CL + CP + CF\emptyset I \tag{38}$$

Yearly maintenance costs for the recalcining system are

$$CM = CLM + CLMA + CER \tag{39}$$

where CER is the cost of equipment which must be replaced such as thermocouples and other minor parts. The cost of equipment replacement was low although costs were not available. Therefore CER was assumed as $1 per day or CER = 365 (ZII).

Makeup Lime and Solids Disposal

Without a recalcining system the yearly cost of new lime and the cost of solids disposal ($CaC\emptyset_3$ plus other solids included in PDS) is

$$CLIME = (CCA\emptyset(CAP) + DC\ (PDS/2000))(PCTY)(8760) \quad (40)$$

When a recalciner is used, the yearly cost of lime makeup to the system and ash disposal can be expressed as

$$CLMUW = (CSLIM + ASHDC)(PCTY)(8760) \quad (41)$$

where

$$CSLIM = SLIM\ (CCA\emptyset) \quad (42)$$

and

$$ASHDC = DC\ (PAPH)/2000 \quad (43)$$

Equipment Capital Costs

Figure 10 shows the cost relation developed from the available field data as

$$EC = 0.0219\ (SUMQN) + 37100.0 \quad (44)$$

where EC is the cost adjusted to February, 1968 dollars. The cost adjustments were made using the local Sewage Treatment Plant Cost Index of the Department of the Interior and bringing the costs to local values for February, 1968. Then the local values were adjusted to the national average values using the respective Sewage Treatment Plant Cost Index.

Installation and Consulting Fees

Based on available information, an installation fee of 10% must be added to the equipment cost[1]. Also a consulting engineers fee of 10% of the sum of the system equipment cost plus installation fee must also be added[1]. Therefore

