FLUIDIZED BED PYROLYSIS OF CATTLE FEEDLOT MANURE

by

CADY ROY ENGLER

B. S., Kansas State University, 1969

A MASTER'S THESIS

submitted in partial fulfillment of the

requirements for the degree

MASTER OF SCIENCE

Department of Chemical Engineering

KANSAS STATE UNIVERSITY
Manhattan, Kansas

1974

Approved by:

Walter P Walawender

Major Professor

CONTENTS

Chapter		Page

CHAPTER I

INTRODUCTION

In recent years citizens of the United States have suddenly become aware of problems regarding preservation of the environment and conservation of our natural resources. These considerations had previously been the concern of only a small group of conservationists. The once limitless supply of natural resources has been found to have definite limits as witnessed by the energy crisis in late 1973. While air and water pollution have received numerous headlines, the equally serious problems of solid waste disposal has been more quietly passed on to city governments who find they are rapidly running out of places to bury the waste. Fortunately, people have now begun to realize the importance of solid waste as a raw material resource and the significance of resource recovery in reducing pollution problems. Most of the research in the area of solid waste utilization has been directed toward the most visible type of waste, municipal solid waste (MSW), because of two factors: the demand for environmentally acceptable methods for treating solid wastes and the availability of MSW in large quantities in central locations.

While the magnitude of the municipal solid waste problem has indeed become large with over 200 million tons (dry, ash-free basis) produced annually, agricultural wastes are produced in even greater quantities with the generation rate estimated at 600 million tons per year as reported by Wender, et al. (1). The amounts of various types of solid wastes that are expected to be generated in the U.S. during 1974 are listed in Table I. The amounts expected to be collected are also included. Although most of the agricultural wastes do not present a serious pollution problem, the development of large modern feedlots has resulted in accumulation of large quantities of manure in relatively small areas and created significant environmental problems. Feedlot manure generation has been estimated in a variety of ways but experience in major feeding areas has indicated that about two tons per year of semi-composted manure with a 50% moisture content can be expected to accumulate for each head of feedlot capacity (2). For a 50,000 head feedlot this would amount to manure production of 100,000 tons per year.

From our survey of feedlots in southwestern Kansas, there are areas where feedlot capacities exceed 500,000 head within a 50-mile radius. In other parts of the country there are even more dense feedlot concentrations; for example, over 600,000 head capacity is available within a 15-mile radius in the Hereford-Dimmitt area of the Texas high plains. Disposing of the manure generated in such intensified feeding areas by the conventional means of applying to the land as fertilizer has become quite expensive. In some locations the cost of hauling and spreading manure has become so great there is no demand and the manure has had to be stockpiled. Providing an environmentally acceptable means for disposing of large quantities of manure by turning it into a material or energy resource would be an asset not only for the feed-lot operator but also for the local community.

Previous studies at Kansas State University (3,4,5) examined the feasibility of three processes for converting manure into a material and energy resource. Although the studies concentrated on feedlot manure, the technology could be readily adapted to treatment of MSW. All the processes had been demonstrated to be technically feasible by other researchers. The processes were liquefaction to produce an oil-like material based on research at the U.S. Bureau of Mines (6), hydrogasification to produce a natural gas substitute, also based on Bureau of Mines research (7), and pyrolysis to produce synthesis gas composed primarily of carbon monoxide and hydrogen using technology developed at West Virginia University (8). While none of the processes were economically feasible, the pyrolysis process appeared to have the greatest chance of becoming feasible in the near future.

Pyrolysis can be defined as the break-down of complex organic compounds into simpler ones using heat in the absence of oxygen. Products obtained from pyrolysis of a complex substance such as manure include gases, liquids and solids, and the quantities of each depend on the conditions under which pyrolysis occurs. The major variables affecting product distribution are temperature and length of time of the reaction and the heating rate. High temperatures for pyrolysis favor the production of gaseous products while lowering the yield of liquid products. Maintaining the products at the pyrolysis temperature for longer periods of time results in a similar effect through cracking of low molecular weight hydrocarbons. Solids production is most significantly affected by the heating rate. As indicated in the West Virginia University study (8), very rapid heating of the complex organic molecule causes it to be torn into fragments before a stable solid matrix can form.

One of the objectives of the present work is to maximize the production of gaseous products from the pyrolysis of feedlot manure. To achieve this the reactor should operate at the highest practical temperature, subject the feed to a very rapid heating rate, and be large enough to provide a moderate residence time for the vaporized products. Different types of reactors have been developed which achieve some or all of the conditions for maximizing gas production from solid waste pyrolysis. The TTU retort (9) was developed at Texas Tech University to study manure pyrolysis. It is a moving bed reactor containing both combustion and pyrolysis zones. Other reactors include the U.S. Bureau of Mines retort (10) which is a modified coking reactor and the Koppers coal gasifier which uses oxygen and steam to gasify solid fuels. An entrained flow reactor has been developed by Garrett Research and Development Company (11) and a fluidized bed pyrolysis reactor used by West Virginia University researchers (8) to treat MSW. The optimum combination of reaction conditions, however, is most readily achieved in a fluidized bed reactor. A fluidized bed consists of a vessel that is partially filled with particulate solids supported on a porous plate or similar device that also acts as a gas distributor. Gas is introduced into the bottom of the reactor and flows upward through the solid bed. When the flow of gas through the solids becomes large enough so that the pressure drop across the bed equals the pressure exerted by the bed on the gas distributor, the solids either start to move out of the vessel as a slug or the particles rearrange themselves so there is more space between them. This is called the point of incipient fluidization illustrated in Figure 1.

Increasing the gas flow beyond the point of incipient fluidization causes the solid particles to move farther apart and allows less restrictive movement of the particles. The bed is then in the fluidized state under which conditions it has many of the characteristics of a boiling liquid:

1. There is an apparent viscosity for the bed, and the bed obeys Archimedes' principle.
2. The upper surface of the bed remains horizontal when the vessel is tipped.
3. A large light object will float while a heavy one will sink.
4. Fluidized solids will flow out of a hole placed in the side of the vessel below the upper surface of the bed.
5. With a large enough gas rate, gas bubbles will form in the bed.

The above analogy to a boiling liquid is shown in Figure 2.

When the gas rate is high enough to produce bubbles, the solids in the bed become agitated and can move from one location to another usually establishing a definite circulating pattern. Depending on the vessel geometry, gas flow rate and solid particle size, the bubbles may coalesce and become large enough to spread across the entire vessel diameter. The bed is then said to be slugging with the solids above the gas bubble pushed upward as by a piston. Particles rain down through the bubble until it finally disintegrates. In most cases slugging is undesirable; however, it does provide violent agitation of the solids bed and keeps it well mixed. Further

THIS BOOK CONTAINS NUMEROUS PAGES THAT HAVE INK SPLOTCHES IN THE MIDDLE OF THE TEXT. THIS IS AS RECEIVED FROM CUSTOMER.

THESE ARE THE BEST IMAGES AVAILABLE.

discussion of the properties of fluidized beds is available in various texts (12,13,14,15).

During pyrolysis, the manure particles shrink to form an ash and char solid byproduct. Because of the wide particle size distribution for coarsely ground manure, a mixture of manure and the solid pyrolysis byproducts is very inhomogeneous. A consistent quality of fluidization can not be maintained with such a mixture. In addition, the low densities of the solids limit the superficial gas velocity used. These problems can be overcome by using a more dense, inert material such as sand to make up the bulk of solids in the bed.

A fluidized bed composed of an inert material offers several advantages for pyrolysis of solids. At moderately high gas velocities the solids are kept completely mixed so that isothermal operation is possible. Thus the solids being pyrolyzed could be introduced into an isothermal bed of inert material maintained at a high temperature. The rapid mixing and isothermal conditions create a very high heat transfer rate which favors gas production. Because of the rapid reaction rate (less than one second for particles less than 0.1 cm diameter), the pyrolysis feed is quickly consumed. That, coupled with a large inventory of inert solids in the bed, enables high capacities for fluidized bed pyrolysis reactors.

Since pyrolysis is an endothermic process, provisions must be made for the addition of heat to the reactor. Figure 3 shows three ways this can be accomplished. One is to introduce a limited amount of oxygen with the fluidizing gas so that partial combustion occurs in the fluidized bed reactor (Figure 3a). Another is a series fluidized bed system (Figure 3b) in which feed is pyrolyzed in one reactor and the solid residues are then fed to a second reactor where they are burned. Hot combustion gases provide heat for pyrolysis and maintain fluidization in the pyrolysis reactor. Finally, solids acting as a heat transfer medium can be circulated between pyrolysis and combustion reactors as shown in Figure 3c. The first two methods have the disadvantage of diluting the pyrolysis gas with combustion gases. The third method eliminates gas dilution problems but adds the problem of solids circulation.

The purpose of this study is twofold: to examine the economic feasibility and potential for processing feedlot wastes via fluidized bed pyrolysis and to provide partial design information for a pilot scale gasifier system. Chapter II presents the conceptual design of a pyrolysis plant for processing 500 tons per day of dry manure. An economic analysis of the process is given along with sensitivity analysis of several variables that could affect gas production costs.

Using the conceptual design as a basis, the potential for large-scale processing of feedlot manure in southwestern Kansas is studied in Chapter III. Feedlot capacities for various regions are given as well as the sizes of pyrolysis plants that could be supported. Utilization of the resulting synthesis gas for ammonia production and electricity generation is also considered.

The results of experimental studies using air to fluidize mixtures of sand with manure and ash in a simulation unit are presented in Chapter IV. The objectives of the studies were to observe bed behavior under conditions approximating those estimated for the manure pyrolysis process and to provide design data concerning elutriation rates for fine sand, manure and ash. Results of the elutriation experiments were compared to existing correlations and used to estimate the amount of ash build-up in the solids for a pilot scale gasifier.

REFERENCES

1. Wender, I., F.W. Steffgen and P.M. Yavorsky, "Clean Liquid and Gaseous Fuels from Organic Solid Wastes," in Recycling and Disposal of Solid Wastes, T.F. Yen, ed., Ann Arbor Science Publishers, Ann Arbor, Michigan, 1974, p. 47.

2. Reddell, D.L., W.H. Johnson, P.J. Lyerly and P. Hobgood, "Disposal of Beef Manure by Deep Plowing," Livestock Waste Management and Pollution Abatement, American Society of Agricultural Engineers, St. Joseph, Michigan, 1971, p. 235.

3. Walawender, W.P., L.T. Fan and L.E. Erickson, "Feedlot Manure and Other Agricultural Wastes as Future Material and Energy Resources: I. Introduction and Literature Review," Report No. 36 of the Institute for Systems Design and Optimization, Kansas State University, Manhattan, Kansas, April 1972.

4. Walawender, W.P., L.T. Fan, C.R. Engler, and L.E. Erickson, "Feedlot Manure and Other Agricultural Wastes as Future Material and Energy Resources: II. Process Descriptions," Report No. 45 of the Institute for Systems Design and Optimization, Kansas State University, Manhattan, Kansas, March 1973.

5. Walawender, W.P., L.T. Fan, C.R. Engler, and L.E. Erickson, "Feedlot Manure and Other Agricultural Wastes as Future Material and Energy Resources: III. Economic Evaluations," Report No. 46 of the Institute for Systems Design and Optimization, Kansas State University, Manhattan, Kansas, July 1973.

6. Appell, H.R., S. Friedman, Y.C. Fu, P.M. Yavorsky and I. Wender, "Converting Organic Wastes to Oil," Agricultural Engineering, 53, 3, 1972, p. 17ff.

7. Feldman, H.F., "Pipeline Gas from Solid Wastes," Chemical Engineering Progress, 67, 12, 1971, p. 51ff.

8. "Solid Waste: A New Natural Resource," Staff of the Fluidized Bed Gasification Project, Department of Chemical Engineering, West Virginia University, Morgantown, W. Va., 1971.

9. Massie, J.R., Jr. and H.W. Parker, "Continuous Solid Waste Retort-Feasibility Study," paper presented at the 74th National Meeting, A.I.Ch.E., New Orleans, March 1973.

10. Crentz, W.L., "Agricultural Wastes - An Energy Resource of the Seventies," presented at the World Farm Foundation Symposium, Anaheim, California, December 1971.

11. McMath, H.G., R.E. Lumpkin and A. Sass, "Production of Gas from Western Sub-bituminous Coals by the Garrett Flash Pyrolysis Process," paper presented at the 66th Annual Meeting, A.I.Ch.E., Philadelphia, November 1973.

12. Kunii, D. and O. Levenspiel, Fluidization Engineering, John Wiley and Sons, New York, 1969.

13. Fluidization, J.F. Davidson and H. Harrison, editors, Academic Press, New York, 1971.

14. Zenz, F.A. and D.F. Othmer, Fluidization and Fluid-Particles Systems, Reinhold, New York, 1960.

15. Leva, M., Fluidization, McGraw-Hill, New York, 1959.

Table I

Solid Waste Generation and Collection in the U.S.[a]

Type of Waste	Wastes Generated[b] 10^6 tons/yr.	Wastes Collected[b] 10^6 tons/yr.
Agricultural and food	412	25
Manure	210	30
Urban	160	80
Logging and manufacturing	70	7
Industrial	50	7
Municipal sewage solids	18	2
Miscellaneous	80	10
Total	1000	161

[a]Wender, et al. (1)
[b]Moisture-and ash-free organic solids

THIS BOOK CONTAINS NUMEROUS PAGES WITH DIAGRAMS THAT ARE CROOKED COMPARED TO THE REST OF THE INFORMATION ON THE PAGE.

THIS IS AS RECEIVED FROM CUSTOMER.

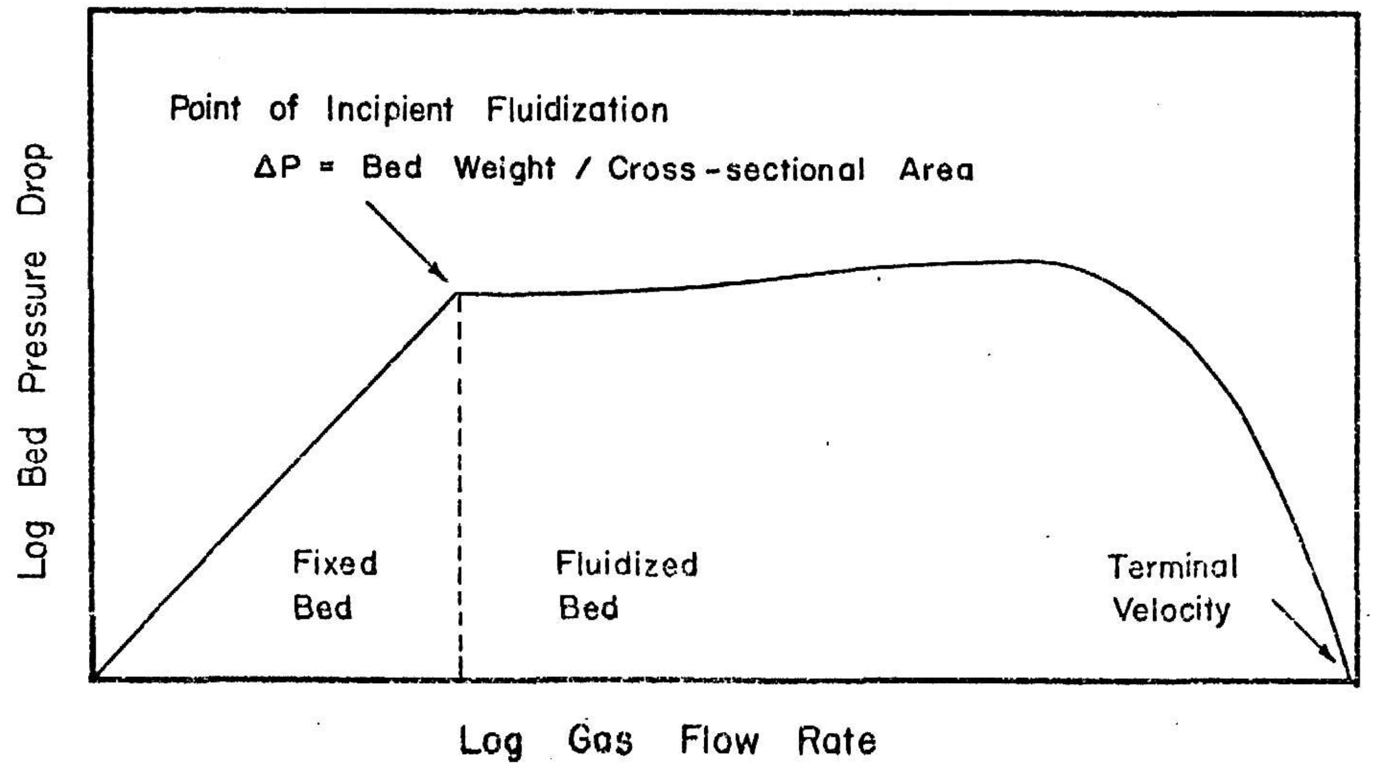


Fig. I. Pressure Drop Diagram for a Bed of Solids.

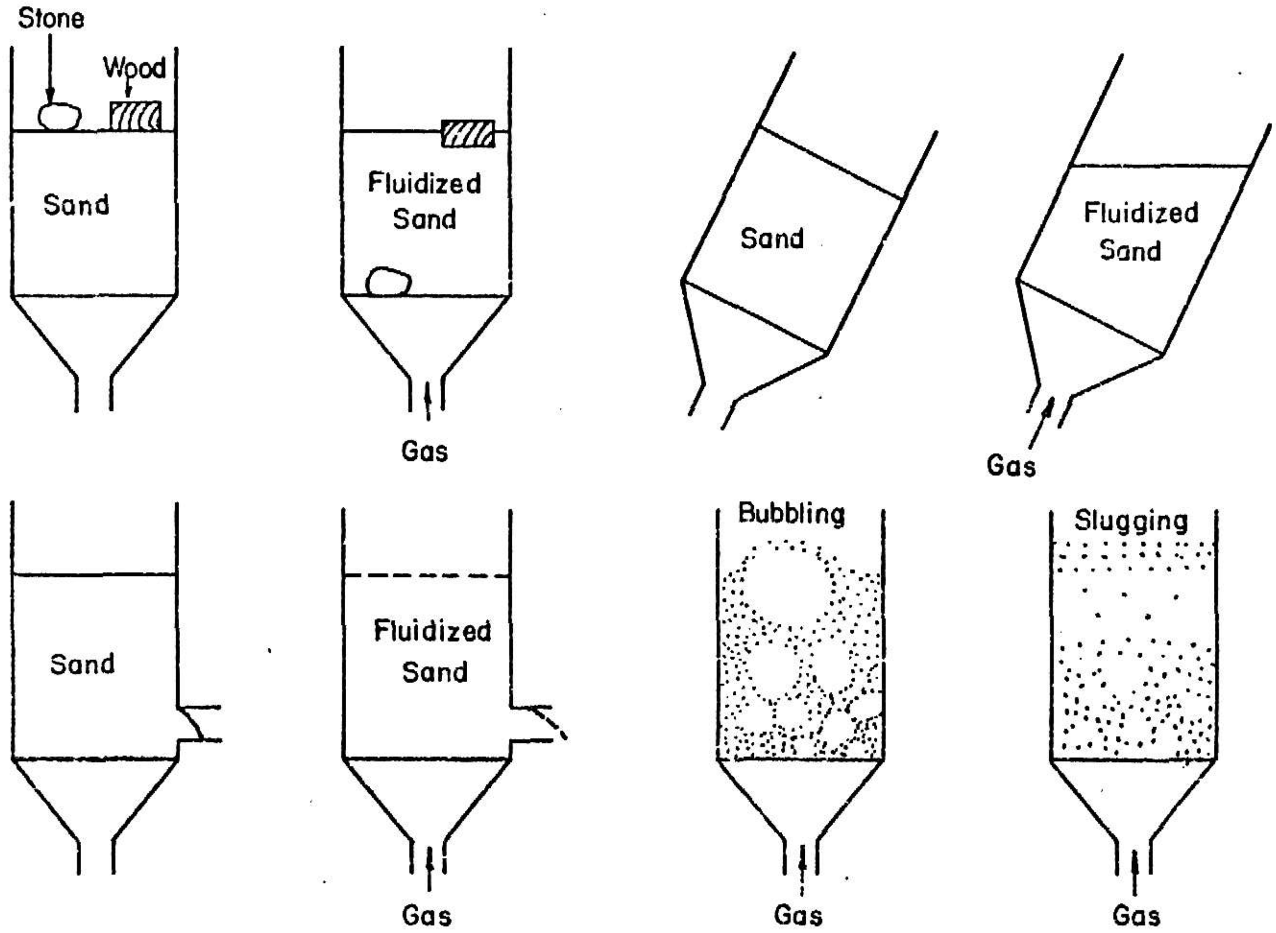


Fig. 2. Liquid-like Behavior of Gas Fluidized Beds.

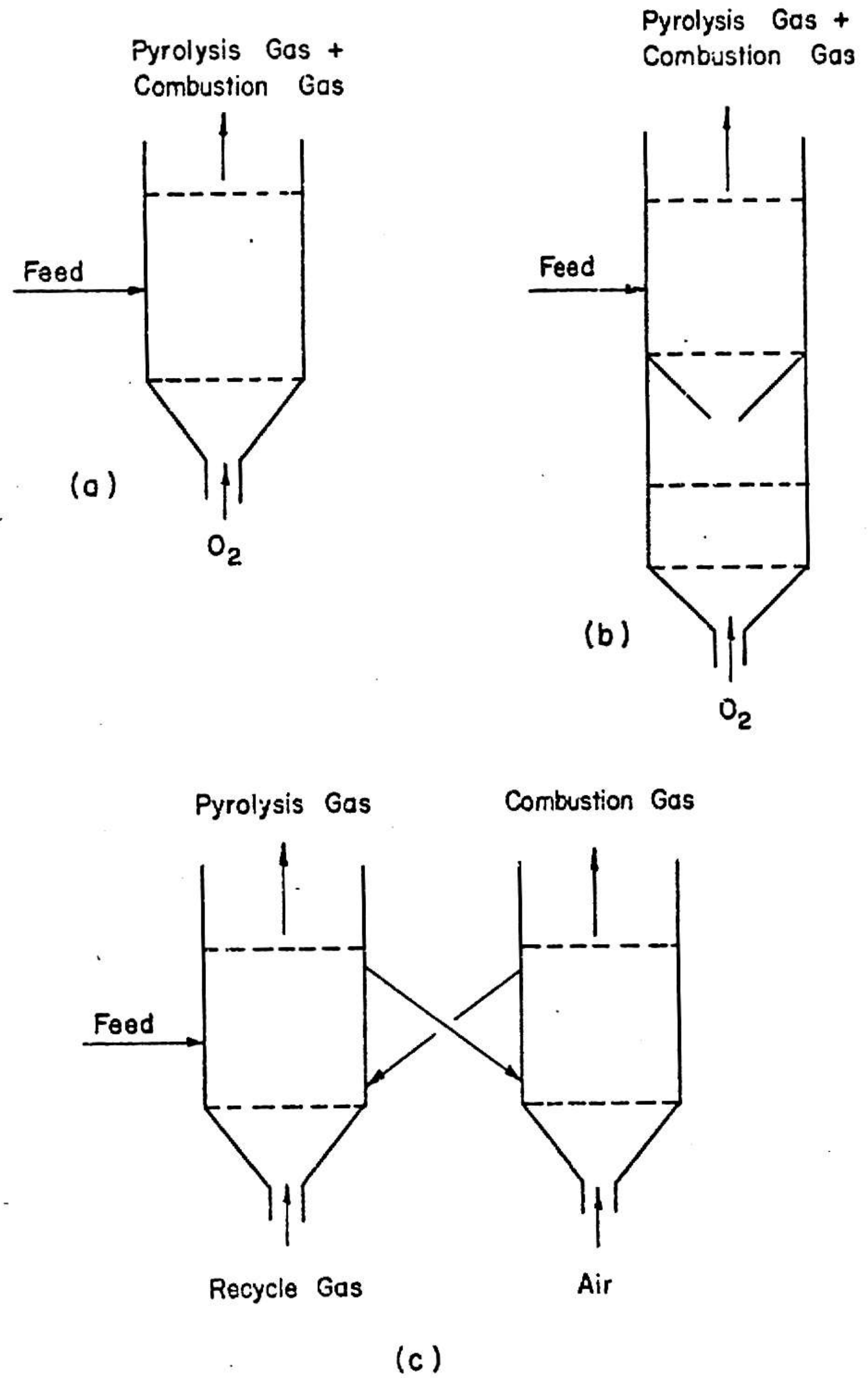


Fig. 3. Methods of Supplying Heat of Pyrolysis.

CHAPTER II

CONCEPTUAL DESIGN STUDY

INTRODUCTION

The cattle-feeding industry is important to the economy of Kansas and other Great Plains states. Recent trends in this industry are a rapidly increasing number of cattle on feed and significant increases in the capacities of individual feedlots. In addition, the larger feedlots tend to be grouped within relatively small areas. The large animal to land ratios thus created have led to serious manure disposal problems and environmental hazards.

On a weight basis manure generated by beef cattle is the major product of the industry. Approximately two tons per head per year of semicomposted manure containing 50% moisture are produced. The old practice of using manure as a fertilizer is not feasible on such a large scale so that often the manure is simply discarded at considerable expense to the operator. Such disposal creates a high potential for causing environmental problems. Converting such a liability into an asset would insure lower beef production costs while preventing damage to the environment.

One approach to manure utilization is through chemical processing to provide material and energy resources. Preliminary studies of three processes (1,2) indicate that a pyrolysis process to generate synthesis gas is most nearly compatible with the current economic situation. The resultant synthesis gas product could be used to provide a clean low-Btu gas for power generation, a starting material for ammonia synthesis or a starting material for methanol production. In addition to the synthesis gas, ash produced as a byproduct could be used as a nitrogen-free fertilizer after suitable processing.

In this chapter a conceptual plant design is developed for a plant processing 500 T/D (tons per day) of dry manure to yield a synthesis gas composed of CO, H_2 and CH_4. Estimated capital investment and operating costs for the plant are presented along with the results of sensitivity analysis for the effects of variations in the size of the plant, raw manure moisture content, manure composition, and transportation costs on the cost of producing the synthesis gas.

DESIGN CONSIDERATIONS

The chemical composition of manure, its moisture content, and quantity per-head vary widely with rations fed, climatic conditions and age of the cattle. Chemical compositions of feedlot manure have been reported by several investigators (3,4,5,6,7,8). To provide a basis for the conceptual plant design, the composition reported by Appell et al. (3) was used. Table I shows the composition of raw manure as well as the chemical composition of the moisture-and-ash-free portion. For comparison, a typical composition of municipal solid waste as reported by Appell and Wender (9) is also shown.

A moisture content of 50% was used for the design which corresponds to a semicomposted manure in a non-arid climate. Since the moisture content is subject to variation and none of the pyrolysis data were taken with feed moisture content above 10%, drying the raw manure to 10% moisture provided a conservative design approach and provided a uniform feed for the pyrolysis reaction.

As mentioned earlier, the quantity of manure available averages about 2 tons per head per year (50% moisture) or approximately 5.5 pounds per day on a dry basis. While there are feedlot capacities up to 750,000 head within a 50-mile radius, a more realistic concentration is about 200,000 head in such an area. That concentration was used to set the conceptual plant design capacity at 500 T/D.

A hypothetical stoichiometry for the pyrolysis reaction was developed from municipal solid waste gasification data reported in a study conducted by West Virginia University (10). Although data were available from manure gasification studies (4,5,8), data from West Virginia University were taken at process conditions most nearly approaching those visualized for the conceptual plant. As the composition of municipal solid waste shown in Table I is quite similar to that of manure, the gasification data should be applicable. In developing the stoichiometry, it was assumed that the ratios of the major constituents in the product gas would remain constant (i.e. mole ratio $CO_2 : H_2 : CO : CH_4 = 1 : 2.28 : 2.15 : 0.68$). In addition, it was assumed that all the oxygen in the manure would be converted to either CO or CO_2. The carbon and hydrogen left after balancing the gas ratios were assumed to form the char product. Because the 10% moisture content of the feed was greater than that used by the West Virginia University group, it was further assumed that 25% of the char would react with steam to form CO and H_2. The resulting hypothetical stoichiometric balances follow:

Manure: $C_{4.35}H_{6.36}O_{2.31}N_{0.24}S_{0.04} \rightarrow 1.20\ CO + 0.55\ CO_2 + 0.38\ CH_4 + 1.26\ H_2 + 0.24\ NH_3 + 0.04\ H_2S + C_{2.22}H_{1.52}$(char)

Char: $0.25\ C_{2.22}H_{1.52} + 0.56\ H_2O \rightarrow 0.56\ CO + 0.75\ H_2$

Overall: $C_{4.35}H_{6.36}O_{2.31}N_{0.24}S_{0.04} + 0.56\ H_2O \rightarrow 1.76\ CO + 0.38\ CH_4 + 2.01\ H_2 + 0.55\ CO_2 + 0.24\ NH_3 + 0.04\ H_2S + 0.75\ C_{2.22}H_{1.52}$

Further details of the hypothetical stoichiometry calculations are given in Appendix A. The gases produced in a plant processing 500 T/D of dry manure are given in Table II.

Clean synthesis gas production amounts to 10.7 million standard cubic feet per day including only CO, H_2 and CH_4.

The above stoichiometry was used to determine the heat of reaction at the pyrolysis conditions. Standard heats of combustion for manure and char were calculated using the Dulong formula and are given in Table III along with other pertinent thermochemical information. The heat of the overall reaction at 1500°F (pyrolysis reactor temperature) was calculated to be 828 Btu/lb. DAFM (dry, ash-free manure) as shown in Appendix A. The heat for the pyrolysis reaction can be supplied by combustion of the char product. Details of the heat balance are also given in Appendix A.

It has been shown that the synthesis gas yield increases as the rate of heating of the solid feed increases (4,10). One way to achieve rapid heating of the feed is to use a fluidized bed of inert material such as sand for the pyrolysis reaction. With such a bed, near isothermal conditions can be maintained. Heat can be supplied to the pyrolysis reactor either by partial combustion within the reaction vessel or by circulating the sand to a separate combustion vessel as was used in the conceptual design.

Although a circulating fluidized system is more complicated than having a single reaction vessel, separating the combustion and pyrolysis reactions has several advantages. The primary advantage is that air can be used to supply oxygen for combustion without the nitrogen diluting the synthesis gas product. In addition, separate reaction vessels would be inherently safer because there would be less possibility of developing an explosive mixture of gases during any system failures.

Studies by Wen, et al. (11) indicate that the synthesis gas yield improves as the residence time of the volatile products at pyrolysis reaction temperature is increased. That reduces the amount of liquids produced by allowing them to crack to lighter compounds. Gas residence time can be increased by increasing the freeboard height above the fluidized bed. For the design a freeboard of 3 to 4 times the expanded bed height was used.

It was assumed that the ash and char particles formed during pyrolysis would not agglomerate and would be much smaller than the circulating sand particles. Based on those assumptions, the ash and char would be elutriated by the pyrolysis gas after an equilibrium concentration of fines was reached in the fluidized bed. Solids separated from the pyrolysis gas would be fed to the combustion reactor where the char would be burned completely. Ash leaving the reaction system with the combustion gas would be separated for nitrogen-free fertilizer recovery.

The pyrolysis gas would need to be cooled and compressed to a specified delivery pressure prior to use in another process. In addition, removal of CO_2 and other undesirable products may be necessary. Several absorption processes such as hot carbonate, aqueous ammonia, alkanolamine, and water are available for removing CO_2. Since water scrubbing required higher pressure than other methods, it was selected as a conservative design approach due to the high compression and pumping requirements.

CONCEPTUAL DESIGN

The process flow diagram for a plant to produce synthesis gas from 500 T/D of dry manure is shown in Figure 1. Major sections of the plant include feed preparation and drying, reaction, and gas clean-up. The plant produces 10.7 million SCF/day of synthesis gas having a heating value of about 400 Btu/SCF (about 40% of the heating value of natural gas) and 139 T/D of ash. A detailed equipment list is given in Table IV.

Manure is transported from the feedlot to the plant by truck. The receiving facility is a storage building large enough to store three days' supply of raw manure and provide room for unloading and transfer to the dryer feed conveyor. The storage area is suitably enclosed to meet esthetic and environmental requirements.

A front end loader (PL-101) transfers the raw manure to the feed conveyor (CV-101), which in turn, takes it to the dryers (DR-101 A & B). Since particle size reduction is required as well as drying, a cage mill/flash drying system was selected to perform both operations in one step. In the dryer the moisture content of the solids is reduced to about 10 wt.% and the particle size reduced to 60 mesh. Wet gas leaves the dryer at about 400°F and is flared to control odor.

Dry solids are stored in hoppers (V-101 A & B) which contain approximately 1/2-day's feed supply. The dry manure is metered continuously through rotary feeder valves into a high velocity stream of recycle pyrolysis gas which conveys the feed into the bottom of the fluidized bed reactor (R-101).

The pyrolysis reactor (R-101) consists of a steel shell 12 feet in diameter and 35 feet high lined with 18 inches of firebrick. The reactor is equipped with an alloy grid to support the 6-foot bed of fluidized sand and operates at 1500°F and 50 psig. The combustion reactor (R-102) is of similar construction with a shell diameter of 16 feet and height of 44 feet. It operates at 1750°F and 50 psig with a fluidized bed height of 10 feet. Design calculations for the reactors are given in Appendix B. Flow of solids between the two reactors is maintained and controlled by injection of gas into vertical sections of the circulating lines to create specific gravity differentials.

Introduction of the manure into the pyrolysis reactor causes very rapid formation of gas and solid particles. The superficial gas velocity in R-101 is about 3.25 feet/second. The gas, together with the entrained ash and char, passes through a two-stage multicyclone installation (S-101) for recovery of the solids which are sent to the combustion reactor. Part of the gas is compressed to 60 psig to be recycled to the pyrolysis reactor to maintain fluidization; the remainder is used to preheat combustion air in E-102 and then sent at 600°F to the quench column (T-101).

In the combustion reactor (R-102), char formed in the pyrolysis reaction is burned to provide heat for the pyrolysis. Because more heat is available from the char than is needed for pyrolysis, excess air is used to maintain the combustion temperature at 1750°F. The hot combustion gases and air are then used to dry the raw manure. The superficial gas velocity in the combustion reactor (R-102) is also about 3.25 feet/second.

Ash is removed from the combustion reactor (R-102) by entrainment in the gas and then separated from the gas in a louvered collector followed by a multicyclone (S-102). The ash is cooled in hollow-flight cooling conveyors (E-101 A-C) and sent to collection hoppers (V-102 A & B). The hoppers hold approximately two days' production of ash which is removed from the plant by truck.

The pyrolysis product gas flows to a quench column (T-101) where it is cooled to 150°F by countercurrent contact with recirculating water. The quench column is 5 feet in diameter, 15 feet high and packed with 10 feet of 2-inch Raschig rings. The cool gas is compressed to 250 psig in the compressor (C-103) and fed to the CO_2 absorption column (T-102). Carbon dioxide is removed from the synthesis gas by high-pressure countercurrent contact with water. The absorption column is 11 feet in diameter, 28 feet high and packed with 20 feet of 2-inch Raschig rings. The synthesis gas product is available from the CO_2 absorption column (T-102) at about 150°F and 240 psig.

Carbon-dioxide-saturated water is regenerated in the regenerator (T-103) which is 12 feet in diameter and 25 feet high. The upper part of the column is packed with 10 feet of 2-inch Raschig rings, and the tower is constructed so that air can circulate upward through the packing. Water from the quench tower is also circulated through the regeneration column for cooling and removal of absorbed materials.

ECONOMIC ANALYSIS

To evaluate the feasibility of manure pyrolysis, investment and operating costs were estimated using the preceeding conceptual design. Using these estimated costs as a base, variations were made in the size of the plant, raw manure moisture content, manure composition and transportation costs to determine their effects on the profitability of the operation. That is, a sensitivity analysis of the plant was carried out.

Capital Investment

Capital investment was estimated by the method described by Guthrie (12). For the processing equipment bare module costs, data from the article were used whenever applicable. Installed equipment cost for the dryers was taken from an article by Grzelak (13). If indirect factors were not given for a particular piece of equipment, the indirect cost was assumed to be 38.5% of installed equipment cost. Details of the cost estimation are given in Appendix B.

The bare module cost for buildings includes a compressor house, control house, administrative offices, shop area and structure for the equipment excluding the dryers. For all buildings the lowest cost categories were used.

The cost of site development was estimated at 10% of equipment f. o. b. as suggested by Peters and Timmerhaus (14). Offsite facilities include water, instrument air and fuel-gas systems, a flare, fire protection equipment, power distribution and yard lighting.

All bare module costs were based on 1968 information escalated to mid-1974 costs by a factor of 1.355 as estimated from the Marshall and Stevens Index. A summary of capital investment costs for the processing equipment is given in Table V. The fixed capital investment cost of $5.65 million (MM) was estimated by applying factors for contingency and contractors fees as shown in Table VI. Working capital was estimated at 15% of fixed capital investment to give the total capital investment required of $6.50 MM.

Annual Operating Costs

The annual operating costs were estimated at $2.94 MM using information from Peters and Timmerhaus (15). A summary of these costs is presented in Table VII. The plant was assumed to operate 350 days per year in making the estimates.

Since it was difficult to assign a raw material cost to manure, it was assumed there would be no cost for the manure at the feedlot. However, the cost of transporting the manure to the plant was taken as a raw material cost. In effect this represents an indirect payment to the feedlot operator because it eliminates the disposal costs which he may incur. Transportation costs were estimated to average $1.50/ton assuming $0.06/ton-mile for an average distance of 25 miles from feedlots within a 50-mile radius. The cost of $0.06/ton-mile represents the low end of high-volume short-haul trucking rates.

Utilities were estimated on the basis of using electric motors for all plant drivers. The cost of electricity was taken as $0.011/kilowatt-hour. The only other utility included was make-up water for the scrubbing system at $0.11/thousand gallons.

Operating labor costs were calculated assuming four men required per shift at an hourly wage of $4.90. A factor of 4.2 was used to allow for continuous operation of the plant. Supervisory labor was taken as 10% of operating labor.

Maintenance and repairs were calculated as 7.5% of the fixed capital investment to include both materials and labor. Operating supplies were assumed to be 15% of maintenance and repairs, and laboratory charges were estimated at 10% of operating labor. A 20-year plant life with no salvage value for the equipment was assumed to calculate depreciation using the straight line method. Taxes and insurance were estimated at 1% each of fixed capital investment.

Administrative costs were taken as 40% of operating labor; and plant overhead as 50% of the sum of operating labor plus supervisory labor plus maintenance and repairs. A small allowance was made for research and development which would be on a contract basis as needed. Interest was calculated assuming 2/3 of the total capital investment would be borrowed at a 10% annual rate.

Profitability Analysis

Because our previous study (2) indicated that the plant would not be profitable under current economic conditions, studies of the effects of several variables on profitability were made. For these studies it was assumed that the ash could be sold as a raw material for fertilizer manufacture at $17.50/ton. The current value of the synthesis gas was assumed to be $0.25/MSCF. (Well-head price of natural gas is now running as high as $0.70/MSCF.)

The cost of processing the manure assuming the above product values and a 350-day/year plant operation is $1.15 MM/year or $3.30/ton raw manure. The break-even sales price for the gas would be $0.56/MSCF. To yield a discounted-cash-flow (DCF) return of 16% before taxes, the gas would have to sell for $0.85/MSCF assuming a 20-year project life.

In the analyses that follow, the gas production cost including a 16% before-tax DCF return is calculated as a function of different variables. The DCF calculations were made using these assumptions:

(1) a constant annual income over the project life,
(2) a project life of 20 years,
(3) no salvage value at the end of the project life,
(4) continuous cash flow, and
(5) ash credit.

The DCF calculation procedure is outlined in Peters and Timmerhaus (16).

Figure 2 shows gas production cost as a function of plant size. With the 500 T/D plant as a base, capital investments for other size plants were calculated using the assumptions given previously. Operating labor varied from a minimum staff of four men/shift to 15 men/shift for a plant processing 5,000 T/D. A plant capacity of about 3,000 T/D would begin to be competitive at gas prices projected for the near future.

Variations in gas production cost with raw manure moisture content are shown in Figure 3. For moisture content less than the base case, it was assumed that excess char would be recycled back to the pyrolysis reactor for gasification to CO and H_2. For moisture content higher than the base case, part of the dried manure was used directly in the combustion reactor to provide additional heat for drying. It was assumed that capital investment and operating costs would remain the same as in the base case and that only the volume of gas produced would change.

Data from five additional sources (4,5,6,7,8) were used to check the variability of manure composition. Pyrolysis gas compositions and yields were calculated in the same manner as for the base case using the different manure analyses. While the gas yields were somewhat scattered, there appeared to be a slight increase as carbon content increased (Figure 4).

Figure 5 shows the effect changes in manure composition have on the gas production cost. Although it is difficult to predict the actual gas yields to expect from different manure compositions, it appears that composition would have only limited effects.

Figure 6 shows the effect of manure transportation costs and disposal credits on gas production cost. These are shown together since one is a cost/ton of raw material and the other a credit/ton of raw material with no overlap between the two. For the base case, transportation charges account for about 17% of the total annual income required to give a 16% before tax DCF return. To make such a return on investment at current gas prices, the plant would have to receive a credit of approximately $4.72/ton with no charge for transportation.

DISCUSSION

Several additional points should be made regarding the analysis. These include the economic environment under which the plant would operate, the end uses for the synthesis gas product, and trends in the manure composition analyses used.

Economic Environment

The economic environment assumed for the preceeding analyses has been of an industrial nature which requires payment of corporate income taxes and which also requires a considerable return on investment. Another economic environment can be considered which is that of a municipality or other governmental agency. In that case funds could be obtained at moderate interest rates by way of bonds with no net profit expected and no income tax to pay. For a municipal venture to be feasible, either the gas would have to sell at essentially the break-even sales price or feedlot operators charged a processing fee. For the base case these values would be $0.56/MSCF or $3.30/ton. While current economics do not appear promising for the base case, a larger plant coupled with rising natural gas prices predicted for the near future could readily change the situation.

Synthesis Gas Uses

For the production of synthesis gas to be a reasonable method of disposing of manure, there must be a market for the gas. Several alternatives are possible as the synthesis gas composition is similar to that obtained from partial oxidation of natural gas. The alternatives include ammonia or menthanol production and power generation.

A 500 T/D manure pyrolysis plant could ultimately produce approximately 11.5 MM SCF/day of H_2 by reforming the methane and converting the CO through the water-gas shift reaction. That amount of H_2 could be converted to about 180 T/D of ammonia. A typical ammonia plant (like the one operated by Farmland Industries, Inc., near Dodge City, Kansas) produces about 210,000 tons/year. Although it would take a large manure processing plant to completely supply such an installation, the base plant discussed above could supply about 30% of the H_2 requirements.

Methanol production requires a H_2 to CO molar ratio of at least 2 to 1. By shifting some of the CO to H_2 with the water-gas shift reaction and removing the CO_2 formed, enough gas with a 2 to 1 ratio can be produced from the 500 T/D plant to supply a 180 T/D methanol plant. That methanol production rate equals the rate from existing small low-pressure plants. Methanol can be used as a fuel, an industrial solvent, or a raw material for producing a variety of chemical products.

While manure could be burned directly to produce steam to generate electricity, environmental considerations many favor production of low-Btu synthesis gas for electricity generation. The synthesis gas route, by allowing easier recovery of nitrogen and sulfur, would prevent oxide emission problems. Preliminary investigations indicate that the synthesis gas route has at least as good thermal efficiency as direct combustion. Present Kansas feedlots could provide enough gas to supply about 13% of the state's residential demand.

Manure Composition

Several trends were noticed while studying the variations in gas production with the different manure compositions reported in the literature. One trend was that as the mole fraction of carbon increased, the fraction of oxygen decreased as would be expected (Figure 7). However, the mole fraction of hydrogen at different carbon fractions (also shown in Figure 7) remained about constant with increasing carbon content. Although the samples reported in the literature were from widely separated locations, differences in composition were not large. These trends support the finding that the amount of synthesis gas produced should not vary much with differences in manure composition.

PROCESS APPRAISAL

The economic analysis presented indicates improvements in the process or significant changes in the cost of producing synthesis gas by conventional methods would be needed for manure pyrolysis to become economically feasible. Another factor that will enhance the pyrolysis approach is increasing pressure for feedlot operators to dispose of waste in an esthetic manner without harming the environment.

One area of the conceptual design in which significant changes could be made is in the feed preparation and drying section. Since the dryers represent approximately 25% of the capital investment, reductions in drying equipment should affect the economics significantly. In addition, the excess heat produced in the combustion reactor could be used to generate steam to reduce the plant's utility consumption. A laboratory investigation of the effect of feed moisture content on the synthesis gas yield is needed to ascertain the amount of drying required. Another investigation should be made to determine the effect of manure particle size on gas yield.

The gas clean-up section of the proposed plant could be improved to reduce annual operating costs. One of the low pressure CO_2 absorption processes mentioned could significantly reduce utility costs, particularly if excess heat were available from the combustion reactor to provide the higher temperature required for solvent regeneration.

Although improvements in reactor design could be made, it is unlikely that costs in that area could be reduced significantly. However, other pyrolysis schemes such as one-stage gasification should be studied.

Finally, locations for a pyrolysis plant should be investigated. Since the larger feedlot concentrations are located away from industrial centers, the optimum balance between raw material supply and product utilization needs to be considered. Additionally, the benefits of large-scale plants may be realized to a certain extent by using crop wastes and other organic wastes as supplemental process feeds. The location of such waste sources should be included in determining plant sites.