$$CI = EC/10 \quad (45)$$

$$CEF = (EC + CI)/10 \quad (46)$$

1. Data from a conversation with Mr. Stewart Peterson, N.E. Region Program Director, FWPCA, Boston, Mass.

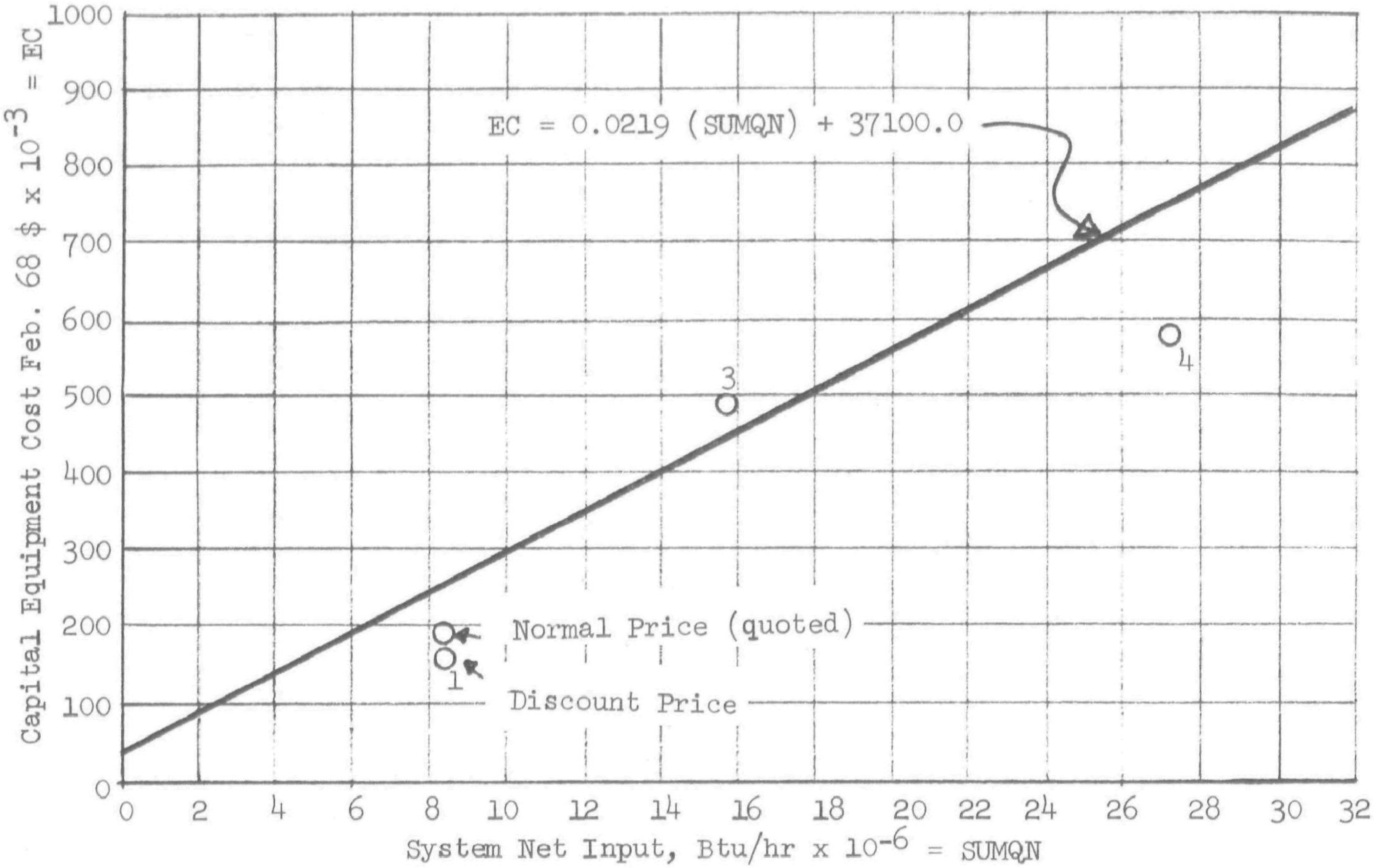

Figure 10 CAPITAL EQUIPMENT COST VS SYSTEM NET INPUT TO BED

Equipment Cost adjusted from purchase date and location of installation to Feb., 1968 and national average Sewage Treatment Plant Cost Index of the Department of the Interior. This Index is presented quarterly in the Engineering News-Record.

and the total system capital cost is

$$TSC = EC + CEF \tag{47}$$

in February, 1968 dollars. This cost can be adjusted to any year by using the national average Sewage Treatment Plant Construction Cost Index of the Department of the Interior. This index appears quarterly in the Engineering News Record Magazine.

Total Cost of Maintenance and Operation

The total yearly system maintenance and operating cost is

$$TCM\emptyset = CM + C\emptyset \tag{48}$$

The total yearly cost of maintenance, operation, lime makeup and solid waste (ash) disposal with a recalciner, from the inlet of the dewatering device through the exhaust stack is

$$TCLIW = TCM\emptyset + CLMUW \tag{49}$$

COMPUTER PROGRAM

Listing of Program

The complete program developed for use as a subroutine is listed below in Fortran IV.