REFERENCES

1. Walawender, W.P., L. T. Fan, C. R. Engler, and L.E. Erickson, "Feedlot Manure and Other Agricultural Wastes as Future Material and Energy Resources: II. Process Descriptions," Report No. 45 of the Institute for Systems Design and Optimization, Kansas State University, Manhattan, Kansas, March 1973.

2. Walawender, W.P., L.T. Fan, C.R. Engler, and L.E. Erickson, "Feedlot Manure and Other Agricultural Wastes as Future Material and Energy Resources: III. Economic Evaluations," Report No. 46 of the Institute for Systems Design and Optimization, Kansas State University, Manhattan, Kansas, July 1973.

3. Appell, H.R., Y.C. Fu, S. Friedman, P.M. Yavorsky and I. Wender, "Converting Organic Wastes to Oil," Agricultural Engineering, 53, 3, p.17ff, 1972

4. Herzog, K.L., H.W. Parker and J.E. Halligan, "Synthesis Gas from Manure," presented at 73rd National Meeting, AIChE, June 1973.

5. Cox, J.L., E.J. Hoffman, R.W. Hoffman, W.G. Willson, J.A. Roberts and D.L.Stinson, "Gasification of Organic Waste," presented at 165th National Meeting, A.C.S., April 1973.

6. Fu, Y. C., E. G. Illig and S. J. Metlin, "Conversion of Manure to Oil by Hydrotreating," presented at 166th National Meeting, A.C.S., August 1973.

7. Kiang, K. D., H. F. Feldmann and P. M. Yavorsky, "Hydrogasification of Cattle Manure to Pipeline Gas," presented at 165th National Meeting, A.C.S., April 1973.

8. Schlesinger, M. D., W. S. Sanner and D. E. Wolfson, "Energy from the Pyrolysis of Agricultural Wastes," presented at A.C.S. Meeting, New York, August 1972.

9. Appell, H. R. and I. Wender, "The Conversion of Urban Refuse and Cellulosic Wastes to Oil," U.S. Bureau of Mines Pittsburgh Energy Research Center, progress report (1970).

10. "Solid Waste: A New Natural Resource," (Staff of the Fluidized Bed Gasification Project), Dept. of Chemical Engineering, West Virginia University, Morgantown, W. Va., May 1971.

11. Wen, C. Y., R. C. Bailie, C. Y. Lin and W. S. O'Brien, "Production of Low BTU Gas Involving Coal Pyrolysis and Gasification," presented at 165th National Meeting, A.C.S., April 1973.

12. Guthrie, K. M., "Capital Cost Estimating," Chemical Engineering, 76 (March 24, 1969), p. 114.

13. Grzelak, R. J., "C-E Raymond Flash Drying Systems," Cost Engineering, July 1965, p. 4.

14. Peter, M. S. and K. D. Timmerhaus, Plant Design and Economics for Chemical Engineers, 2nd Ed., McGraw-Hill, New York, 1968, p. 112.

15. Ibid., pp. 126-138.

16. Ibid., pp. 238-242.

Table I

Compositions of Manure and Municipal Solid Waste

	Manure, wt %[a]		Municipal Solid Waste, wt. %[b]
Constituent	As Used	Dry, ash-free	Dry, ash-free
Carbon	18.9	52.2	55.1
Hydrogen	2.3	6.4	7.3
Oxygen	13.3	36.9	36.6
Nitrogen	1.2	3.3	0.8
Sulfur	0.4	1.2	0.2
Ash	13.9	----	----
Water	50.0	----	----

[a] Appel, et al. (3)

[b] Appel and Wender (9)

Table II

Synthesis Gas Production From 500 tons/day Plant

Constituent	MM SCF/day
CO	4.55
H_2	5.20
CH_4	0.98
CO_2	1.43
H_2S	0.10
NH_3	0.62
H_2O	0.80

Table III

Thermochemical Data

Constituent	ΔH_c (Btu/lb)	C_p (Btu/lb °F)
Dry, ash-free manure	-8,750	0.75
Char	-17,060	0.34
Ash	---	0.23
Sand	---	0.28

Table IV

Major Process Equipment

Equipment Number	Name	Size	Comments
Reactors			
R-101	Pyrolysis	12' φ x 35'	C.s. shell, lined with 18" firebrick
R-102	Combustion	16' φ x 44'	
Towers			
T-101	Water quench	5' φ x 15'	Pack with 10' of 2" Rashig rings
T-102	CO_2 absorber	11' φ x 28'	Pack with 20' of 2" Rashig rings
T-103	Regenerator	12" φ x 25'	Pack upper half with 10' of 2" Raschig rings
Solids Removal			
S-101	Pyrolysis gas clean-up	15,000 CFM	Multicyclones
S-102	Combustion gas clean-up	26,000 CFM	Louvered collector followed by multicyclones
Dryers			
DR-101 A&B	Flash dryer and pulverizer	20,000 lb/hr water removal	Gas fired for start up
Compressors			
C-101	Pyrolysis recycle	175 HP	
C-102	Combustion air	1300 HP	
C-103	Pyrolysis gas	625 HP	
Exchangers			
E-101 A-C	Ash cooler	250 ft^2	16" x 20' hollow flight conveyor
E-102	Air preheater	3410 ft^2	Plate fin exchanger
Pumps			
P-101	Water quench	450 GPM, 75 psi	
P-102 A-C	CO_2 absorption	11,000 GPM, 300 psi	
Storage Bins			
V-101 A&B	Dry manure	15' φ x 20'	Cone bottom hopper
V-102 A&B	Ash	15' φ x 20"	Cone bottom hopper
Conveyors			
CV-101	Raw manure	16 CFM	12" screw conveyor

Table V

Installed Costs for Major Process Equipment Items

Item	M$
DR-101 A & B	1,098
R-101 and R-102	445
S-101 and S-102	106
Compressors	747
Towers	308
Pumps	353
Miscellaneous[a]	68
Total	3,125

[a]Includes heat exchangers, storage hoppers and conveyors

Table VI

Capital Investment Estimate

	M$
Bare Module Costs	
Processing Equipment	3,125
Buildings	280
Site Development	125
Offsite Facilities	545
Total Bare Module Cost	4,075
20% Additional Item Allowance	815
Subtotal	4,890
5% Contractor	245
Total Module Cost	5,135
10% Contingency	514
Fixed Capital Investment	5,649
Working Capital (15%)	847
Total Capital Investment	6,496

Table VII

Annual Manufacturing Costs

	M$	M$
Raw Material		
Manure Transportation	525	
Sand Replacement	5	
Total Raw Materials		530
Utilities		
Electricity	370	
Make-up Water	92	
Total Utilities		462
Direct Production Cost		
Operating Labor	171	
Supervisory Labor	17	
Maintenance & Repairs	424	
Operating Supplies	64	
Laboratory Charges	17	
Total Direct Production Cost		693
Fixed Charges		
Depreciation	283	
Taxes & Insurance	113	
Total Fixed Charges		396
Overhead		
Administration	68	
Plant Overhead	306	
Research & Development	50	
Interest	433	
Total Overhead		857
Total Annual Manufacturing Cost		2,938