Flow Chart

The flow chart for the computer program developed is shown in Figure 11.

```
// FOR
*IOCS(CARD,1132PRINTER)
*LIST ALL
      REAL NFP
C        LIME RECALCINATION IN FLUIDIZED BED REACTOR
   10 FORMAT(1H0,'COSTS DO NOT INCLUDE BUILDINGS OR LAND REQUIRED FOR EQ
     1UIPMENT')
  111 FORMAT(1H0,'COSTS DO NOT INCLUDE LAND FOR STORAGE OF WASTES')
   11 FORMAT(2F10.4)
   12 FORMAT(1H0,' SYSTEM IS TOO SMALL TO BE ACCURATELY SIZED BY THIS PR
     1OGRAM')
   13 FORMAT(1H.,' CASE',I3)
   14 FORMAT(5F10.4)
   15 FORMAT(1H0,8(5F16.4,//,1X))
   16 FORMAT(1H0,6F16.4)
      I=0
  100 READ(2,14 ) X,Y,PCTY,FC,EPC
      READ(2,14 ) CLR,CLMR,CLMS,DC,PERW
      READ(2,14 ) CAP,QSL,YRS,CCAO,W
      READ(2,11 )DEWAL,PVS
      I=I+1
      WRITE(3,13 )I
      ZII=1.
      PI=3.14159
      PDS=CAP*2000./(X*Y)
      IF(PDS-4000.) 110,120,120
  110 WRITE(3,12 )
      GO TO  180
  120 QCAO=5.*10.**6
      SUMQN=QCAO*CAP
      ENDO=(2.69*10.**6)*CAP
      SUMQ=SUMQN+ENDO
  130 SUMQI=SUMQN/ZII
      AGC=SUMQI/241000.
      DGC=(4.*AGC/PI)**.5
      IF(DGC-16.) 150,150,140
  140 ZII=ZII+1.
      GO TO  130
  150 ZI=PDS/ZII
      DFC=DGC
      DFO=(DGC-1.5)/.49
      AFO=(PI*DFO**2)/4.
      VE=SUMQI/23700.
      VS=VE/1.5
      HE=10.
      HS=HE/1.5
      HFXP=.5
      HP=3.
      HC=8.5
      HF=16.
      ORH=HP+HC+HE+.5+HF+HEXP
      NFP=FLOAT(IFIX(2.3+0.00087*ZI+0.500001))
      IF(NFP) 160,160,170
```

```
160 NFP=1.
170 HPIZI=320.+2.439*(ZI/1000.)+1.0665*(ZI/1000.)**2
    OHPZI=321.6+3.39*(ZI/1000.)+0.523*(ZI/1000.)**2
    HPI=HPIZI*ZII
    OHP=OHPZI*ZII
    CP=OHP*EPC*0.746*PCTY*8760.
    SLIM=CAP*(((PERW+DEWAL)/100.)+W)
    CO2=2000.*CAP*44./56.
    PAPH=PDS-CAP*2000.-CO2+(W+PERW/100.)*2000.*CAP-PDS*PVS/100.
    QAPR=CAP*2000.*.217*(1650.-400.)
    QFO=(SUMQ-QSL*PDS-QAPR)/ZII
    CFO=QFO*PCTY*8760.*FC/10.**6
    CFOI=CFO*ZII
    CL=.25*CLR*PCTY*ZII*8760.
    CLM=CLMR*PCTY*8760.*ZII/8.
    CLMA=CLMS*PCTY*8760.*ZII/150.
    CO=CL+CP+CFOI
    CER=365.*ZII
    CM=CLM+CLMA+CER
    CLIMF=(CCAO*CAP+DC*PDS/2000.)*PCTY*8760.
    ASHDC=DC*PAPH/2000.
    CSLIM=SLIM*CCAO
    CLMUW=(CSLIM+ASHDC)*PCTY*8760.
    EC=0.0219*SUMQN+37100.
    CI=EC/10.
    CEF=(EC+CI)/10.
    TSC=EC+CI+CEF
    TCMO=CM+CO
    TCLIW=TCMO+CLMUW
    WRITE(3,10 )
    WRITE(3,111)
     WRITE(3,15 ) X,Y,PCTY,FC,EPC,CLR,CLMR,CLMS,DC,PERW,CAP,QSL,YRS,CC
   1AO,PDS,ZII,SUMQN,SUMQ,DGC,ZI,DFO,VE,ORH,NFP,HPIZI,OHPZI,CO2,PAPH,Q
   2APR,QFO,CFO,CFOI,CP,CL,CLM,CLMA,CO,CER,EC,TSC,CM
    WRITE(3,16 ) CLIME,CLMUW,TCLIW,W,DEWAL,PVS
180 PAUSE 1111
    CALL DATSW(4,MORE)
    GO TO ( 100, 190),MORE
190 CALL EXIT
    END
```

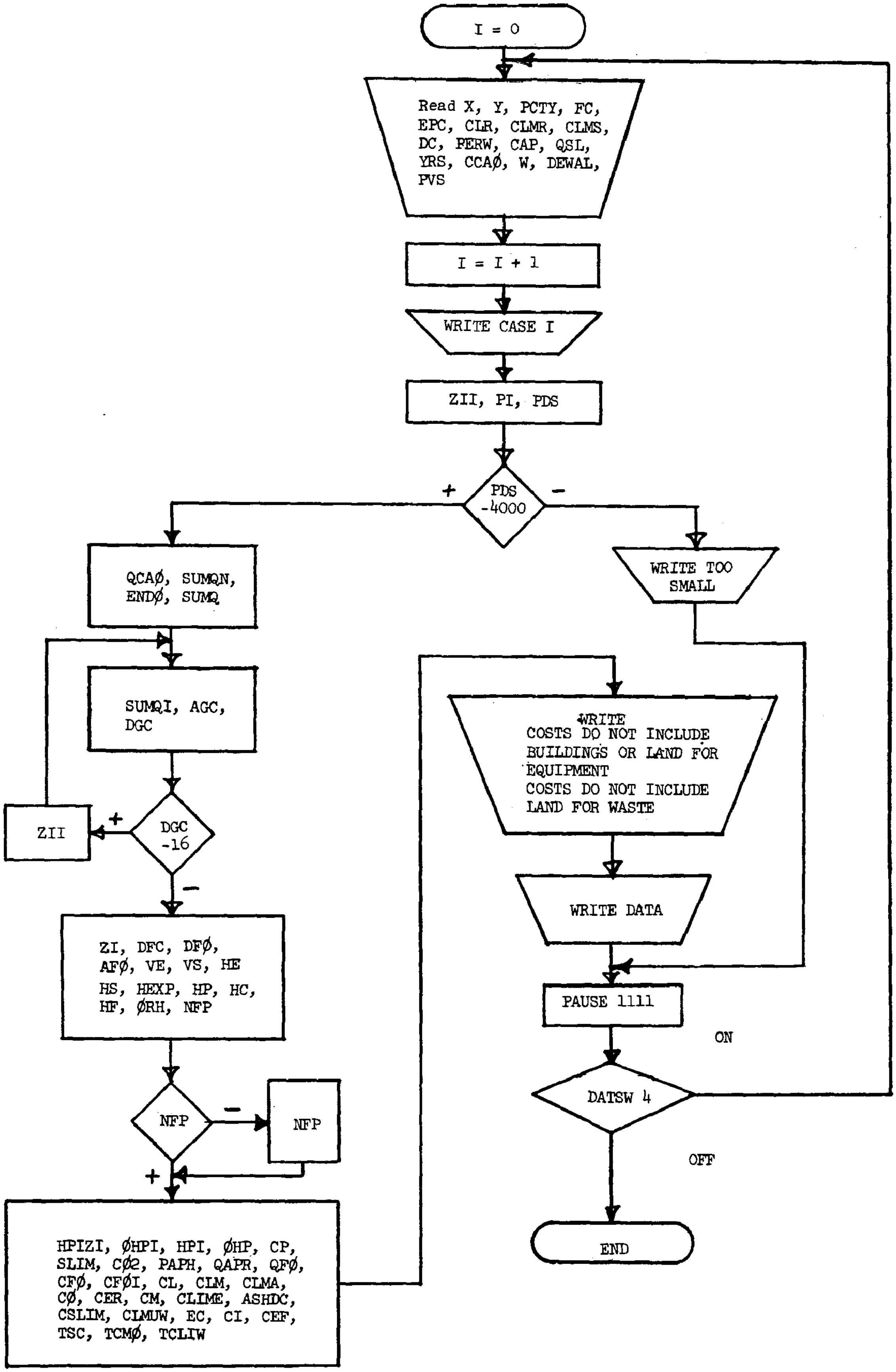

Figure 11 FLOW CHART

How to Use Program

Operation

The operation of the computer program requires an input of four data cards. Samples of these data cards are shown in Figure 12. Also data switch 4 must be set OFF for normal machine exit and ON for additional data input (i.e. cases 1,2,3, etc.). Nothing else is required of the operator. The input stream to be used is the material leaving the dewatering device but with zero water content (since water is completely removed in the cage mill).

The output data generated by the input data in Figure 12 is shown in Figure 13.

Reactor Sizing Limit

As with any mechanical equipment there is a maximum tolerable size above which a fluidized bed reactor cannot be economically constructed using standard available materials. The physical size and operating stress combine to make this limit apparently a unit with a reactor calciner grid diameter of 16 feet. This is based on observations of fluidized bed equipment installed at the 23 locations from which data were obtained in performances of FWPCA contract 14-12-415. If a larger than 16 feet diameter reactor is required a series of reactors having equal diameters is selected by iteration in the program. The number of equal diameter reactors selected per system is ZII. The capacity of each reactor in pounds of dry solids per hour is then

$$ZI = PDS/ZII$$

The system cost curves are based on field data on equipment with feed capacities from approximately 4000 to 25,000 pounds of dry solids per hour (Figure 8).