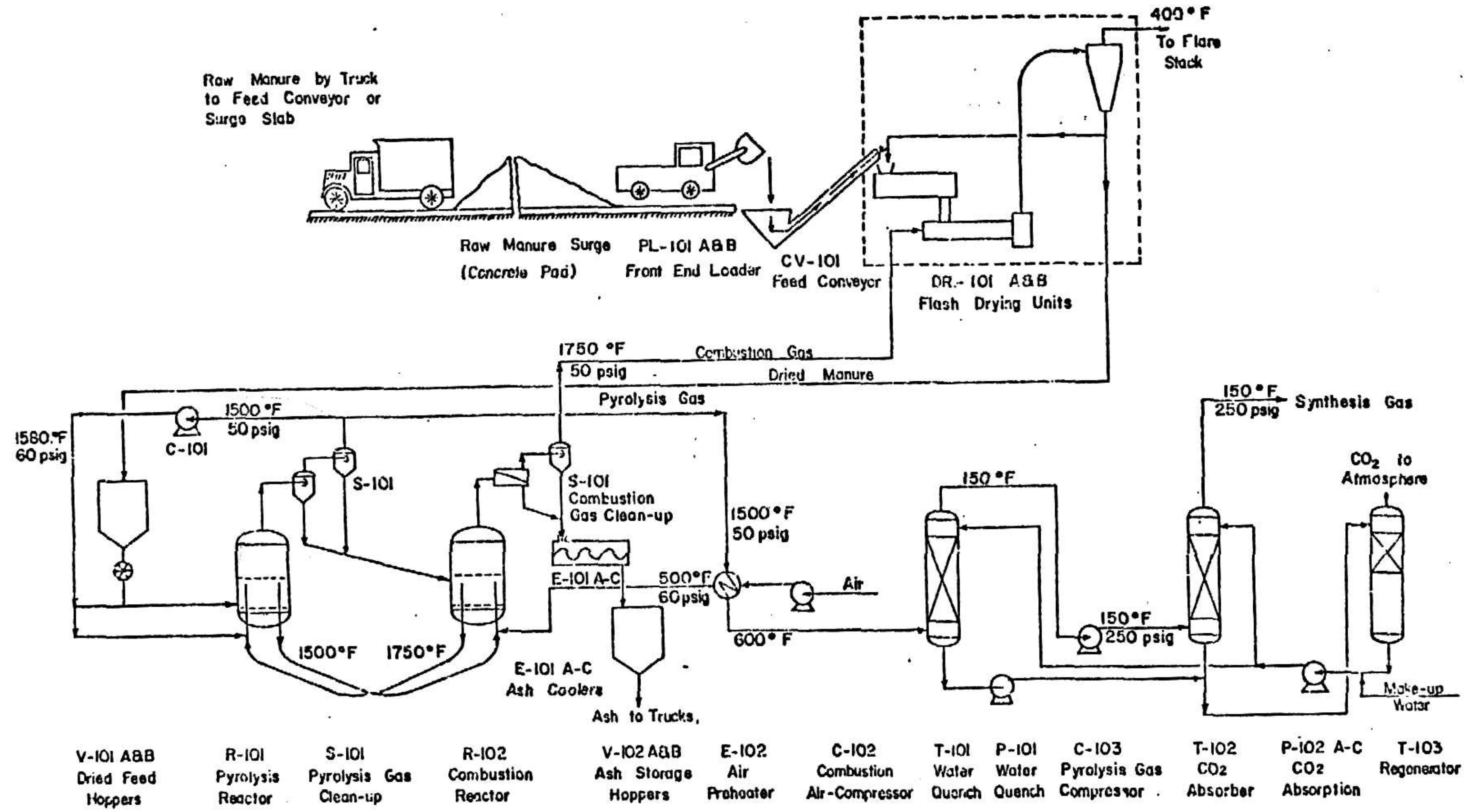


Figure 1. Process Flow Plan

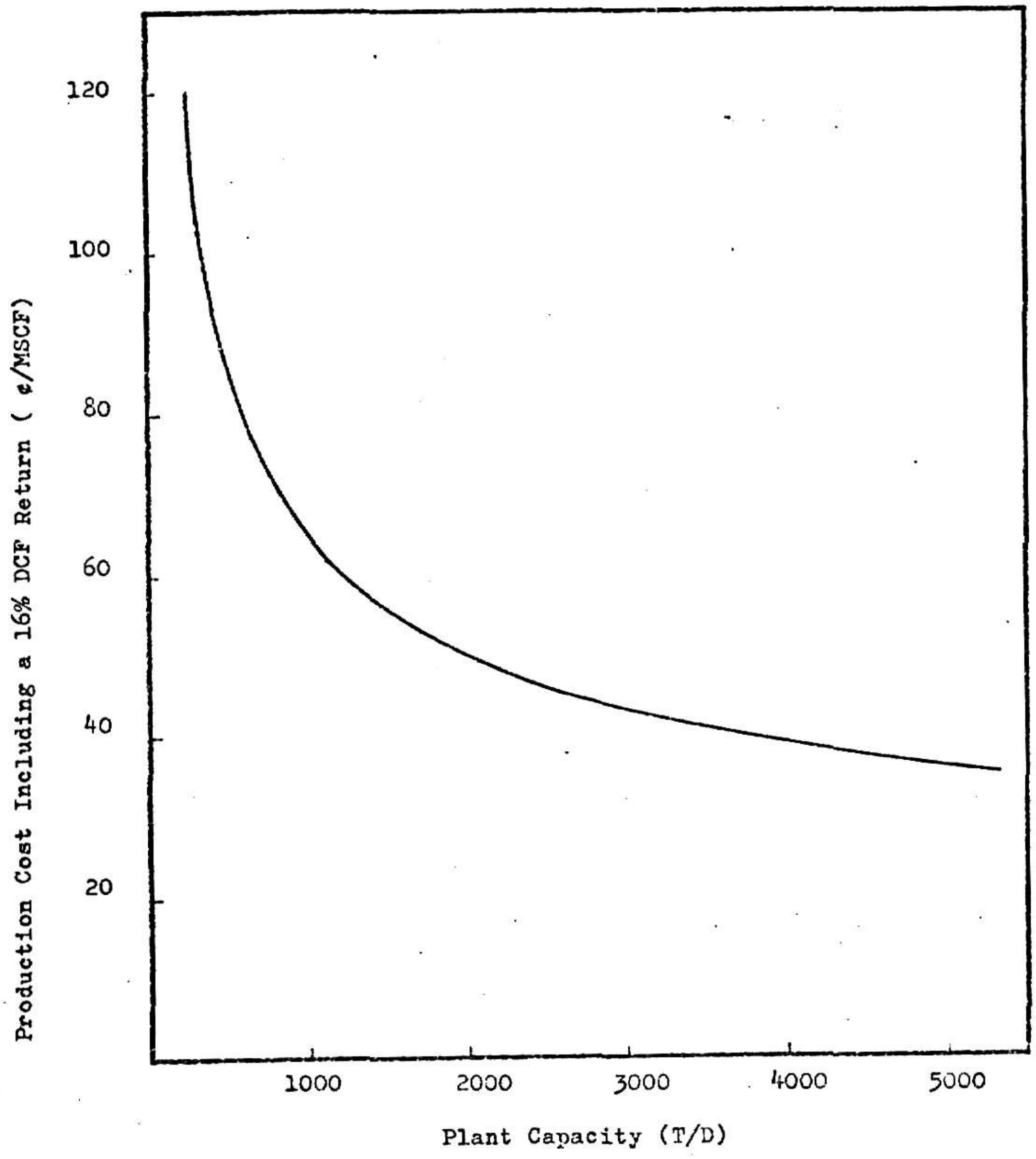


Figure 2. Plant Capacity Effect on Gas Production Cost

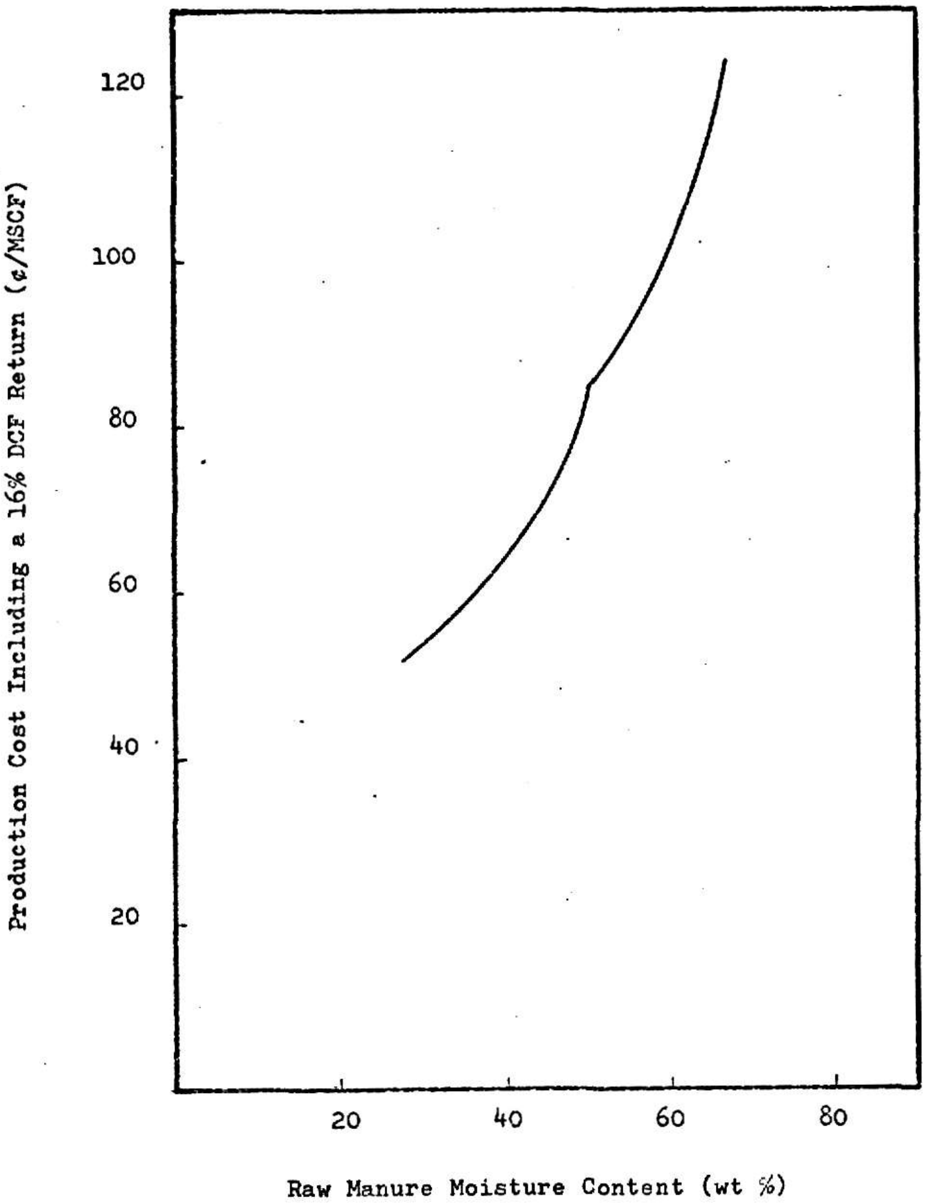


Figure 3. Manure Moisture Content Effect on Gas Production Cost

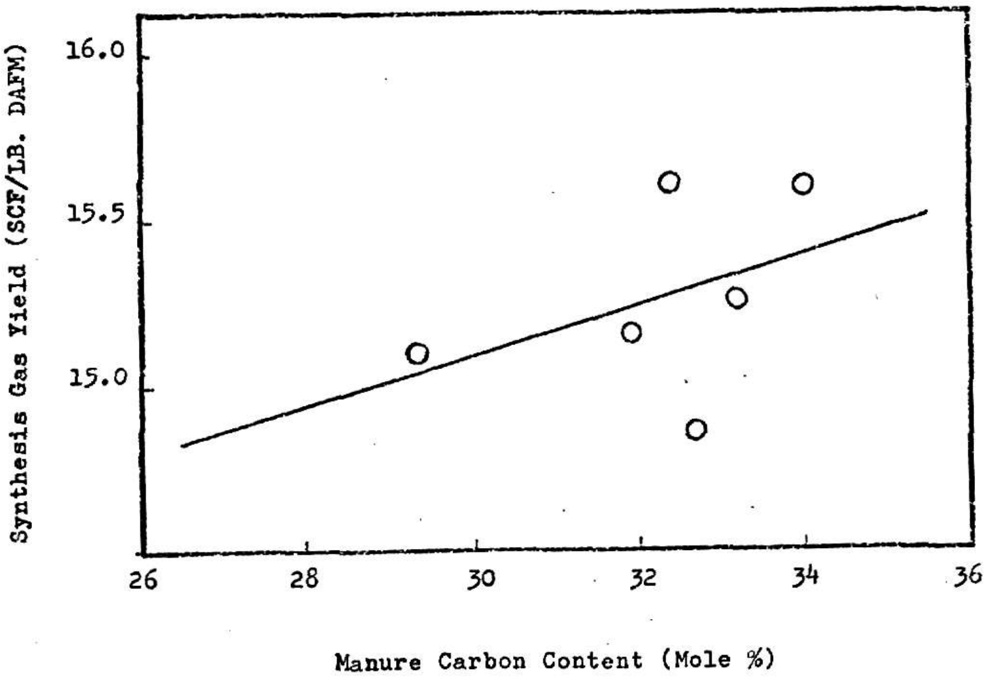


Figure 4. Synthesis Gas Yield as a Function of Carbon Content

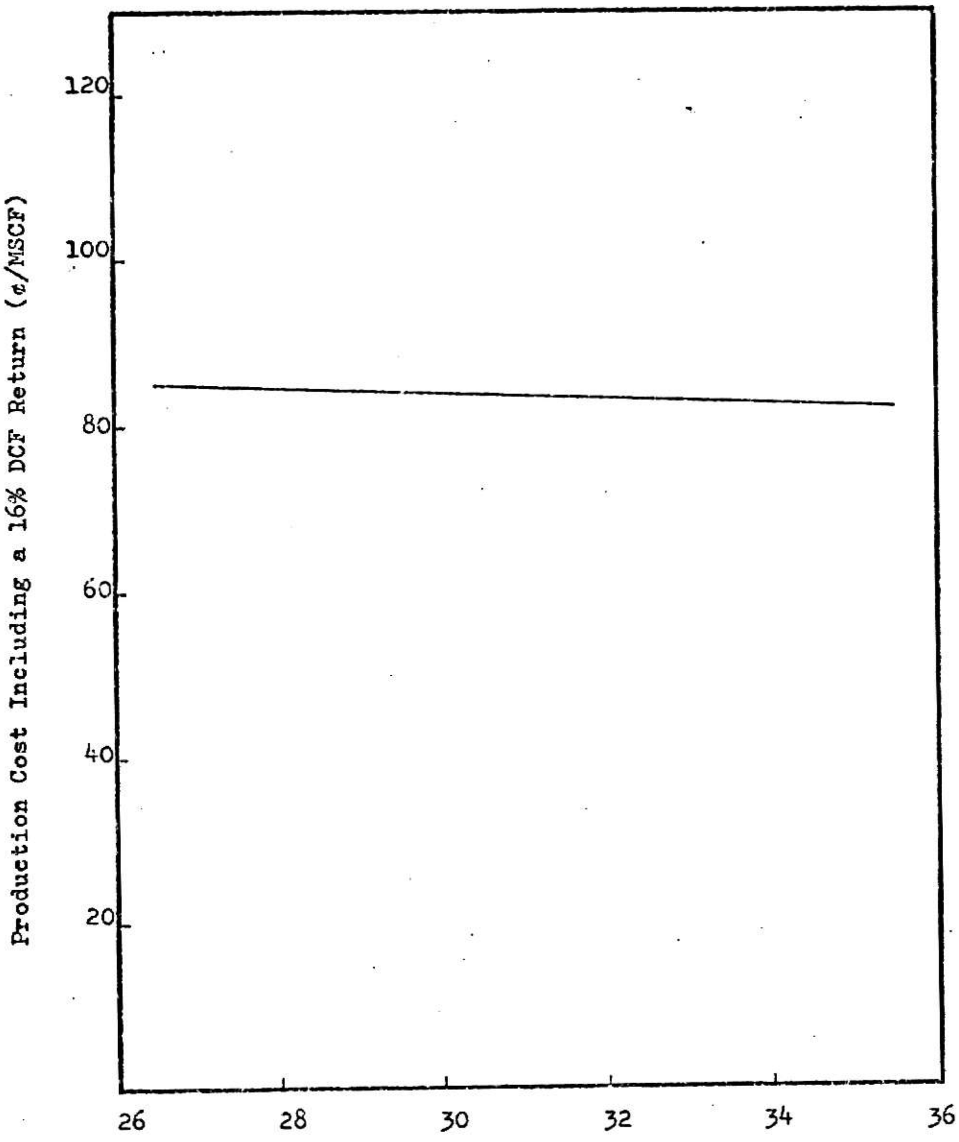


Figure 5. Manure Carbon Content Effect on Gas Production Cost

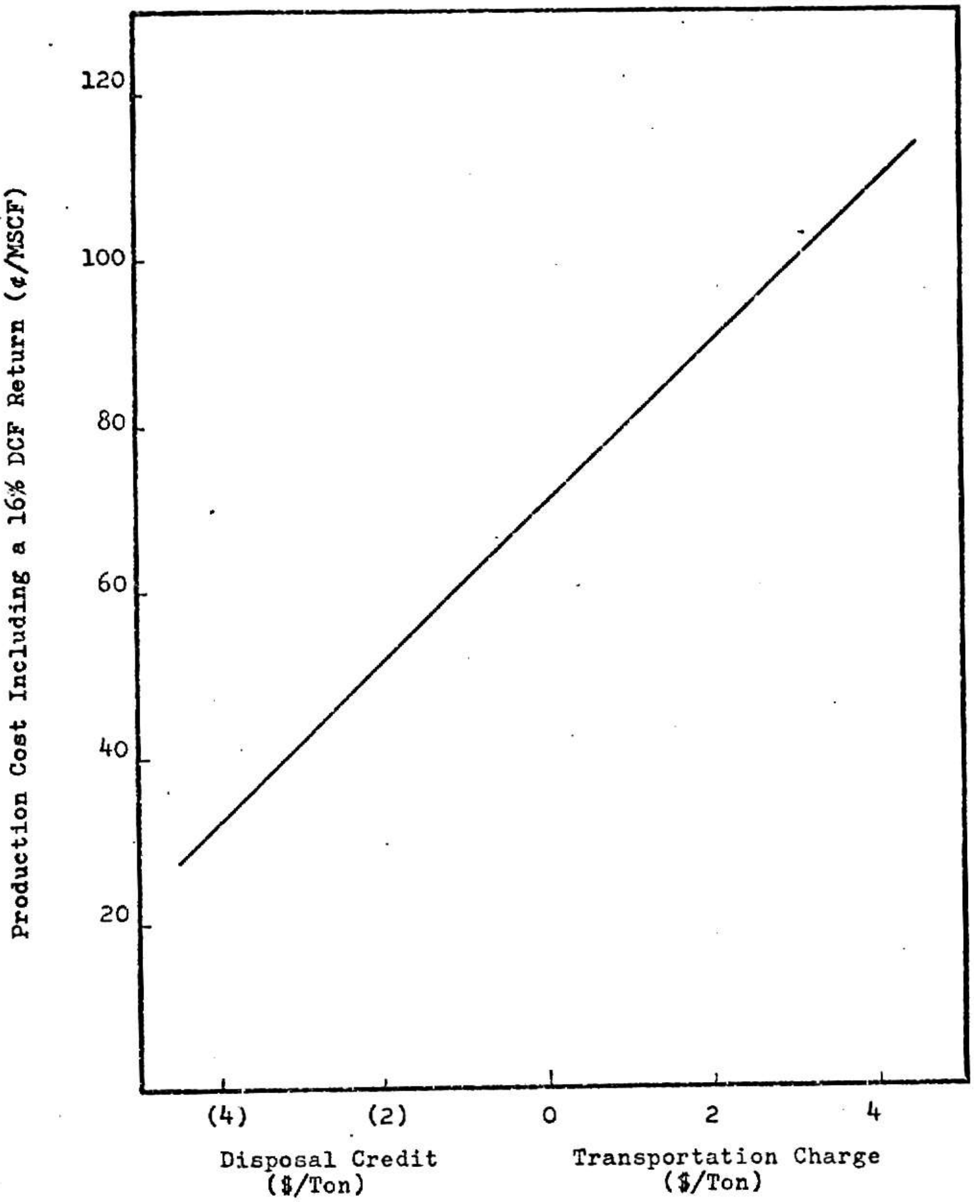


Figure 6. Disposal Credit and Transportation Charge Effect on Gas Production Cost

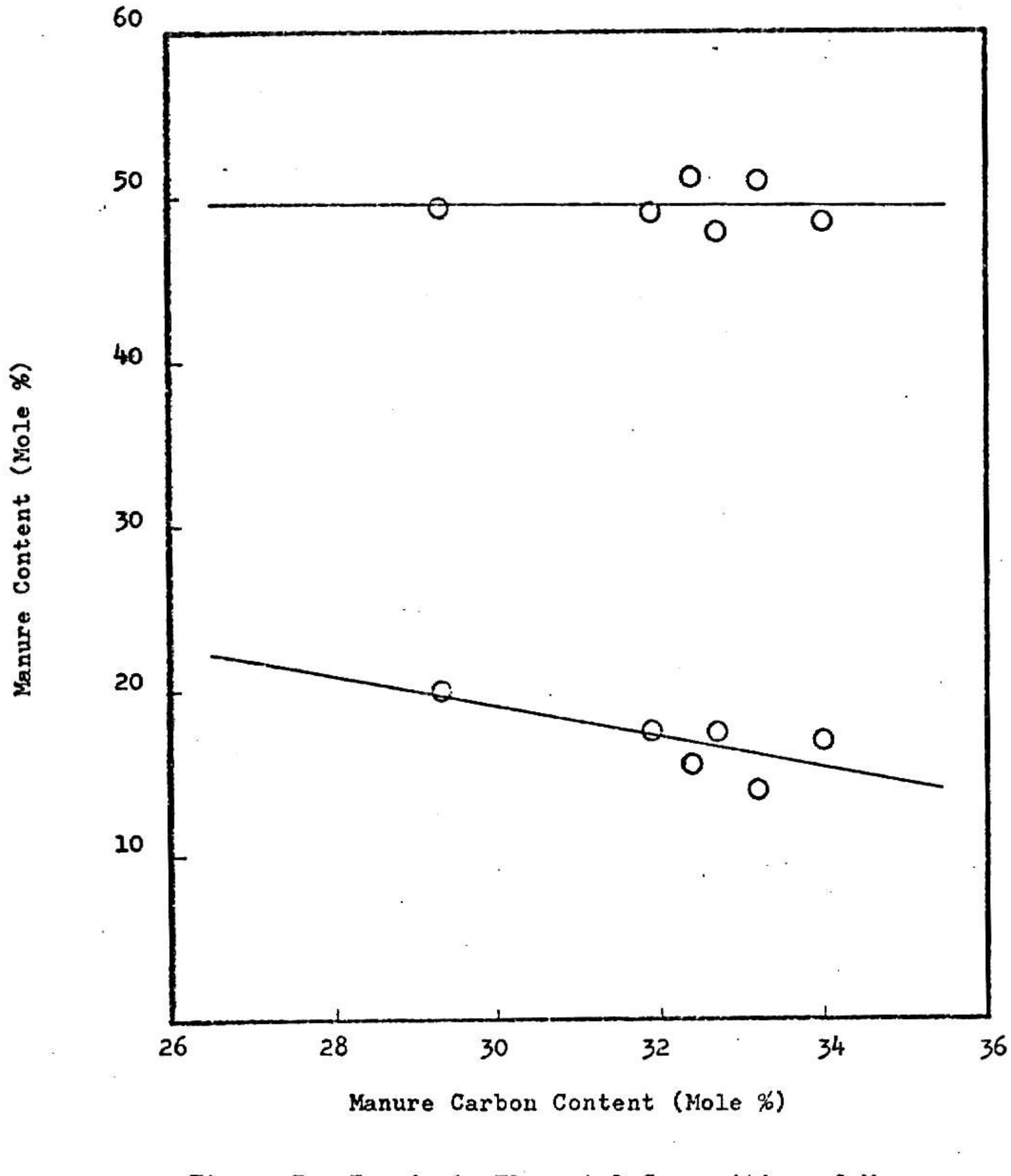


Figure 7. Trends in Elemental Composition of Manure

CHAPTER III

POTENTIAL APPLICATIONS IN SOUTHWESTERN KANSAS

INTRODUCTION

A conceptual design of a plant to produce synthesis gas (CO and H_2 mixture) from feedlot manure via flash pyrolysis was analyzed in Chapter II. While the analysis showed that small-scale plants processing 500 T/D (tons per day) of dry manure would not be economically attractive, larger plants could produce synthesis gas at a cost competitive in the current economic environment. To determine the potential for large capacity pyrolysis plants in Kansas, feedlot manure generation and possible synthesis gas usage in the southwest quarter of the state were studied.

Feedlot capacity data were collected and organized into regions that could support moderate- to large-scale pyrolysis plants. Synthesis gas outputs for the various plants were calculated and the information used to estimate sizes of ammonia production or electricity generating facilities that could be supported in each region. These applications for southwest Kansas are described following a summary of the manure pyrolysis process.

MANURE PYROLYSIS PROCESS

The manure pyrolysis process described in Chapter II is shown in Figure 1 as a flow plan with the equipment designations explained in Table I. Briefly, the process consists of a feed preparation section where incoming manure is dried and ground, a pyrolysis section where the manure is gasified at 1500°F in a fluidized bed reactor, and a gas clean-up section where CO_2 and other undesirable gaseous byproducts are removed. Heat requirements for the pyrolysis reactor and manure drying are supplied by burning char (a solid byproduct of the pyrolysis reaction) in a fluidized bed combustion reactor.

The conceptual design was based on an incoming manure moisture content of 50% which corresponds to semicomposted manure in a non-arid climate. However, for this study a moisture content of 40% was used to reflect drier conditions in southwest Kansas. Approximately 200,000 head of cattle would be required to support a plant processing 500 T/D of manure (dry basis). A plant of that size would produce about 14 MMSCF/D (million standard cubic feet per day) of clean synthesis gas with the composition given in Table II.

Capital investment costs for the conceptual design were estimated on a mid-1974 basis and are shown in Figure 2 as a function of plant capacity which is on a dry basis. Annual operating costs for a moderately large plant processing 2500 T/D would be approximately $10 million as summarized in Table III. Transportation costs were estimated assuming the manure would be available within an average hauling distance of 25 miles. The cost basis for transportation was $0.06/T/mile. An interest rate of 10% was used, and the other costs were simply taken from the previous chapter.

For this study capital investment and operating costs were assumed to be independent of incoming manure moisture content. However, the gas production rate and composition were dependent on moisture content. In all cases studied, the incoming manure was assumed to be 40% moisture so the gas composition given in Table II applies.

FEEDLOT CAPACITIES

Data provided primarily by the Liberal Chamber of

Commerce (1) were used to develop feedlot capacities for regions in western Kansas. Approximate capacities within a 50-mile radius of three major feeding centers are shown in Figure 3 and listed in Appendix C. Capacities in overlapping areas were included in the totals for each of the regions. The total for Liberal also included feedlots in Oklahoma and Texas. Additionally, Figure 3 shows total feedlot capacities within 70- and 100-mile radii around Liberal.

Cattle produce about 5.5 lbs of manure (dry basis) per head daily. For the 50-mile radius data, synthesis gas plants processing up to 1800 T/D could be supported. The 100-mile radius around Liberal could support a plant processing 4125 T/D. These capacities are based on consumption of all manure generated in an area.

Data from Chapter II were used to estimate synthesis gas costs that would yield a 16% before-tax discounted-cash-flow (DCF) return. All operating costs were estimated on a mid-1974 basis, and credit was taken for the value of the ash byproduct. Transportation costs were varied to reflect average distances manure would be hauled in each of the regions.

Table IV lists maximum possible plant capacities for the various regions along with the synthesis gas production rates and gas prices required to yield a 16% before-tax DCF return. For plants processing manure from regions of the same size, the gas price decreases slightly with increased plant capacity. However, for plants drawing from larger regions, increased transportation costs offset savings from larger plant capacities so that gas prices remain essentially the same. The last entry in Table IV shows that gas price increases significantly when only a fraction of the manure from a region is available for pyrolysis. Unfortunately data were not available to make a direct comparison of the synthesis gas price of $0.50/MSCF (thousand standard cubic feet) or $1.25/MM Btu to the cost of producing it by conventional processes such as natural gas reforming.

The cattle feeding industry is closely tied to both the fertilizer industry and electricity usage. Large quantities of fertilizer are used to produce feed for the cattle, much of which is grown near the feedlots. Electricity is used by feedlot operators for numerous parts of the operation such as drying and grinding grain and feeding the cattle. Use of synthesis gas from manure pyrolysis for these two industries was examined in detail. Synthesis gas can be used as the hydrogen source for producing ammonia fertilizers or it can be burned in gas turbines to produce electricity.

AMMONIA PRODUCTION

In ammonia plants natural gas is used to provide hydrogen for the ammonia synthesis reaction and also to provide energy for the operation. A typical ammonia plant uses about 40 MSCF of natural gas per ton of ammonia produced with approximately 23 MSCF/T used for reforming to synthesis gas and 17 MSCF/T for energy supply. In terms of dry manure 1.5 T/T NH_3 are required for fuel and 2.7 T/T NH_3 are required to supply H_2 for a total of 4.2 T/T NH_3. Table IV also lists the sizes of ammonia plants that could be supported by each region. The synthesis gas could be transported easily via pipeline from a gasification plant to an ammonia plant in another location.

The yield from converting manure to ammonia is about 390 lbs N/T dry manure. For raw manure from a commercial cattle feedlot, Herron and Erhart (2) reported an equivalent nitrogen content of 22 lb N/T manure when applied directly to the field. The manure used by Herron and Erhart had 22% moisture content. When corrected to a dry basis, the nitrogen equivalent is 28 lb N/T dry manure. There are advantages to using manure directly as a fertilizer from the standpoint of soil conditioning and availability of other nutrients. However, the application of manure on fields is expensive, and application at a high rate on a limited area is not desirable. Conversion of manure to ammonia would increase its equivalent nitrogen value by nearly 17 times. Additionally, the ash byproduct from manure pyrolysis could be used to supply other plant nutrients after suitable processing. Analysis of manure ash by the U.S. Bureau of Mines (3) indicates that phosphate and potassium contents are about 15 and 10 percents respectively. The remainder of the ash is mostly lime, silica and alumina.

The annual usage of nitrogen from commercial fertilizers was estimated for the southwest quarter of Kansas in order to determine the potential contribution of manure synthesis gas. Sales data from the Kansas Department of Agriculture Control Division (4) were used for the estimate. Based on sales reported for that area, approximately 175,000 T of nitrogen are used annually which is equivalent to 212,000 T of ammonia. An ammonia plant supported by the manure generated in region 1 or 4 (see Table IV) could supply about 70% of the annual nitrogen requirements while the large plant projected for region 5 of Table IV would supply a surplus.

ELECTRICITY GENERATION

Synthesis gas produced from pyrolysis of feedlot manure has a heating value of approximately 400 Btu/SCF. Power generating facilities now using natural gas (heating value about 1000 Btu/SCF) could be readily modified to use the synthesis gas directly or blended with natural gas. Synthesis gas production data were used to estimate the potential generating capacities given in Table IV. The capacities are peak-load ratings for plants utilizing all of the gas produced.

Most existing generating facilities in western Kansas are small and are not located in areas with large feedlot capacities nearby. Exceptions are generating plants at Dodge City and Liberal with capacities of 182.4 and 74.2 megawatts respectively. A manure pyrolysis plant at Dodge City could provide about 40% of the rated capacity there. A pyrolysis plant near Liberal could support additional generating capacity for that area.

For a small regional utility, a municipal type of venture could be used to build the manure gasification plant. In such a case no profit is required and no taxes are paid. A power plant rated at 30 MW capacity could be supported by a 500 T/D manure pyrolysis plant. The cost of producing the gas would be about \$0.42/MSCF or \$1.05/MM Btu. A municipal venture analysis for the larger plant in region 1 of Table IV would reduce the cost of gas to about \$0.31/MSCF or \$0.78/MM Btu.

Feedlot manure could be used as fuel for the small generating plants (1 to 5 MW) in western Kansas. However, it would not be economical to gasify the manure since low-capacity pyrolysis plants are quite expensive. An alternate route would be direct combustion of the manure to fire steam boilers. The economic feasibility of that approach was not studied.

SUMMARY

The potential for utilization of synthesis gas from moderate- to large-scale manure pyrolysis plants appears to be good for southwestern Kansas. In particular, either producing ammonia or generating electricity could directly benefit feedlot operators and other residents of the area. The cost of producing gas for a private venture would be somewhat higher than could be economically justified at present, but realization of projected costs for natural gas could change that in the near future. Municipal funding would reduce the cost of the gas and make it competitive with new natural gas sources.

While this analysis is optimistic about the application of manure pyrolysis to feedlot waste disposal, it is not likely the assumption of using all manure produced in a region for a pyrolysis plant would be valid. Also, the number of cattle actually on feed in a given region can drop sharply because of economic conditions. When only a portion of the manure is available (region 5A, Table IV), the cost of producing synthesis gas goes up considerably.

REFERENCES

1. Liberal Chamber of Commerce, "Liberal, Heart of the Beef Belt," Liberal, Kansas, December 1973.

2. Herron, G.M. and A.B. Erhart, "Value of Manure on an Irrigated Calcareous Soil," Soil Science Society Proceedings 1965, p. 278.

3. Davis, E.G., I.L. Feld, and J.H. Brown, "Combustion Disposal of Manure Wastes and Utilization of the Residue," Bureau of Mines Solid Waste Research Program, Technical Progress Report 46, January 1972.

4. Kansas Department of Agriculture, Control Division, "Tonnage of Commercial Fertilizer Reported by Registrants as Sold in Kansas in the Spring of 1973, by Counties," Topeka, Kansas, 1973.

Table I

Explanation of Symbols in Figure 1

Item	Equipment Description
PL-101 A&B	Payloaders for moving raw manure
CV-101 A&B	Feed conveyors to supply dryers
DR-101 A&B	Flash dryers combined with cage mill grinders
V-101 A&B	Dried manure storage hoppers
R-101	Fluidized bed pyrolysis reactor
C-101	Recycle gas compressor
S-101	Cyclone system to remove solids from pyrolysis gas
R-102	Fluidized bed char combustor
S-102	Cyclone system to remove solids from combustion gas
E-101 A-C	Cooling conveyors to remove ash from process
V-102 A&B	Ash storage hoppers
E-102	Heat exchanger to cool pyrolysis gas with combustion air
C-102	Combustion air compressor
T-101	Pyrolysis gas quench tower
C-103	Pyrolysis gas compressor
P-101	Pump to remove water from quench tower
T-102	CO_2 absorption tower
T-103	Regenerator tower to release CO_2 from water
P-102 A-C	Pumps to recycle water to quench and absorption towers

Table II

Clean Synthesis Gas Production
from 500 T/D (dry basis) Plant

Component	Mole %
CO	40
H_2	48
CH_4	7
CO_2	5

Table III

Annual Manufacturing Costs
for 2500 T/D (dry basis) Plant

	M$
Raw Materials	
Manure Transportation	2,625
Sand	16
Utilities	
Electricity	1,866
Make-up Water	467
Direct Production Costs	
Operating Labor	425
Supervisory Labor	42
Maintenance & Repairs	1,123
Operating Supplies	169
Laboratory Charges	42
Fixed Charges	
Depreciation	749
Taxes & Insurance	299
Overhead	
Administration	170
Plant Overhead	795
Research & Development	108
Interest	1,148
Total Annual Manufacturing Cost	10,044

Table IV

Regional Data for Manure Pyrolysis Plants

Region Number	Region Center	Radius (mi)	Capacity (T/D)*	Gas Rate (10^9 SCF/yr)	Gas Price ($/MSCF)	NH_3 Capacity T/D	NH_3 Capacity MT/yr	Power Plant Rating (MW)
1	Garden City	50	1800	18.0	0.45	430	150.0	140
2	Dodge City	50	1050	10.5	0.52	250	87.5	80
3	Liberal	50	1200	12.0	0.50	290	101.5	95
4	Liberal	70	1900	19.1	0.48	460	161.0	150
5	Liberal	100	4125	41.3	0.50	990	346.5	325
5A	Liberal	100	2060**	20.7	0.58	495	173.2	160

*Dry basis.