CARD #1 X, Y, PCTY, FC, EPC
CARD #2 CLR, CLMR, CLMS, DC, PERW
CARD #3 CAP, QSL, YRS, CCAØ, W
CARD #4 DEWAL, PVS

						CARD #
.56	.75	.25	1.20	.06	CASE 1	1
2.	2.50	4.	5.	10.	CASE 1	2
.1	500.	20.	20.	.01	CASE 1	3
.01	10.				CASE 1	4

0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0
1 2 3 4 5 6 7 8 9 10 11 12 13 14 15 16 17 18 19 20 21 22 23 24 25 26 27 28 29 30 31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53 54 55 56 57 58 59 60 61 62 63 64 65 66 67 68 69 70 71 72 73 74 75 76 77 78 79 80
1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1
2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2
3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3

CASE 1

						CARD #
.56	.75	.25	1.20	.06	CASE 2	1
2.	2.50	4.	5.	10.	CASE 2	2
1.	500.	20.	20.	.01	CASE 2	3
.01	10.				CASE 2	4

0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0
1 2 3 4 5 6 7 8 9 10 11 12 13 14 15 16 17 18 19 20 21 22 23 24 25 26 27 28 29 30 31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53 54 55 56 57 58 59 60 61 62 63 64 65 66 67 68 69 70 71 72 73 74 75 76 77 78 79 80
1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1
2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2 2
3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3

CASE 2

Figure 12 SAMPLE INPUT DATA CARDS

CASE 1

SYSTEM IS TOO SMALL TO BE ACCURATELY SIZED BY THIS PROGRAM

CASE 2

COSTS DO NOT INCLUDE BUILDINGS OR LAND REQUIRED FOR EQUIPMENT

COSTS DO NOT INCLUDE LAND FOR STORAGE OF WASTES

0.5600	0.7500	0.2500	1.2000	0.0600	
2.0000	2.5000	4.0000	5.0000	10.0000	
1.0000	500.0000	20.0000	20.0000	4761.9052	
1.0000	5000001.0136	7690000.0117	5.1396	4761.9052	
7.4278	210.9704	38.5000	6.0000	355.7979	
349.6022	1571.4287	934.2852	542500.0014	4766547.0136	
12526.4804	12526.4804	34269.5391	1095.0002	684.3751	
58.4000	47891.0157	365.0000	146600.0004	177385.9691	
1107.7751					
69871.4377	9937.5898	58936.3750	0.0100	0.0100	10.0000

Figure 13 OUTPUT DATA GENERATED FROM INPUT DATA SHOWN ON FIGURE 12

Accuracy of the relationships developed is uncertain beyond these limits. It is estimated that reasonable approximations can be attained at higher capacities since the reactors are limited to 16 feet calciner grid diameter and systems are built in multiples of similar reactors.

Figure 14 shows a plot of the capital cost of the recalcination system in Feb. 68 dollars versus system capacity.

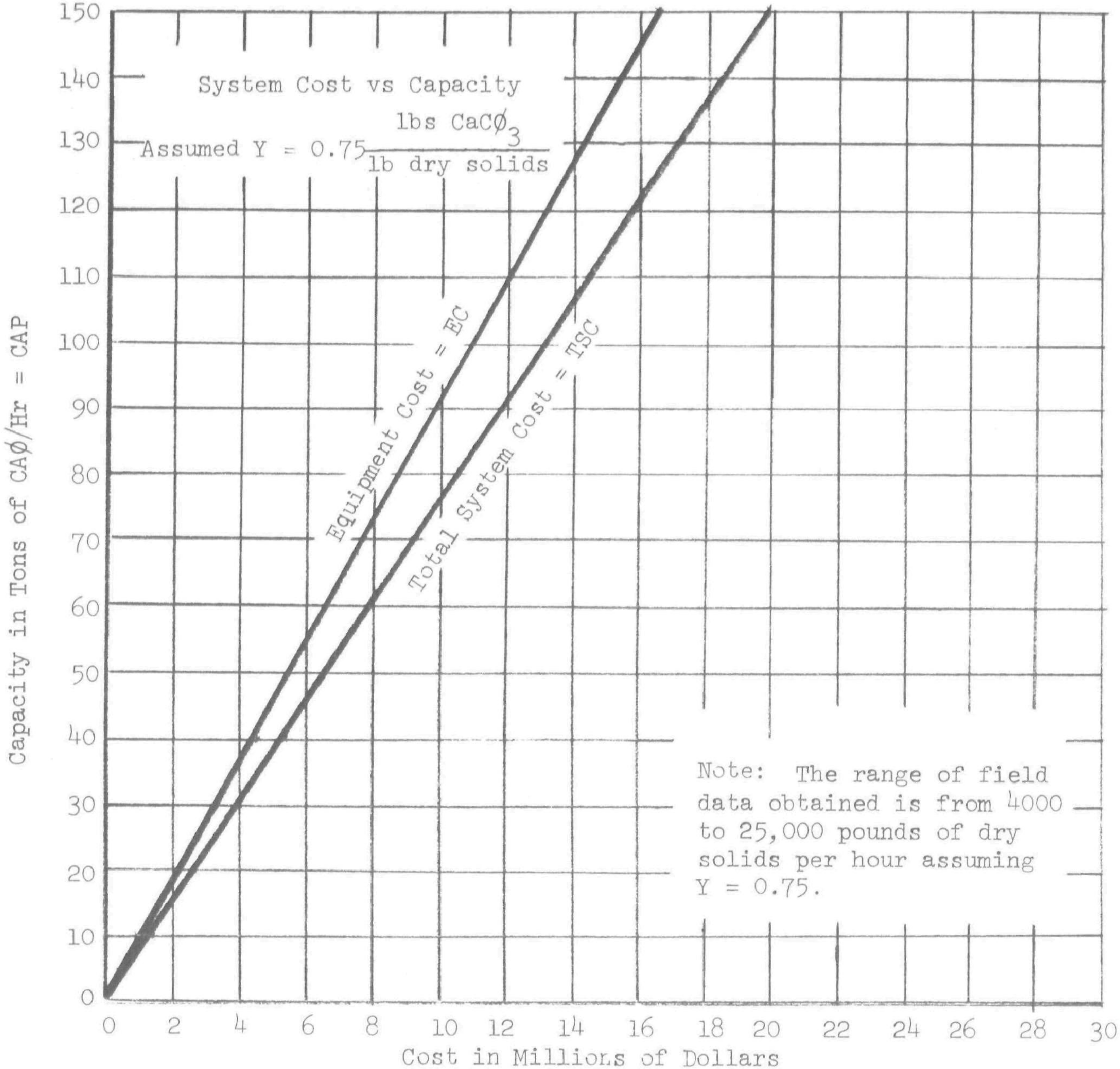

Figure 14 SYSTEM COST IN MILLIONS OF DOLLARS VS CAPACITY

CONCLUSIONS AND RECOMMENDATIONS

Performance of the System

In general, personnel at the installations surveyed were pleased with the performance of the fluidized bed recalcination system. The users of this equipment are mainly concerned with operating cost (\$/ton CaØ) as compared to the cost of makeup lime delivered to the storage bins. There are indications that with the current design practice fluidized bed recalciners can be operated at about 40% over design ratings.

No fluidized bed recalciners are presently being used for lime recovery at sewage treatment plants. It may be possible to recover lime by recalcination in the same reactor used to incinerate the primary sludge. If so a dual purpose incineration-recalcination process would greatly reduce the recalcination costs. Experience at Blue Plains with a multiple hearth recalciner indicates that only recalcination of relatively pure $CaCØ_3$ can be achieved. Quite possibly this will be true for the fluidized bed process. Then lime recovery from $CaCØ_3$ will be limited to the tertiary process. The computer program developed will adequately size and cost the fluidized bed system required for primary or tertiary lime recalcination provided a CaØ pelletized bed is used in the reactor and the equipment is similar to that upon which the model is based.