**Assuming only half of the manure in the region is available for gasification.

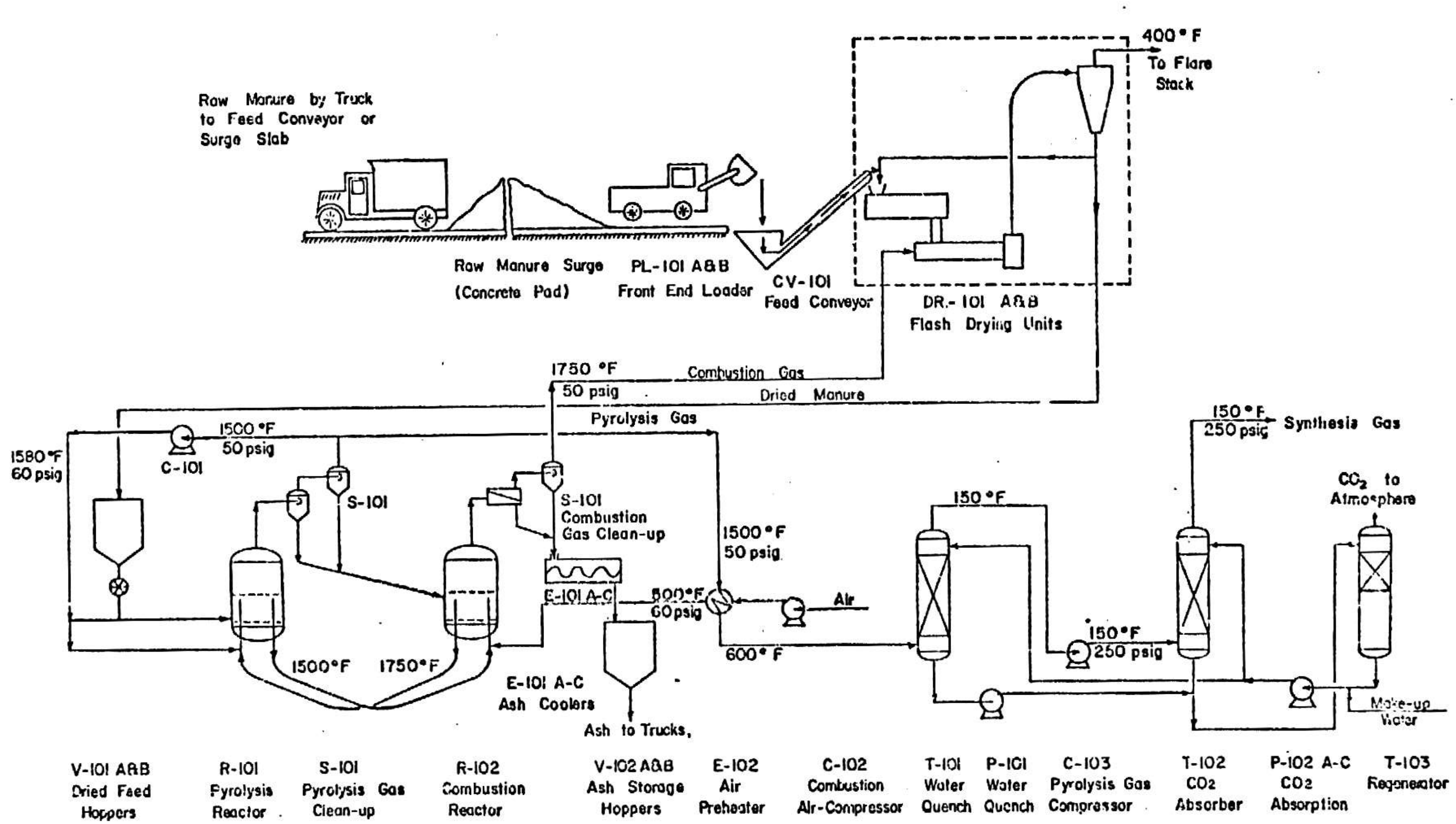


Figure 1. Process Flow Plan

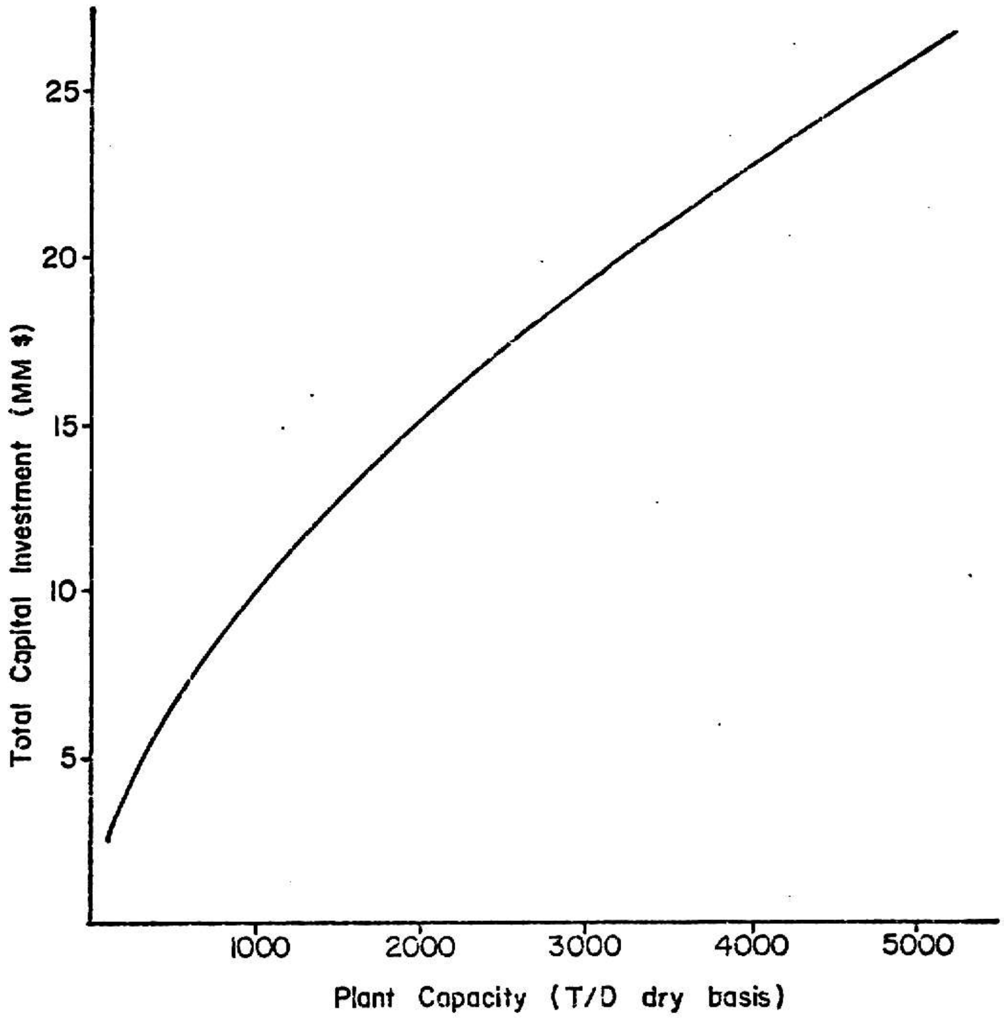
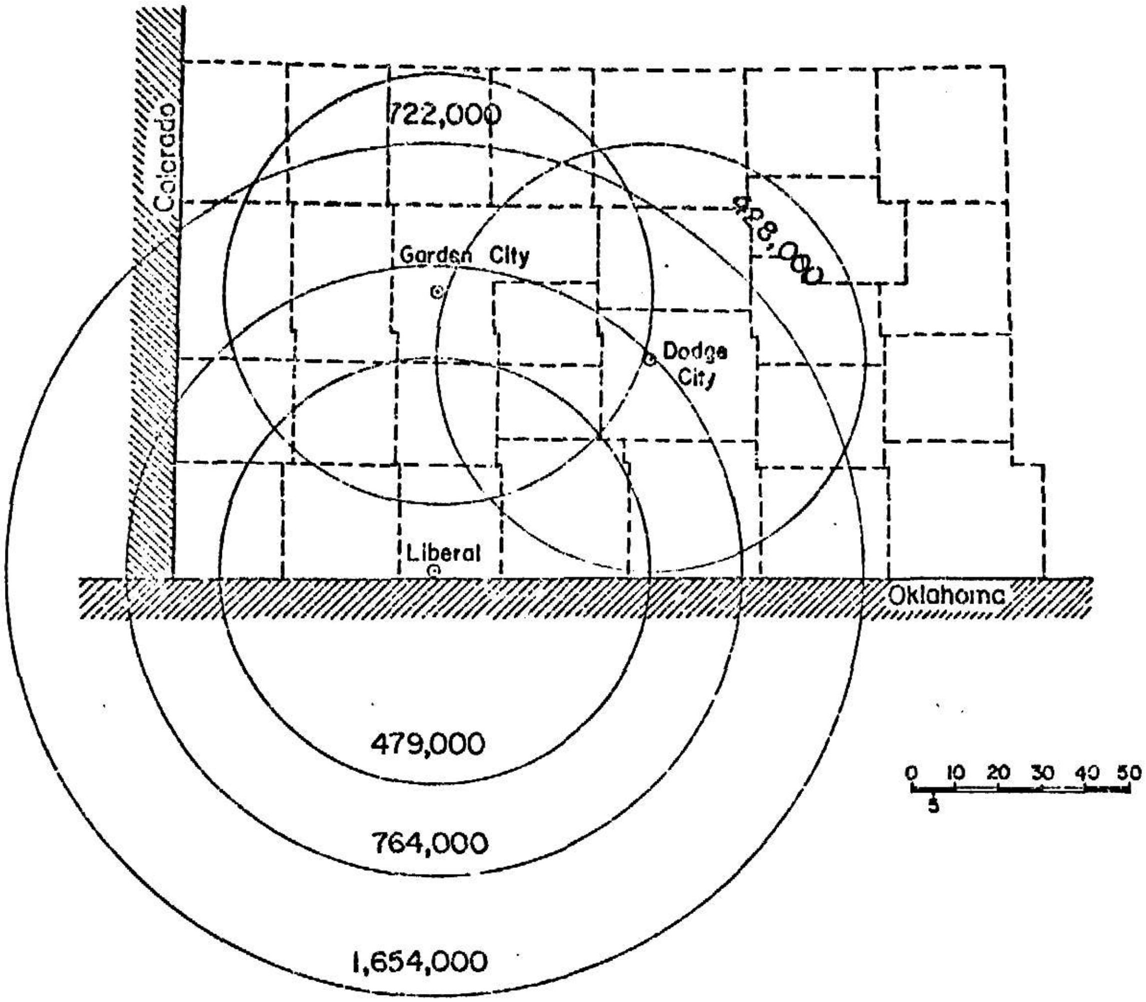


Figure 2. Capital Investment Cost vs. Plant Capacity

Figure 3. Cattle Concentrations in Southwestern Kansas



CHAPTER IV

SIMULATION UNIT DEVELOPMENT

INTRODUCTION

On the basis of the conceptual design study for manure pyrolysis (Chapter II) and its potential applications in southwestern Kansas (Chapter III), the possibility of implementing such a process appears fairly good. However, before a pilot plant can be built, additional information concerning the operation of the central part of the process, the fluidized beds, is required. This phase of the work concentrated on the development of a simulation unit to observe the operating characteristics of the bed, obtain pertinent pilot plant design data, and gain general operating experience with a fluidized bed.

In selecting the equipment for the simulation unit, an effort was made to keep costs low while at the same time provide enough capacity so that duplication of the fluidization column dimensions for a pilot scale gasifier would allow processing an estimated one-half ton per day of dry manure. The column diameter was selected as the minimum that would allow easy application of refractory material during construction of the actual gasifier and was found to be compatible with the desired processing rate.

Because a mixture of manure and solid pyrolysis byproducts would be inhomogeneous, sand was selected as an inert material to make up the bulk of solids in the bed. Additionally, sand would be a good heat transfer medium for use in the pyrolysis reactor. Due to the uncertainties involved in scaling from one size fluidized bed to another, the length-to-diameter ratio for the expanded bed in the simulation unit was set at three to give a reasonable working volume of sand. The ratio of sand to feed for the pilot scale gasifier would be about 88 pounds sand per pound of dry, ash-free manure fed per minute whereas the ratio used in the conceptual design was only 12.

The air compressor was selected to give a superficial velocity of about 100 cm/sec. Other equipment items were chosen for compatibility with the above requirements. Although operating conditions for the simulation unit were well defined, the quality of fluidization could not be predetermined.

Because of the lack of information for irregular shapes and sizes of particles, elutriation of manure, ash and char particles was of particular interest. Elutriation is the selective removal of one size or type of particle from a fluidized bed by the gas stream. Numerous investigators have studied the process of elutriation and excellent summaries of much of the work have been presented by Kunii and Levenspiel (1) and by Leva and Wen (2). While several investigations have been made with particles in the size range of sand, the beds were kept shallow or the superficial velocities low in order to maintain smooth fluidization. However, Wen and Hashinger (3) have shown that operating a bed in a slugging mode can significantly reduce the elutriation rate.

Elutriation has been found to follow a first order rate expression as long as the gas exits above the transport disengaging height (1). The rate of removal of solids of size d_p from the bed can be written as

$$-\frac{1}{A}\frac{d\,W(d_p)}{dt} = \frac{K^*\,W(d_p)}{W} \qquad (1)$$

where $W(d_p)$ is the weight of the solids of size d_p in the bed, A is the cross-sectional area of the bed, t is time, W is the total weight of solids in the bed and K^* is the specific elutriation rate constant. For experimental work it is more convenient to use the elutriation constant

$$K = \frac{K^*\,A}{W} \qquad (2).$$

The elutriation rate expression can be integrated for a batch experiment to give

$$\frac{W(d_p)}{W_o(d_p)} = \exp(-Kt) \qquad (3)$$

if the amount elutriated is small in relation to the total weight of the bed (<20%). From experimental measurements of the elutriation rate, a plot of ln $(W(d_p)/W_o(d_p))$ vs t can be constructed. The resulting curve should be linear near $t = 0$ with the negative value of the slope being the elutriation constant K.

For continuous feeding with no solids outflow except by elutriation (see Figure 1), a material balance for a given particle size yields

$$F_o(d_p) - KW(d_p) = \frac{dW(d_p)}{dt} \qquad (4)$$

which can be integrated to give

$$W(d_p) = \frac{F_o(d_p)}{K}(1-e^{-Kt}) \qquad (5)$$

with the boundary condition that $W(d_p) = 0$ at $t = 0$. At steady state (infinite time) the elutriation constant can be obtained as

$$K = \frac{F_o(d_p)}{W(d_p)} \qquad (6).$$

Literature correlations have related K* rather than K to system properties since it has been shown by Wen and Hashinger (3) that K depends on the bed dimensions whereas K* is independent of bed dimensions. The most widely used correlations for predicting K* are those presented by Wen and Hashinger (3) and Yagi and Aochi (4). Both correlations relate a dimensionless elutriation constant to particle Reynolds and Froude numbers. From a dimensional analysis, Wen and Hashinger also included a density ratio in their correlation. The correlation of Wen and Hashinger is represented by the equation

$$\frac{K^*}{\rho_g(u_o-u_t)} = 1.52\text{x}10^{-5}\left[\frac{(u_o-u_t)^2}{gd_p}\right]^{0.5}\left[\frac{d_p\rho_g u_t}{\mu}\right]^{0.725}\left[\frac{\rho_s-\rho_g}{\rho_g}\right]^{1.15} \qquad (7)$$

and the correlation of Yagi and Aochi is approximated by

$$\frac{K^*d_p}{\mu}\cdot\frac{gd_p}{(u_o-u_t)^2} = 0.0015\left[\frac{d_p\rho_g u_t}{\mu}\right]^{0.6} + 0.01\left[\frac{d_p\rho_g u_t}{\mu}\right]^{1.2} \qquad (8)$$

as given by Leva and Wen (2).

To aid in the design of a pilot scale gasifier as well as obtain operating experience with a fluidized bed, elutriation rates were studied using the simulation unit. Elutriation constants were determined for various test materials and compared with constants estimated from the above correlations.

EXPERIMENTAL EQUIPMENT

The experimental fluidized bed simulation unit made possible the visual observation and study of bed behavior. The unit consisted of an 8-inch diameter plexiglas fluidization column with an air compressor supplying the fluidizing gas. A rotameter was used to measure the gas flow rate, and either one or two cyclones in series were used to remove solids from the exit gas stream. A screw feeder could be attached to the column for continuous feeding experiments and a solids overflow line was provided. A schematic diagram of the system is shown in Figure 2, and the components are described below.

8-Inch Fluidization Column

The fluidization column was constructed of plexiglas pipe with an eight-inch inside diameter and $\frac{1}{4}$-inch walls. The column was composed of several sections to permit easy assembly and modification and had an overall height of 12 feet above the gas distributor. The base for the column was eight-inch inside diameter steel pipe two feet high and contained the air inlet. A photograph of the air supply piping and lower part of the column set up for continuous feeding and solids removal is provided as Figure 3.

The gas distributor consisted of two perforated plates each with 88 1/8-inch holes placed on a 0.707-inch square pitch. The hole patterns on the two plates were offset to prevent solids from dropping into the base. The plates were separated by a 0.035-inch gasket. Details of the perforated plates are shown in Figure 4.

Air Compressor

Air for fluidization was supplied by a Worthington Model 20 RS 80B air compressor. The compressor was a sliding vane oil-cooled type capable of delivering 150 SCFM at 20 psig with a maximum discharge pressure of 125 psig. The compressor was powered by a 20 hp motor.

Dust Collectors

Two cyclones were used to remove dust from the exit air stream. Figure 5 shows the cyclone used for most of the experimental work. It had a four-inch inside diameter with a cylindrical top section eight inches high and a conical bottom section also eight inches high. The cone tapered from the four-inch diameter at the top to a one-inch outlet at the bottom. The air inlet was two inches inside diameter and was approximately tangential.

The second cyclone, which was placed in series with the first when used, had a four-inch inside diameter and was 16 inches high. There was no conical section and the dust outlet was again one inch. The air inlet was a tangentially arranged rectangle one inch by two inches. Air outlets for both cyclones were two inches in diameter and extended below the air inlets.

Provisions were also made to attach a bag filter to the end of the air outlet. However, the bag filter was not used for any quantitative purposes but only to control dust emissions during some of the continuous feeding experiments.

Static Mixers

Provisions were made to install Koch static mixer elements at various locations in the dense bed. The elements used were six-inch diameter by six inches high and were suspended in the center of the column. Figure 6 shows one of the mixer elements installed in the column.

Screw Feeder

For continuous feeding experiments a Vibra-Screw screw feeder was used. The feeder had a one-inch diameter screw with a variable speed drive. Feed rates for ground manure from 100 to 800 g/min could be obtained. The bin was sealed to prevent loss of feed material and had a vibrator to improve flow of material to the screw.

Rotameter

A Fischer & Porter model 10A3565A Flowrator was used for measuring air flow to the column. The meter had a maximum capacity of 123.6 standard cubic feet per minute.

Pressure Drop Measurement

Pressure taps were installed in the base of the column, in the flanges of two of the column sections, and in the top plate of the column. Tubing from the taps could be attached to either mercury or water filled manometers.

EXPERIMENTAL TEST MATERIALS AND PROCEDURES

Materials used in the fluidization column were prepared from either horticultural sand or dried feedlot manure. Sand for the bulk of the bed material was cleaned by fluidizing batches in the column for about two hours at the maximum air velocity (115 cm/sec). Sand remaining in the column was then screened to remove any particles greater than 10 mesh. Table I shows a typical sand particle size distribution based on sieve analyses. The fine sand removed from the bed during cleaning was collected in the cyclone and used either as collected or after separation into sieve fractions for test material. Solids density for the sand was taken as 2.6 g/cm^3 (5).

Additional test materials were prepared from dried feedlot manure obtained from feedlots operated by the Kansas State University Department of Animal Science and Industry. For some tests the manure was ground on a Wiley mill to pass a $\frac{1}{4}$-inch screen. For other tests the manure was heated in a reducing kiln to between 1000 and 1500°F to produce an ash and char mixture. The resulting ash product was passed through a 10 mesh screen to remove rocks and other large particles. Because of large initial attrition rates for both the manure and ash, only materials elutriated during previous runs were used for some of the tests. Table II gives typical particle size distributions for the test materials. Solids densities were found by pressing pellets of test materials under about 10,000 psi. Values obtained were 1.43 g/cm^3 for the ash and 1.16 g/cm^3 for the ground manure.

Batch experiments were run using clean sand mixed with a test material (fine sand, ground manure or ash) in known proportions. The mixture was placed in the fluidization column and air flow started. The mixture was fluidized for a short time at a low air velocity to insure that the bed was well mixed and moving freely. The air control valve was then quickly adjusted to give the desired air rate and timing started. Samples were collected from the cyclone solids discharge at appropriate time intervals for each run. Data recorded during each run included maximum and minimum bed heights and bed pressure drop. Following each run, the bed and all samples collected from the cyclone were weighed to check the material balance and determine the elutriation rate. For batch experiments using test material with a non-uniform particle size, sieve analyses were run on the elutriated material.

For continuous feeding experiments, the screw feeder was set up to feed into the column about nine inches above the distributor plates and $\frac{1}{4}$-inch from the wall. A weighed quantity of sand was placed in the column, and test material was weighed and put in the feed hopper. The air control valve was adjusted to give the desired air rate before starting the feeder. Solids overflow was not used for any of these experiments. Elutriation rates were measured during each run; and after the run weights of the bed, material left in the feeder and elutriated material were measured to check the material balance.

OPERATING CHARACTERISTICS

The fluidized bed operating conditions were held constant for all of the experimental work and only the type of material being elutriated varied. The superficial air velocity was about 97 cm/sec and the column was operated near atmospheric pressure. Air temperature varied from 80 to 130°F depending on the ambient air temperature and how long the compressor had been operating. Total bed weight for each run was approximately 20 kg.

With the given air velocity and bed size the bed operated as a slugging bed. Minimum bed height was about 60 cm and the maximum approximately 110 cm. Static height of the bed was 38 cm. During operation only one gas slug was observed in the bed at any given time. Static electricity was a problem during all of the runs although it was worse for some than for others. Grounding the column reduced the static somewhat, but it could not be completely eliminated.

Pressure drop across the distributor plates was measured with the column empty and was found to be 59 cm H_2O (4 cm Hg). Because of the slugging operation, pressure drop across the bed varied considerably and it was difficult to determine an average value. The average bed pressure drop was approximately 5 cm Hg with fluctuations from 2 to 12 cm Hg. The average bed pressure drop compared favorably with the theoretical pressure drop calculated from the weight of the bed and the cross section (4.5 cm Hg).

Koch static mixers were placed at various locations in the bed in order to observe the effect on slugging. With only one element located in the region from the middle of the bed to the expanded bed surface, the maximum bed height was significantly reduced to about 69 cm while the minimum height changed only slightly to 53 cm. The average pressure drop across the bed remained the same (5 cm Hg) but the fluctuations in pressure drop were reduced to about $\pm$ 1 cm Hg. Addition of a second mixer element did not noticeably change the quality of fluidization from that obtained with only one element.

Bed circulation was quite good both with and without static mixers. During slugging operation, solids generally moved downward near the walls and upward in the center of the column. Occasionally gas slugs would carry solids a short distance upward near the walls. The static mixers did not appear to affect the rate of circulation although no quantitative measurements were taken.

Continuous feeding of solids did not in general alter the quality of fluidization. However, when the amount of solids in the bed was increased because of feeding, the slugging became more violent. Also, when the ground manure was fed, large manure particles tended to accumulate in the top layers of the bed. Only a few qualitative trials were run with continuous feeding and withdrawal of sand. Again the quality of fluidization was unchanged. The solids discharge rate fluctuated somewhat because of the slugging, but it did not appear that would cause any significant problems for pilot plant design.

Rates for a given speed setting on the feeder were reproducible within ± 10 g/min when disconnected from the fluidization column. However, when feeding into the column, average rates obtained from material balances were 10 to 25 g/min higher than during calibration of the feeder. There were also severe problems in maintaining steady feed rates for the experiments. Pressurization of the feeder and feed hopper with air from the column inlet caused an initial surge of feed that could not be controlled by the screw speed. Finally, it was found that sealing the feed system so that it was maintained at column pressure but with no air actually flowing through the system allowed fairly steady feed rates.

EXPERIMENTAL RESULTS

The initial batch runs were made using fine sand with a wide particle size distribution as the test material. However, satisfactory material balances were not obtained for the various size fractions so most of the data was of limited value. Results from those runs were only used to estimate an elutriation constant for the largest size fraction (d_p = 0.0283 cm). Elutriation constants for smaller sizes of sand were obtained using fine sand fractions having narrow particle size distributions. Experimental conditions for all the batch runs used to obtain elutriation constants are listed in Table III. Values for K and K* are reported in Table IV. The experimental data are tabulated in Appendix D, and an example of the treatment of the data to obtain elutriation constants is given in Appendix E.