Since there were no fluidized bed recalciners in operation at sewage plants information on lime loss rates from such a system were not available. Based on the survey information the lime loss rate in the dewatering process is (DEWAL) 20%. The lime loss rate caused by required bed blowdown (PERW) to remove lumps of impurities from the bed is roughly estimated to be in the range of less than 1% for relatively pure $CaCO_3$ feed. It should be higher for impure $CaCO_3$ feed. Stack losses (W) are negligible because gas cyclone and scrubber efficiencies are very high (99+%).

Slaking difficulties were encountered at some plants because of slaking being attempted at too low a temperature.

Factors to be Considered

For a fluidized bed recalciner, the cost of operating and maintenance labor varies depending on the situation. In some cases permanent operators and maintenance personnel are assigned on the basis of peak work loads rather than normal work loads. In general, two man-hours per day should be sufficient to operate each recalciner. When problems occur, men must be available to correct them. Thus if a man is used to operate a system only part-time, he and other personnel must be available on short notice to tend to any malfunction and maintenance problems of the reactor system.

The recalcination system with the cost of land for the system and related disposal areas, buildings to house equipment, lime makeup, operation and maintenance must be weighed against a non-recalcining installation with greater lime makeup costs, larger storage areas for makeup lime and $CaCO_3$, and greater solids disposal costs.

The efficiency of combustion equipment will vary depending on how it is operated. The computer program developed assumes that near-optimum performance will be maintained. Excess air, bed conditions, pelletizing efficiency (i.e., soda ash concentration in bed), and use of waste heat in cooler or pub mill will play important parts in the operating efficiency. Optimum efficiency is seven to eight million Btu's per ton of recalcined CaØ.

In the installations surveyed, the capital costs of buildings, equipment, land and taxes were apparently neglected (or unimportant because of tax depreciation or other reasons). In general, the volume of the building required to house a fluidized bed recalcination system is roughly four times the volume of the reactor alone. A clearance of about ten feet is allowed both above and below the reactor.

Equipment Design

The system analyzed in this program includes a pellet cooler below the calcining zone. All fluidized bed recalcination systems should have such a cooler because:

1) the heat removal from the pellets is required so they can be handled and stored for reuse,
2) the heat removed from the pellets would otherwise be wasted if it were not used to preheat the fluidizing air and reduce fuel costs.

Care should be taken in design of the slaking process to insure that sufficiently high slaking temperatures are provided. This can be accomplished by using the heat from the hot exhaust gases leaving the reactor at about 1600°F. These gases are quenched with air or water to 1000°F before entering the cage mill. It would be wise to use the wasted heat for slaking or some other function such as fluidizing air preheat after the cooler section.

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TERMINOLOGY

AFC	=	Cross-sectional area of grid in calciner of each reactor, ft^2
AGC	=	Cross-sectional area of freeboard in calciner of each reactor, ft^2
AFØ	=	Cross-sectional area of freeboard zone above calcining section in each reactor, ft^2
ASHDC	=	Total system ash disposal cost, current $/hr
CAP	=	System calcining design capacity (all reactors combined), tons CaO/hr
CCAØ	=	Cost/ton lime delivered to location, current $/ton
CEF	=	Consulting engineers fee for total recalcining system, Feb. 1968 $
CER	=	Yearly cost of maintenance for total recalcining system, current $/yr
CFØ	=	Cost of fuel for operation of each reactor over a one-year period, current $
CFØI	=	Cost of fuel for operation of all reactors (ZII) in system handling PDS over one-year period, current $
CI	=	Installation cost for total system less building, Feb. 1968 $
CL	=	Cost of operating labor per year for total recalcining system, each reactor requires separate operator for 1 hr in each 4 hrs of operation, $/yr
CLIME	=	Total yearly purchased and disposal cost of new lime and other process solids, current $/yr without recalciner
CLIMR	=	Total yearly purchased lime and disposal cost of recalcination waste products for plant with recalciner, current $/yr
CLM	=	Yearly cost of maintenance labor for total system, each reactor requires 1 hr in 8 hrs of operation, current $/yr
CLMA	=	Yearly cost of maintenance service labor for total system, current $/yr
CLMR	=	Cost of maintenance labor rate, current $/man hr

CLMS = Cost of maintenance service labor rate, current $/man hr

CLMUW = Cost of lime makeup and solids disposal with recalciner, current $/yr

CM = Yearly cost of maintenance for total recalcining system, current $/yr

CØ = Yearly cost of operation of total system, current $/yr

CØ2 = Carbon dioxide liberated in reaction of CaCØ$_3$ CaØ + CØ$_2$ for total system capacity, pounds/hr