It was found from runs using ash and ground manure (runs 23 and 25) that the starting materials contained large agglomerates that broke down to elutriable sizes very rapidly upon fluidization. Approximately 20% of the ground manure initially too large to be elutriated was broken down. For the ash about 70% of the material initially greater than 170 mesh broke down to a size less than 170 mesh upon fluidization. To obtain values of K for those runs, it was assumed that the starting concentration of elutriable material was equal to the total amount actually elutriated from the bed during the run. For the manure and ash runs the data were analyzed on the basis of all material being the mean particle size for each case since the test materials could not be easily separated into narrow particle size fractions.

Four batch runs were made with Koch static mixers placed at various locations in the bed. Elutriation constants for those runs have been included in Table IV. While only limited data were taken, it appears that the elutriation rate was increased slightly by the addition of static mixers. This is in agreement with the observations of Wen and Hashinger (3).

Before the experimental data could be compared with existing correlations, it was necessary to select appropriate parameters to describe the irregularly shaped particles. Two parameters wore required, a sphericity (or shape factor) and a diameter correction factor to convert average diameters from sieve analyses to equivalent spherical diameters (the diameter of a sphere having a volume equal to the particle volume). Test materials were examined under the microscope for general shape characteristics, then values of the sphericities were estimated as 0.6 for sand and 0.4 for both manure and the ash and char mixture using data from Perry's Handbook (6). The diameter correction factor for all materials was estimated as 1.6 by averaging data presented by Brown, et al. (7) for various particle shapes except disks.

Using the estimated parameters, terminal velocities were calculated for the test materials. The procedure given by Kunii and Levenspiel (1) was used for the calculations. The gas density and viscosity were based on average experimental conditions and held constant for the calculations. Figure 7 presents the calculated terminal velocities as functions of equivalent spherical diameter for the three test materials (sand, manure and ash). Because the terminal velocity for the largest sand particles elutriated ($d_p \approx 0.028$ cm) must be slightly less than the superficial gas velocity, it was possible to obtain an indication of the validity of the estimated parameters. The terminal and superficial gas velocities were 85 and 97 cm/sec respectively thus giving reasonable agreement.

The experimental values for the specific elutriation constant K* are plotted in Figure 8 against the equivalent spherical diameters as log K* vs log d_p. The data for sand cover a wide range of particle sizes with the elutriation constant increasing as particle size decreases. Data presented by Wen and Hashinger (3) show a similar behavior. Such a trend is expected as the elutriation driving force, the difference between the superficial and terminal velocities, increases as the particles decrease in size.

The correlation for estimating K* proposed by Wen and Hashinger (3) is shown in Figure 9 along with the experimental data. The dashed lines represent approximate bounds for the data used in developing the correlation. Limits for the operating conditions under which the correlation should apply were given as

$$0.0041 \text{ cm} < d_p < 0.0147 \text{ cm}$$

$$0.00017 \text{ g/cm}^3 < \rho_g < 0.0012 \text{ g/cm}^3$$

$$1.3 \text{ g/cm}^3 < \rho_s < 5.0 \text{ g/cm}^3$$

$$22 \text{ cm/sec} < u_o < 132 \text{ cm/sec}$$

Experimental conditions were all within the limits except the gas density which averaged 0.0013 g/cm^3. Although most of the sand data fall within the dashed lines, the smallest sand (d_p = 0.0056 cm), manure and ash are well out of the bounds for the correlation.

Figure 10 shows the experimental data plotted in a similar manner on the correlation proposed by Yagi and Aochi (4). No range of operating conditions was given for the correlation. Again the smallest sand, manure and ash data fall well outside the bounds of data used in developing the correlation.

The correlations were tested by estimating values of K* for the sand diameters used in the experiments. The estimated values are plotted in Figure 11 as log K* vs log d_p. For reference the curve for the experimental data from Figure 8 is also shown. The estimated elutriation constants increase as expected in going from the largest particle diameter to the medium diameters. However, the estimated constants for the smaller particles then decrease giving considerable error.

Values of K* were estimated from both correlations for a range of spherical particle diameters. Results were similar with maximum values predicted for intermediate particle diameters. This indicates failure of both correlations in adequately predicting elutriation constants for small diameter particles. Although the failure was not examined in detail, it appears to be the result of rapidly diminishing influence of the $u_o - u_t$ term used.

Further testing of the Wen and Hashinger correlation was accomplished by using values of K* from the correlation to estimate results that could be expected from continuous feeding experiments. The estimated results were then compared to experimental data. For the experimental runs either manure or ash was fed, and the column initially contained only sand so that equation 5 applied. Table V gives pertinent data for the continuous runs. Only runs 33 and 34 were used for comparison because of the difficulty in handling the attrition that occurred during the other runs. The comparison for run 33 (manure) is given in Table VI and for run 34 (ash) in Table VII. Sample calculations are presented in Appendix F.

All of the predicted results were too low with the intermediate particle sizes (d_p = 0.012 to 0.024 cm) showing the closest agreement. Poor estimation for the larger diameter particles in run 34 was probably caused by experimental error due to the small quantities involved. The larger particles in the elutriated manure (run 33) were primarily hair fibers and fragments of leaves so that a poor estimation of the physical properties (density and shape correction factors) could have been the source of error. For the smaller particles in both cases, the unreliability of the correlation is again evident.

Both the batch and continuous feeding experiments gave consistently higher elutriation rates than were predicted from either of the correlations with the deviations becoming greater as the particle size decreased. Wen and Hashinger (3) obtained similar results in analyzing one of their batch experiments using spherical particles as shown in Table VIII.

In an attempt to improve the usefulness of the correlations for design purposes, an extrapolation procedure was devised as follows:

1. Estimate values of K* using both correlations and plot as log K* vs log d_p.
2. Draw a curve through the values for large and intermediate particle diameters and extrapolate to the smaller particle diameters.

The resulting curve should be similar to the experimental data curve shown in Figure 8 and should give an improved estimation of K* for the smaller particles.

The above procedure was applied to manure particles resulting in the curve for K* shown in Figure 12. Values from the curve were then used to estimate results for run 33 which are compared to the experimental values in Table IX. Although the overall results show improvement, there is still considerable error in predicting behavior of the large particles. As mentioned earlier, that is probably due to the physical nature of the large particles.

A similar analysis was made for the ash data from run 34. The curve used for estimating K* is shown in Figure 13 and results of the analysis are given in Table X. In this case, fairly good agreement with the experimental data was obtained.

Finally, an estimate of the amount of ash build-up for a pilot plant system processing one-half ton per day of dry manure was made using the extrapolation procedure. Figure 14 shows the pilot plant system based on the conceptual design of Chapter II. It is composed of pyrolysis and combustion reactors with sand circulating between them as a heat transfer medium. Quantities determined were the amounts of ash and char elutriated from the pyrolysis reactor and the amount of ash circulating with the sand at steady state. For the calculations it was assumed that the ash and char would be formed instantaneously in the pyrolysis reactor and would all be of an elutriable size as represented by the elutriated ash analysis given in Table II. It was also assumed that agglomeration would not occur and that all the char would be burned instantaneously in the combustion reactor.

The calculations (given in Appendix G) indicate that at steady state less than 5% of the circulating solids should be ash. The load on the pyrolysis reactor cyclone was estimated as 50 g/min or 11 grains/ft^3. Since the extrapolation procedure tends to underestimate elutriation rates, the calculated amount of ash build-up should be conservatively large. However, the estimated load on the cyclone is probably somewhat small.

The average specific elutriation constant estimated for the pilot plant was 0.052 g/cm^2 sec compared with an experimental value of 0.81 g/cm^2·sec for the simulation unit. The lower estimated value is reasonable since the pyrolysis gas density is much less than that of air thus providing a lesser buoyant force for the particles. Partially offsetting the effect of density is a slightly higher estimated viscosity for the pyrolysis gas than for air.

SUMMARY

The fluidized bed simulation unit was used to observe bed behavior under operating conditions approximating those estimated for the manure pyrolysis process. With the selected conditions the bed operated as a slugging bed. However, the addition of Koch static mixers in the upper portion of the dense bed effectively broke up the gas slug and significantly reduced bed fluctuations.

In addition to observing bed behavior, elutriation of fine sand, ground manure and ash was studied to obtain information for pilot plant design. Experimental values for the specific elutriation constant K^* determined from batch runs were compared to two existing correlations and found to give only limited agreement. However, both correlations seriously underestimated K^* for smaller diameter particles as the values predicted were lower than values for slightly larger particles. By using the correlations to estimate K^* for ranges of particle diameters, it was possible to extrapolate the data for larger diameters to obtain curves of log K^* vs log d_p for different types of particles. Predicted results for continuous feeding of both manure and ash using values of K^* from the curves gave fair agreement with experimental results although consistently underestimating the elutriation rate.

The same method of using the correlations to obtain a curve for log K^* vs log d_p was used to estimate elutriation constants for the manure pyrolysis pilot plant. Calculations based on the resulting elutriation constants indicated that less than 5% of the circulating solids should be ash. From the estimation it appears that build-up of ash in the circulating sand should not be a problem unless particles too large to be elutriated are present either by introduction with the manure feed as gravel or similar solids or by agglomeration of ash during pyrolysis or combustion.

NOMENCLATURE

A	bed cross-sectional area, cm^2
d_p	particle diameter, cm
$\bar{d}_p$	mean particle diameter, cm
$E(d_p)$	total elutriated of particle size d_p, g
F_o	total feed rate, g/min
$F_o(d_p)$	feed rate of particle size d_p, g/min
$F_2(d_p)$	elutriation rate for particle size d_p, g/min
g	acceleration of gravity, 980 cm/sec^2
K	elutriation constant, min^{-1}
K^*	specific elutriation constant, $g/cm^2.sec$
Re_p	particle Reynolds number = $\frac{d_p U_t \rho_g}{\mu}$, dimensionless
t	time, sec or min
U_o	superficial velocity, cm/sec
U_t	terminal velocity of particle, cm/sec
W	total weight of solids in bed, g
$W(d_p)$	weight of particle size d_p in bed, g
$W_o(d_p)$	initial weight of particle size d_p in bed, g
μ	gas viscosity, g/cm.sec
ρ_g	density of gas, g/cm^3
ρ_s	density of solids, g/cm^3
ϕ_s	particle sphericity, dimensionless

REFERENCES

1. Kunii, D. and O. Levenspiel, Fluidization Engineering, John Wiley and Sons, New York, 1969, p. 312ff.

2. Leva, M. and C.Y. Wen, "Elutriation," in Fluidization, edited by J.F. Davidson and D. Harrison, Academic Press, New York, 1971, p. 627ff.

3. Wen, C.Y. and R.F. Hashinger, "Elutriation of Solid Particles from a Dense-Phase Fluidized Bed," A.I.Ch.E. Journal, 6, p. 220.

4. Yagi, S. and T. Aochi in Fluidization Engineering by D. Kunii and O. Levenspiel, John Wiley and Sons, New York, 1969, p. 315.

5. Handbook of Chemistry and Physics, 48th edition, The Chemical Rubber Company, Cleveland, Ohio, 1967, p. B278ff.

6. Perry's Chemical Engineers' Handbook, Fourth ed., McGraw-Hill, New York, 1963, p. 5-50.

7. Brown, G.G., et al., Unit Operations, John Wiley and Sons, New York, 1950, p. 77.

Table I

Clean Sand Particle Size Distribution

$\bar{d}_p$ (mm)	wt. %
1.410	1.1
1.005	7.0
0.715	20.1
0.505	21.3
0.358	30.8
0.252	16.3
<0.208	3.4

$\bar{d}_p = 0.400$ mm

Table II

Manure and Ash

Particle Size Distributions (Wt. %'s)

$\bar{d}_p$ (mm)	Manure		Ash	
	Ground	Elutriated	Screened	Elutriated
3.490	8	-	-	-
1.689	15	-	-	-
1.524	-	-	3	-
1.194	14	-	5	-
0.743	23	-	2	-
0.423	10	-	9	-
0.299	8	25	9	1
0.235	6	17	12	3
0.150	4	11	16	6
0.106	2	6	18	7
0.075	2	6	12	13
0.052	3	10	10	64
<0.043	5	25	4	6

Table III

Batch Experimental Conditions

Run No.	Test Material	Mean Particle Size(cm)	Bed Weight(g)	% Test Material
5	Sand	(1)	19,506	6.8
10		0.0126	20,000	10.0
12		0.0083	20,000	6.1
13		0.0052	20,000	6.8
14		0.0052	20,000	5.6
15		0.0083	20,000	6.1
16(2)		0.0083	20,003	6.0
18(2)		0.0052	20,000	4.9
19(3)		0.0083	20,000	5.9
20(4)		0.0083	20,000	6.0
21		0.0035	20,000	5.0
23	Ash and Char	(5)	19,832	6.6
24		(6)	20,000	6.9
25	Manure	(7)	19,954	4.3
26		(8)	20,000	3.6

(1) Wide particle size distribution used to estimate 0.117 mm faction

(2) Koch static mixer 12" from distributor.

(3) Koch static mixer 6" from distributor.

(4) Koch static mixers 6" and 18" from distributor.

(5) Wide particle size distribution including material >10 mesh.

(6) Particle size distribution given in Table II for elutriated ash and char.

(7) Particle size distribution given in Table II for ground manure.

(8) Particle size distribution given in Table II for elutriated manure.

Table IV

Elutriation Constants

Material	Run No.	$\bar{d}_p$ (cm)	K(min^{-1})	K*($g/cm^2.sec$)
Sand	5	0.0283	0.0232	0.0233
	10	0.0202	0.0449	0.0462
	12	0.0133	0.213	0.219
	15	0.0133	0.165	0.170
	13	0.0083	0.334	0.344
	14	0.0083	0.334	0.344
	21	0.0056	0.954	0.981
Ash and Char	23	0.0093(1)	0.712	0.727
	24	0.0093(1)	0.864	0.889
Manure	25	0.0115(1)	0.589	0.605
	26	0.0115(1)	0.447	0.460
Sand(2)	16	0.0133	0.318	0.327
	19	0.0133	0.252	0.259
	20	0.0133	0.197	0.203
	18	0.0083	0.575	0.592

(1) $\bar{d}_p$ taken as mean particle diameter of elutriated material.

(2) Runs with Koch static mixers.

Table V

Continuous Feeding Experiments

Run No.	Feed[1]	Initial Bed Wt. (g)	F_o (g/min)	t (min)	Elutriated (g)	Final Bed Wt. (g)
27	Ground manure	20,000	363	11	1,021	22,972
29	Ground manure	20,000	247	12	894	22,029
33	Elutriated manure	20,000	164	10	1,328	20,310
34	Elutriated ash	20,000	128	11	1,378	20,035

(1) Typical particle size analyses given in Table II.

Table VI

Estimated and Actual Results for Run 33

$\bar{d}_p$ (cm)	K^* ($g/cm^2 \cdot sec$)[a]	F_o (g/min)	$W(d_p)$ (g)	Estimated $E(d_p)$ (g)	Actual $E(d_p)$ (g)	Estimated as % of Actual
.0478	.045	18	146	34	145	23
.0376	.093	26	171	89	214	42
.0240	.141	25	136	114	198	58
.0170	.133	15	84	66	125	53
.0120	.105	12	75	45	100	45
.0083	.073	23	165	65	182	36
.0048	.037	45	378	72	366	20
		164	1155	485	1330	36

[a]Estimated using correlation of Wen and Hashinger (3).

Table VII

Estimated and Actual Results for Run 34

$\bar{d}_p$ (cm)	K^* ($g/cm^2 \cdot sec$)[a]	F_o (g/min)	$W(d_p)$ (g)	Estimated $E(d_p)$ (g)	Actual $E(d_p)$ (g)	Estimated as % of Actual
.0478	.011	1	10	1	4	25
.0376	.069	2	16	6	24	25
.0240	.165	9	47	52	88	59
.0170	.172	16	81	95	155	61
.0120	.143	32	180	172	310	55
.0083	.103	60	400	260	584	44
.0048	.054	8	67	21	74	28
		128	801	607	1239	49

[a] Estimated using correlation of Wen and Hashinger (3).

Table VIII

Estimated and Actual Results for Spherical Particles[a]

$\bar{d}_p$ (cm)	Amount Elutriated (g) Estimated	Actual	Estimated as % of Actual
.0147	153	148	103
.0099	298	291	102
.0071	320	414	77

[a]Wen and Hashinger (3)

Table IX

Estimation for Run 33 Using Extrapolation Procedure

$\bar{d}_p$ (cm)	K* $(g/cm^2 \cdot sec)$[a]	$W(d_p)$ (g)	Estimated $E(d_p)$ (g)	Actual $E(d_p)$ (g)	Estimated as % of Actual
.0478	.043	147	33	145	23
.0376	.059	198	62	214	29
.0240	.123	146	104	198	53
.0170	.173	73	77	125	62
.0120	.257	44	76	100	76
.0083	.390	59	171	182	94
.0048	.720	64	386	366	105
		731	909	1330	68

[a]From Figure 12.

Table X

Estimation for Run 34 Using Extrapolation Procedure

$\bar{d}_p$ (cm)	K* (g/cm^2-sec)[a]	$W(d_p)$ (g)	Estimated $E(d_p)$ (g)	Actual $E(d_p)$ (g)	Estimated as % of Actual
.0478	.0105	10	1	4	25
.0376	.0297	19	3	24	13
.0240	.105	60	39	88	44
.0170	.187	76	100	155	65
.0120	.310	102	250	310	81
.0083	.500	123	537	584	92
.0048	.970	8	80	74	108
		398	1010	1239	82

[a]From Figure 13.

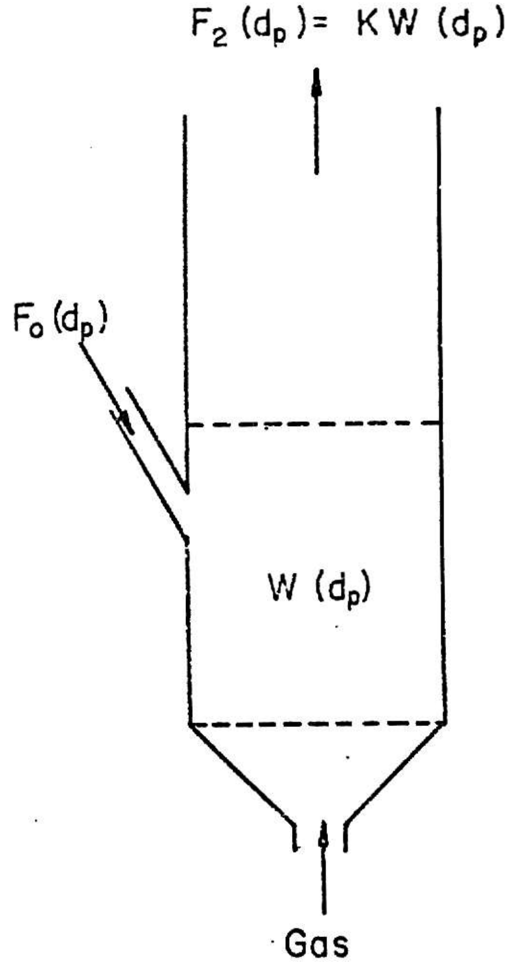


Fig. 1. Steady - State Elutriation of Particle Size d_p.

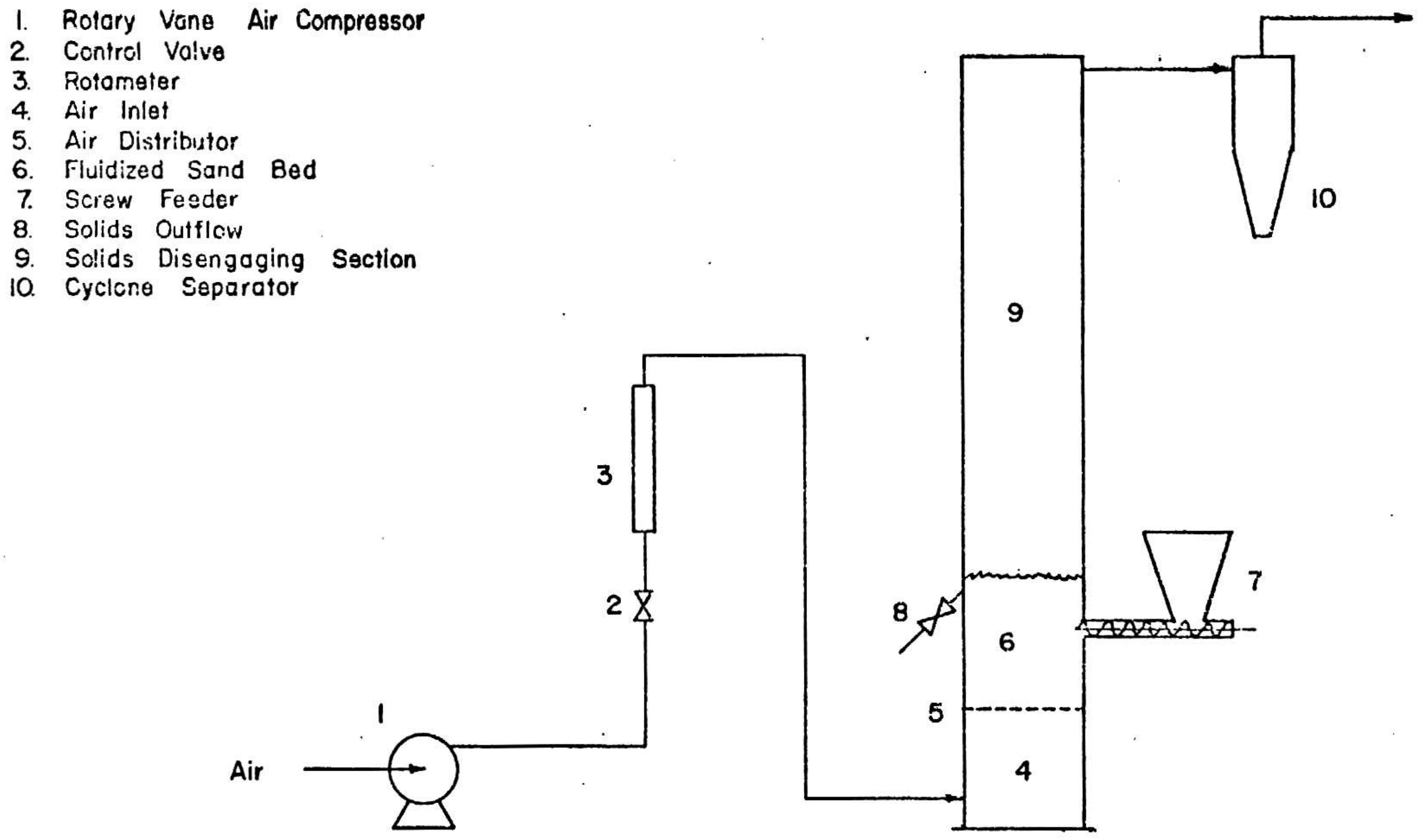


Fig. 2. Fluidized Bed Simulation Unit.