CP = Yearly cost of electrical power for total recalcining system, current $/yr

CLR = Cost of operating labor, current $/man hr

CSLIM = Cost of lime lost from system be dewatering, blowdown of bed and stack losses, current $/hr

DC = Disposal cost of solid wastes for removal from plant site, current $/ton

DEWAL = Percent of lime lost in dewatering process, 100 (#CaØ lost) per # CaO recovered

DFC = Freeboard diameter in calciner of each reactor, ft

DFØ = Diameter of Freeboard zone above calcining section in each reactor, ft

EC = Total system capital equipment cost for equipment from and including dewatering device to stack inlet not including land and buildings required, Feb. 1968 $

ENDØ = Endothermic heat of reaction (CaCØ CaØ + CØ$_3$), (total for all reactors in system), Btu/hr

EPC = Electrical power cost, current $/KW hr

FC = Fuel cost per million Btu's, current $

HC = Height of cooler section in each reactor, ft

HE = Depth of expanded fluidized bed in each reactor, ft

HEXP = Height of expansion section between calcining and freeboard zone in each reactor, ft

HF = Height of freeboard above calcining section in each reactor, ft

HP	=	Height of air distribution plenum in each reactor, ft
HPI	=	Installed HP in total system HP
HPIZI	=	Installed HP per reactor, HP
HS	=	Static depth of fluidized bed in each reactor, ft
NFP	=	Number of points at which CaØ sludge is fed into the fluidized bed in each reactor, dimensionless
ØHP	=	Operating HP in total system, HP
ØHPZI	=	Operating HP per reactor, HP
ØRH	=	Overall height of each reactor, ft
PAPH	=	Pounds ash per hour, as calcium phosphate and other wastes from total system, pounds/hr
PCTY	=	Fraction of year, week of month to be operated, hrs operated/hrs in year (total hours, not work hours); i.e., continuous operation for one year is for 8760 hrs/yr. For 1/2 time, PCTY = .5 and operating time = 4380 hrs/yr.
PDS	=	Total design pounds of dry solids per hour fed to all reactors in system, pounds/hr
PERW	=	Percent of lime lost in blowdown for periodic removal of inerts in all recalciners, 100(#CaØ lost)/#CaØ recovered.
PVS	=	Percent volatile solids in sludge, 100(pound VS)/pounds dry solids
QAPR	=	Total heat supplied to calcining process through air preheating in the cooler/coolers section(s) of all reactors combined, Btu/hr
QCAØ	=	Gross heat minus endothermic heat of CaØ required per ton of CaO produced, Btu/ton CaØ
QFØ	=	Heat required from fuel in each reactor of system, Btu/hr
QSL	=	Sludge heat input (lower heating value), Btu/pound of dry solids
SLIM	=	Lime lost from system by dewatering, blowdown of bed and stack losses, tons/hr
SUMQ	=	Gross heat input to fluidized bed for all reactors combined in system considered, Btu/hr

SUMQI = Design net heat input to each recalciner in system, Btu/hr

SUMQN = Net heat input to adiabatic fluidized bed (total for all reactors in system), gross heat minus endothermic heat of reaction, Btu/hr

TCLIW = Total yearly cost of operation, maintenance, lime makeup and solids disposal with recalciner, from inlet to centrifuge or vacuum filter, current \$/yr

TCMØ = Total yearly system operating and maintenance cost, current \$/yr

TSC = Total system cost including installation and consultants fees less building and land, Feb. 1968 \$

VE = Expanded volume of fluidized bed in each reactor, ft^3

VS = Static volume of each fluidized bed, ft^3

W = Pounds of CaØ lost through stack per pound CaØ recovered

X = Pounds CaØ per pound $CaCØ_3$ in feed to reactors = 0.56

Y = Pounds $CaCØ_3$ per pound dry solids in feed to reactors

YRS = Estimated operating life of recalciner, yrs

ZI = Design pounds of dry solids per hour each recalciner in the system handles, pounds/hr

ZII = Number of recalciners in system required to handle the design sludge capacity

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