THIS BOOK CONTAINS SEVERAL DOCUMENTS THAT ARE OF POOR QUALITY DUE TO BEING A PHOTOCOPY OF A PHOTO.

THIS IS AS RECEIVED FROM CUSTOMER.

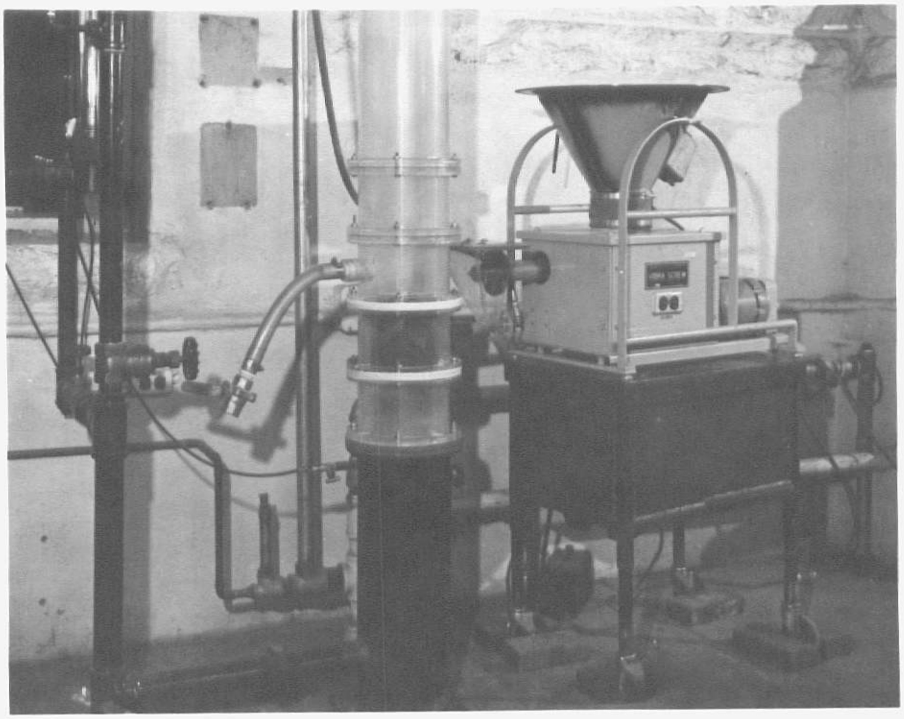


Fig. 3. Bottom Section of Simulation Unit

The above view shows the simulation unit set up for continuous feeding and solids removal. Air from the compressor passes through a rotameter and the piping at the left then enters the air inlet section (solid base of the column). The air distributor plates are located between the base and plexiglas column. A Koch static mixer is shown installed in the second column section. The screw feeder on the right is set up with an experimental pneumatic feed injection system.

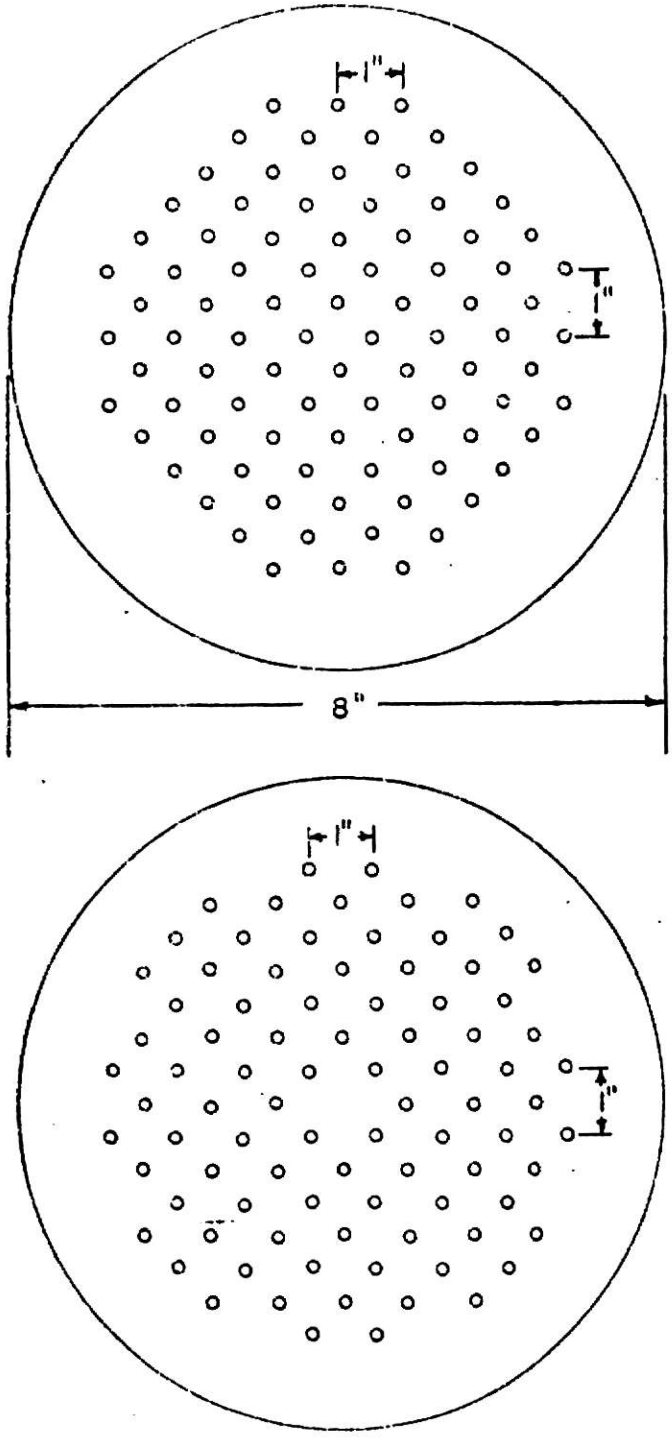


Fig. 4. Distributor Plate Details.

Fig. 5. Solids Disengaging Section

This photograph shows the upper portion of the fluidization column. The cyclone separator is mounted on the wall. The piping from the cyclone leads to a bag filter and outside vent.

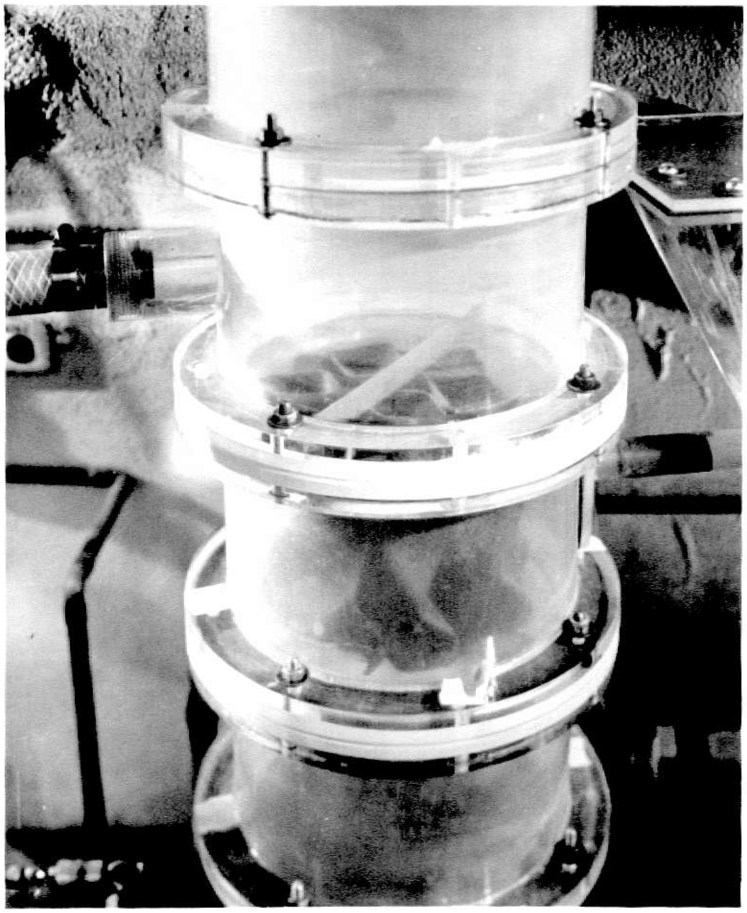


Fig. 6. Static Mixer Installation

Above is a closeup view of a six-inch Koch static mixer installed in the column.

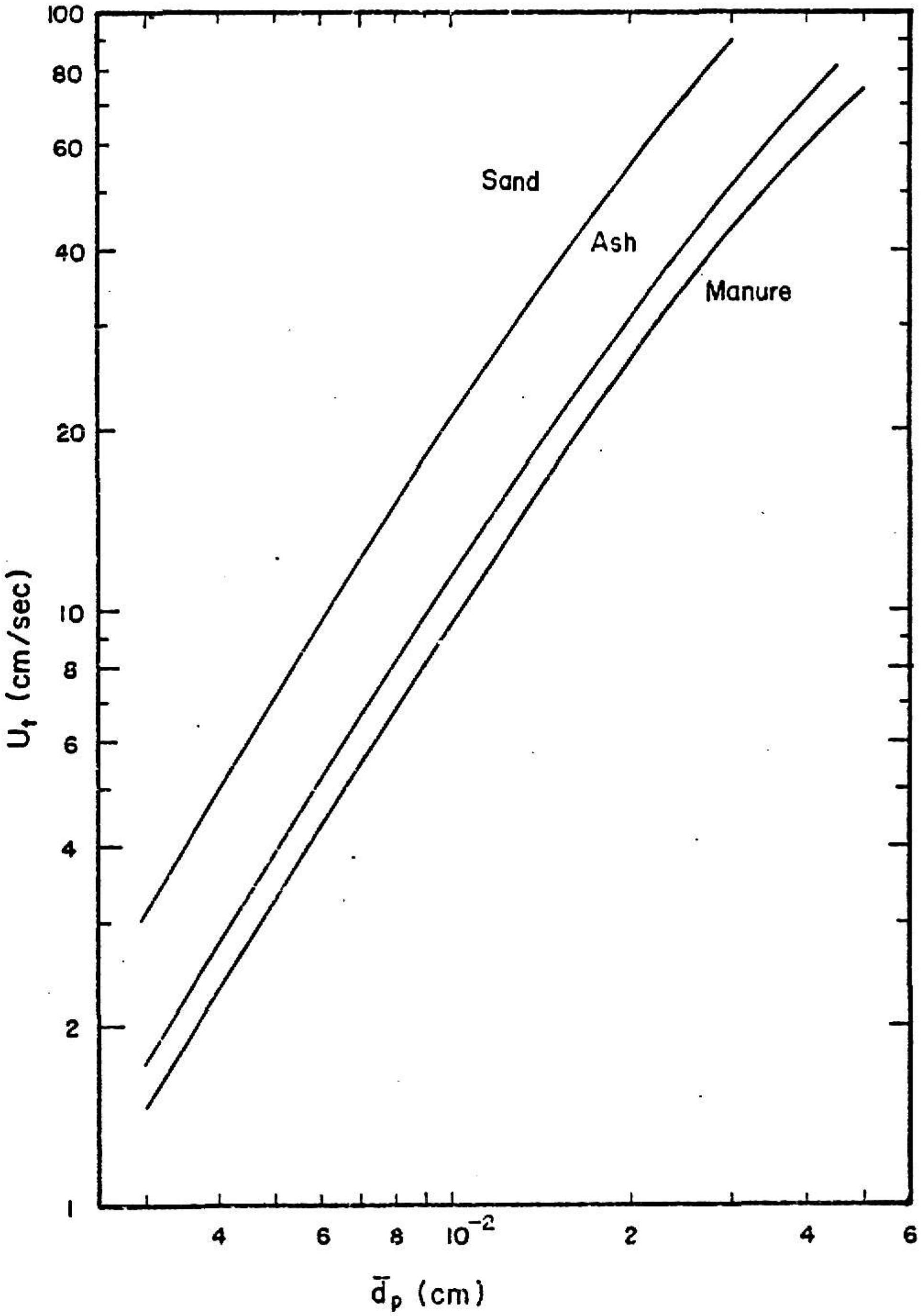


Fig. 7. Terminal Velocities for Experimental Conditions.

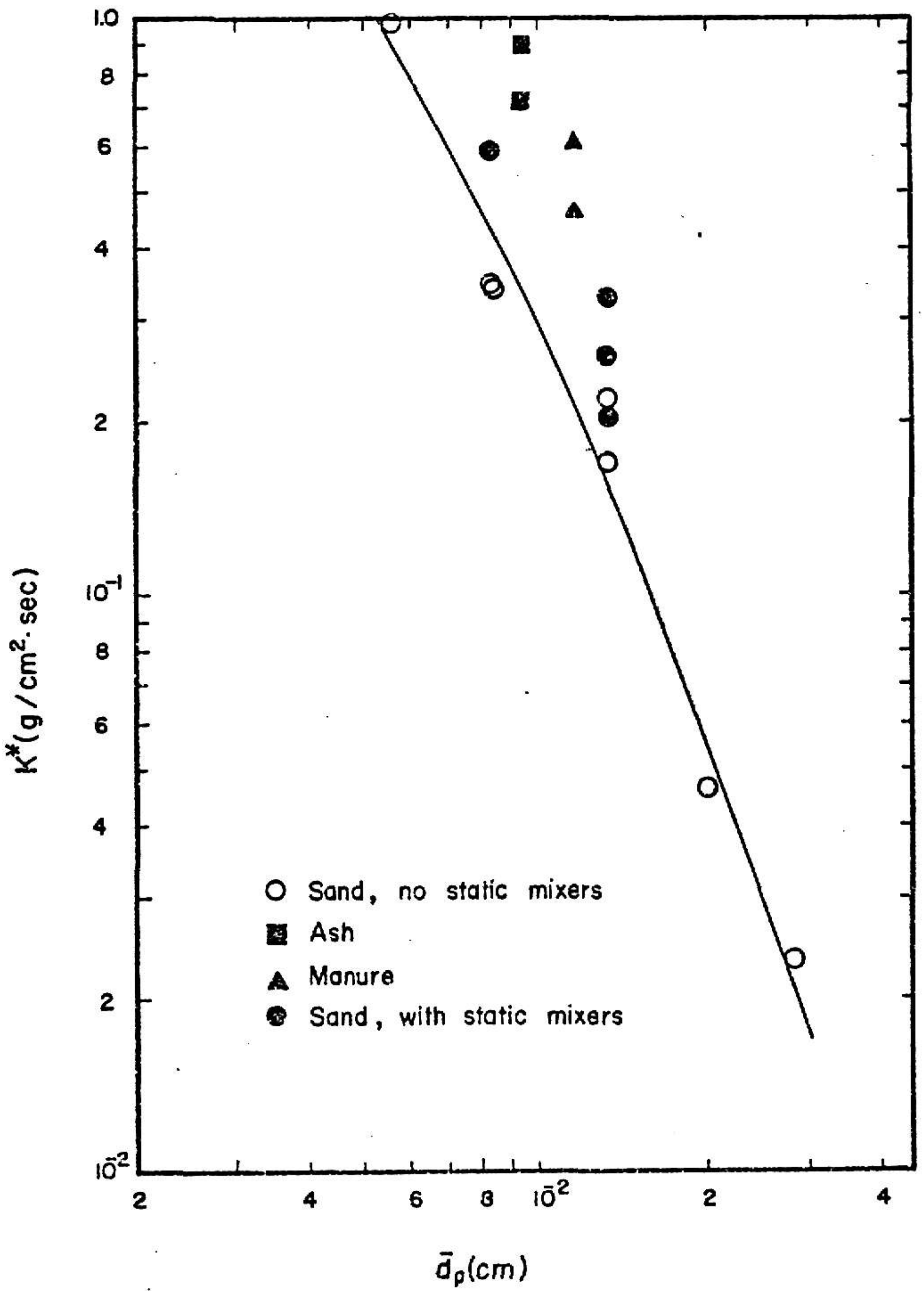


Fig. 8. Experimental Elutriation Constants.

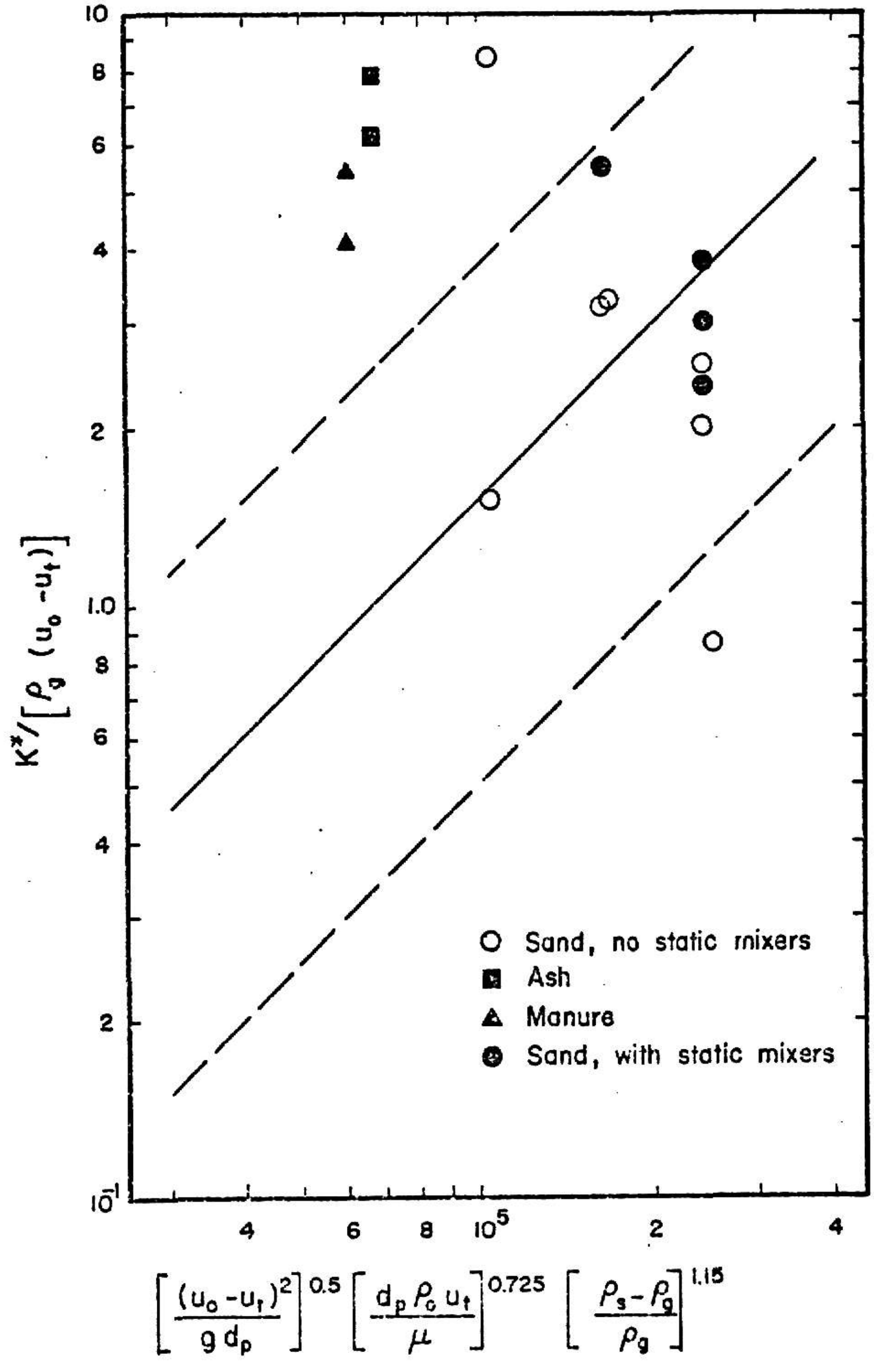


Fig. 9. Correlation of Wen & Hasinger [3].

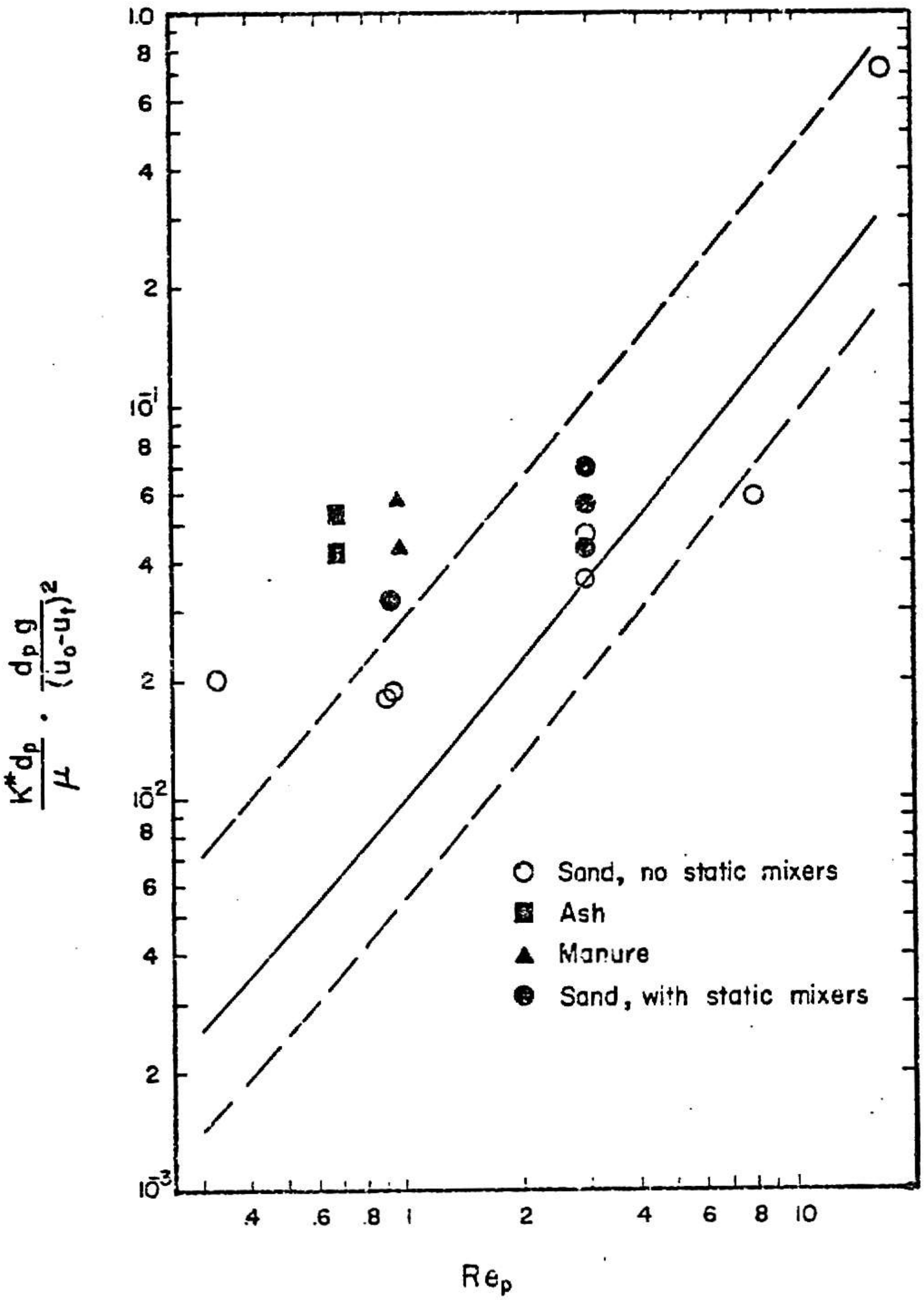


Fig. 10. Correlation of Yagi & Aochi [4].

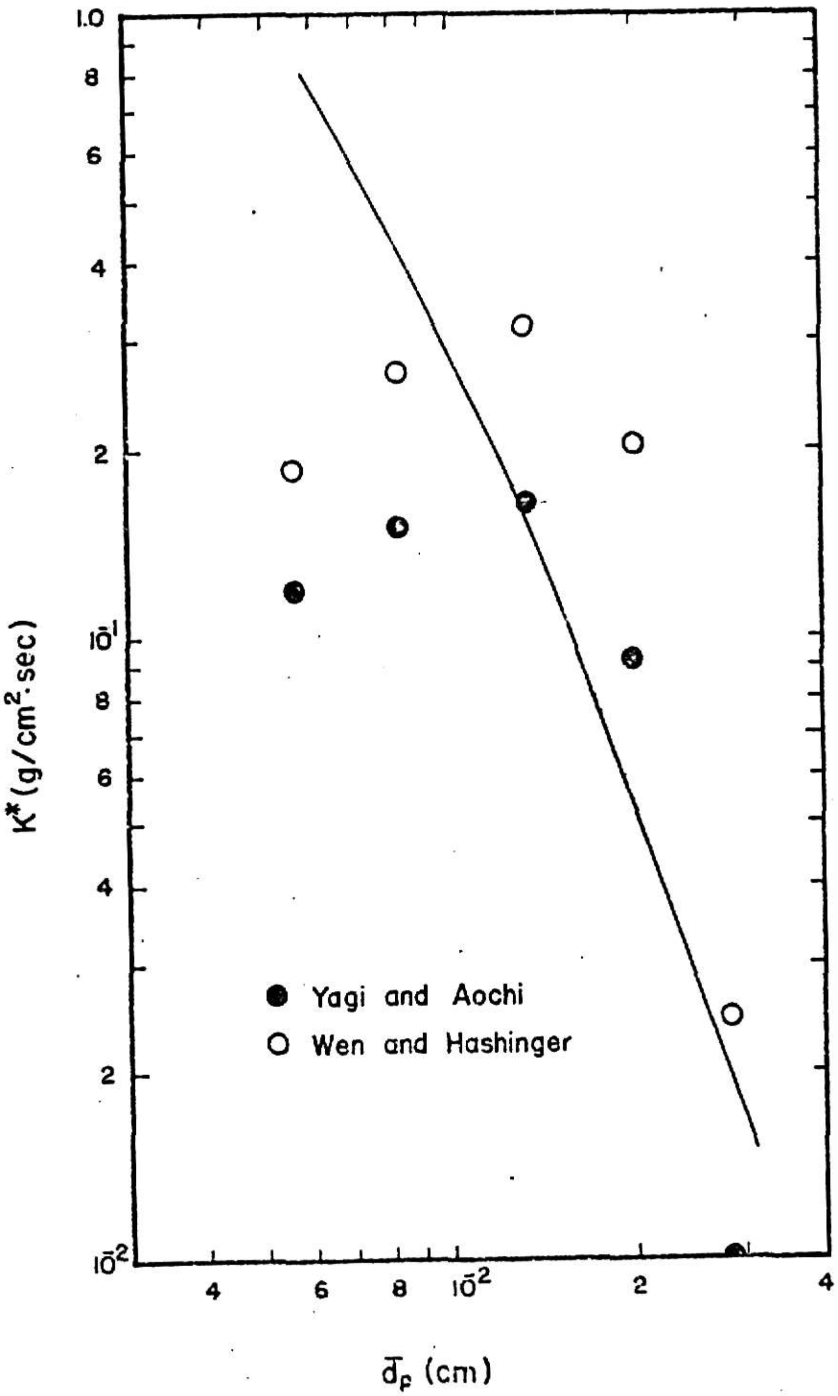


Fig. 11. Estimated Elutriation Constants for Sand.

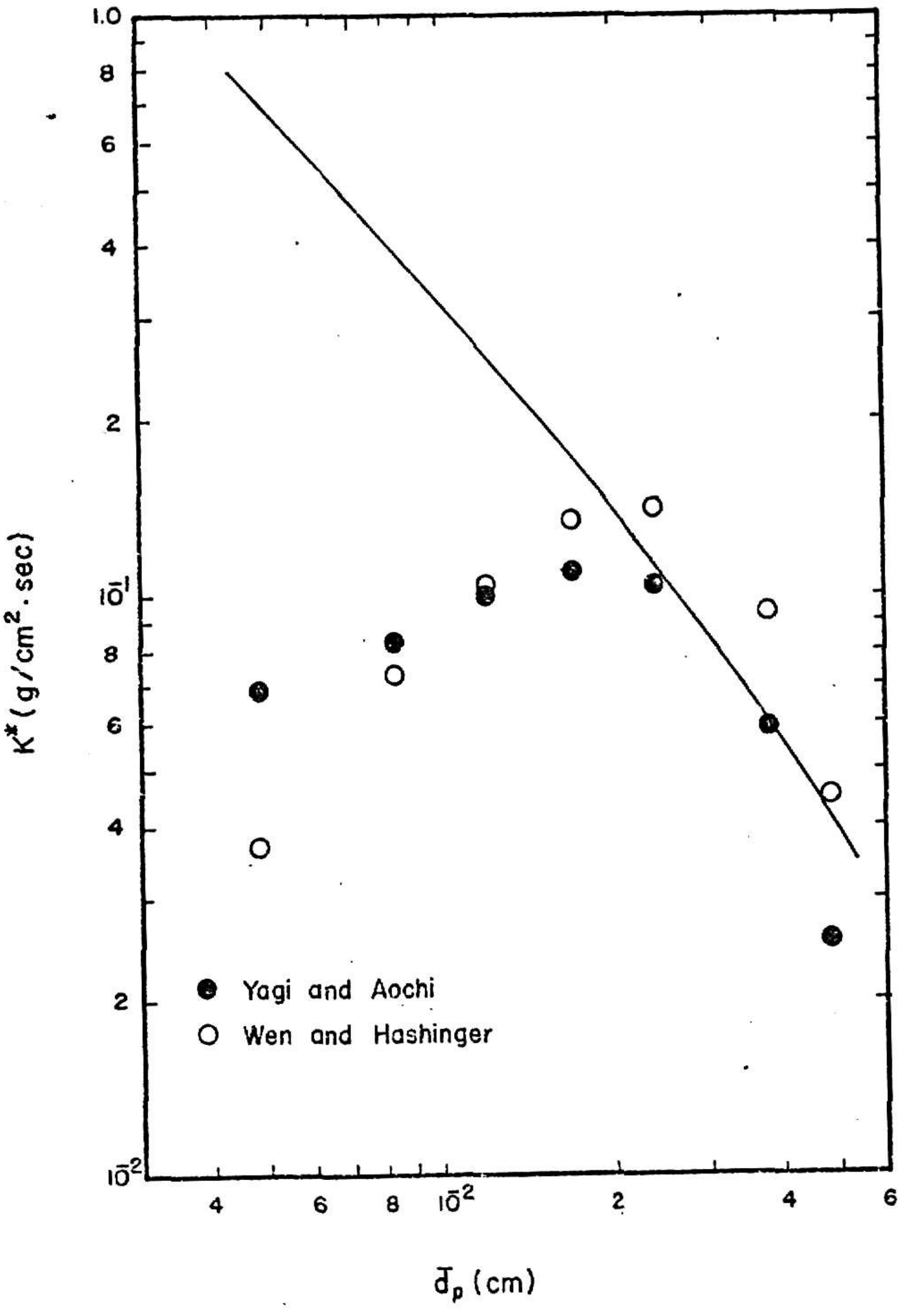


Fig. 12. Estimated Elutriation Constants for Manure.

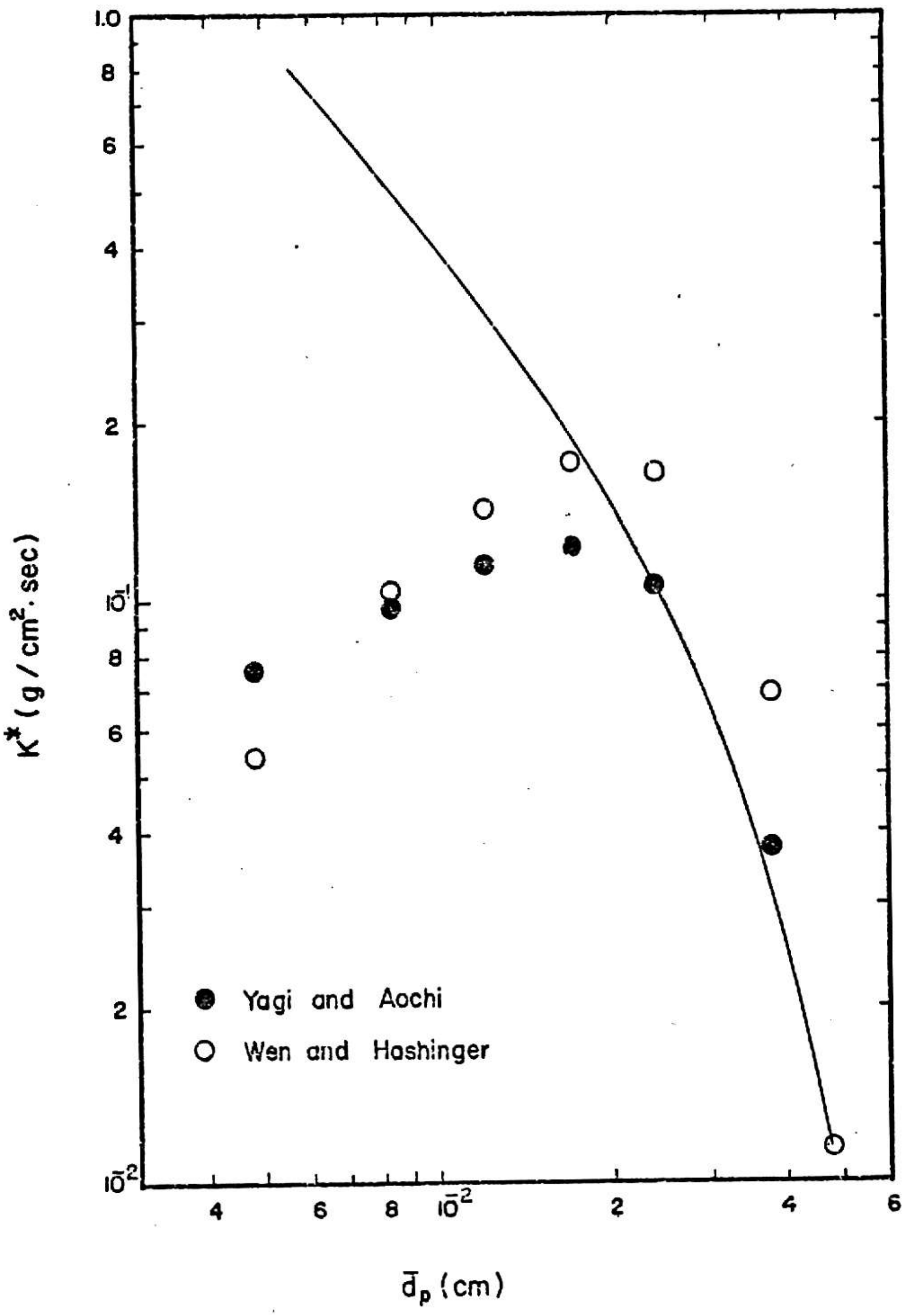


Fig. 13. Estimated Elutriation Constants for Ash.

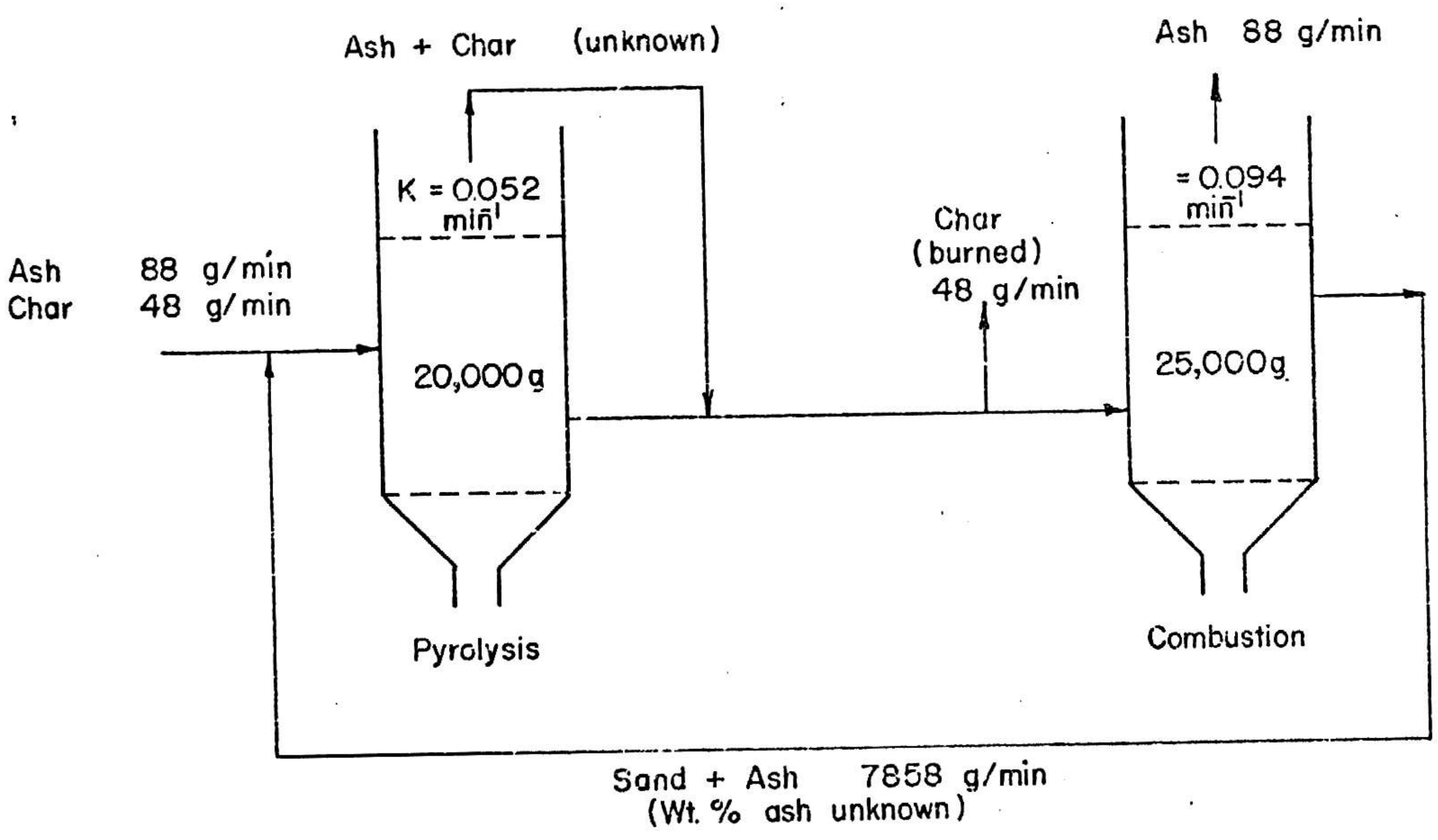


Fig. 14. Pilot Plant System for Elutriation Estimate.

CHAPTER V
CONCLUDING REMARKS

CONCLUSIONS

A conceptual design for a plant to produce synthesis gas from 500 T/D (tons per day) of dry feedlot manure via fluidized bed pyrolysis was developed and used as the basis for an economic analysis. The analysis indicated that a plant of that capacity would not be profitable in the present economic environment but that larger capacities (4,000 T/D of dry manure) could be competitive. The analysis also showed that moisture content of the incoming manure and the cost of transporting manure from the feedlot to the plant significantly affected profitability. With the cost of producing synthesis gas increasing due to increased costs and shortages of natural gas, the process could be profitable in the near future for plants as small as 500 T/D capacity.

Based on a survey of feedlot capacities in southwestern Kansas, the application of manure pyrolysis in that geographical area offers considerable potential. Plants processing in the range of 4,000 T/D of dry manure could be supported although the manure would have to be transported as far as 100 miles. Smaller plants processing around 2,000 T/D were found to be equally as profitable as the large plant because of reduced transportation requirements.

Ammonia production and electricity generation were also studied as processes that could use the synthesis gas produced. It was found that conversion of manure to ammonia via a synthesis gas intermediate increased its equivalent nitrogen value by nearly 17 times. In addition, other plant nutrients would be available in the ash byproduct of pyrolysis. A pyrolysis plant processing 2,000 T/D could support ammonia production of about 150,000 T/yr or an estimated 70% of the nitrogen fertilizer requirements for the southwestern quarter of Kansas. Using the synthesis gas for electricity generation rather than ammonia production, a 2,000 T/D pyrolysis plant could support a generating facility rated at about 150 megawatts.

Because of the favorable potential for manure pyrolysis, a fluidized bed simulation unit was constructed to observe bed behavior under flow characteristics approximating the conceptual design conditions. The bed operated as a slugging bed under the selected conditions. However, Koch static mixers added to the bed effectively broke up the gas slug without significantly affecting bed circulation.

The simulation unit was also used to obtain partial design information for construction of a pilot scale gasifier. Elutriation rate constants were determined for fine sand, manure, and ash and compared to two existing correlations. Examination of the correlations over ranges of particle diameters showed that both failed to adequately predict elutriation constants for the smaller particle diameters. However, it was possible to extrapolate estimated data for the larger diameters to give improved agreement between experimental and estimated values for smaller diameters.

RECOMMENDATIONS

Optimization of the conceptual process design was not done in this work. Considering the large amount of capital required for a manure pyrolysis plant, significant savings might be possible through the use of different types of equipment. Two areas of the conceptual design in particular that should be optimized are the feed preparation and drying section and the gas clean-up section. In addition to optimizing the equipment selection, the effect of moisture content of the pyrolysis reactor feed on synthesis gas yield should be studied either in the laboratory or as part of the pilot scale gasifier studies.

Because the cattle feeding industry is very sensitive to economic fluctuations, further effort is needed to define plant sizes that could realistically be supported in southwestern Kansas. Data concerning the magnitude of changes in the number of cattle actually on feed in the area should be obtained along with information about manure disposal practices currently in use and anticipated for the future. The availability of crop wastes and other organic wastes that could be used as supplementary feeds for a pyrolysis plant should also be studied.

Additional experimentation using the fluidized bed simulation unit that would be of help in designing a pilot scale gasifier includes an extension of the study of static mixer effects on bed operation, testing of other feed systems for continuous feeding and study of circulating systems using a second fluidization column. The effects of static mixers on circulation of solids within the bed should be investigated. Also, other sizes and types of mixer elements should be tried.

Because of the difficulties encountered in maintaining steady feed rates with the screw feeder, other feeding systems should be considered. For example, a screw or weigh belt feeder operating at atmospheric pressure could be used to meter the feed to a rotary air lock which would then inject the feed into a pneumatic transport system. Such an arrangement should effectively move the solids in the adverse pressure gradient encountered.

Maintaining the sand circulation rate estimated from the heat of pyrolysis requirements may be difficult due to the large quantities involved. In particular, the slugging operation of the bed may make steady circulation rates impossible. Addition of a second column to the simulation unit would allow sand circulation between beds to be studied.

Finally, further examination of the correlations available for estimating elutriation constants would be of benefit for fluidized bed design in general. Re-evaluation of the velocity terms used in the correlations could lead to improved elutriation constant estimates for small particle diameters.

ACKNOWLEDGEMENT

The author wishes to express his sincere appreciation to his advisor, Dr. W. P. Walawender, and the other members of his advisory committee, Dr. L. T. Fan and Dr. L. E. Erickson, for their support.

The work was supported in part by the Kansas Agricultural Experiment Station, project Ch.E. 0880.

APPENDIX A

MATERIAL AND ENERGY BALANCE CALCULATIONS

To develop the material balance for the pyrolysis reaction, manure composition data reported by Appel et al. (1) (see Table I), were used along with gas composition data reported by West Virginia University (2) (Table II). The following assumptions were made to obtain the material balance:

1. Ratios of the components CO_2, CO, H_2 and CH_4 in the pyrolysis gas would be 1 : 2.15 : 2.28 : 0.68 from data reported by West Virginia University (2 and Table I).
2. All nitrogen and sulfur in the manure would be converted to NH_3 and H_2S, respectively.
3. All oxygen in the manure would be converted to either CO_2 or CO.
4. Feed to the pyrolysis reactor would contain 10% water.
5. Any carbon and hydrogen remaining after meeting the above requirements would become the char product.

The dry, ash-free composition of manure was used to calculate a pseudo-molecular formula of $C_{4.35}H_{6.36}O_{2.31}N_{0.24}S_{0.04}$ which conveniently has a molecular weight of 100. The stoichiometry for the pyrolysis reaction was determined by the following steps:

1. Oxygen balance to give CO and CO_2 yield as

   $$\text{moles CO} = 2.31\left(\frac{35.5}{2(16.3) + 35.5}\right) = 1.20$$

   $$\text{moles } CO_2 = 2.31\left(\frac{16.3}{2(16.3) + 35.5}\right) = 0.55$$

2. Calculations of H_2 and CH_4 yields from ratio to CO as

   $$\text{moles } H_2 = 1.20\left(\frac{37.1}{35.5}\right) = 1.26$$

   $$\text{moles } CH_4 = 1.20\left(\frac{11.1}{35.5}\right) = 0.38$$

3. Nitrogen and sulfur balances to give

   moles NH_3 = 0.24

   moles H_2S = 0.04

4. Calculation of char composition from carbon and hydrogen balances to give

moles C = 4.35 - (1.20 + 0.55 + 0.38) = 2.22

moles H = 6.36 - 2(1.26) - 4(0.38) - 3(0.24) - 2(0.04) = 1.52

for a char formula of $C_{2.22}H_{1.52}$. The overall pyrolysis reaction can then be written as

$$C_{4.35}H_{6.36}O_{2.31}N_{0.24}S_{0.04} \rightarrow 1.20\ CO + 0.55\ CO_2 + 1.26\ H_2 + 0.38\ CH_4 + 0.24\ NH_3 + 0.04\ H_2S + C_{2.22}H_{1.52}$$

Since the incoming manure moisture content was set at 10%, it was further assumed that steam would react with 25% of the char yielding CO and H_2. The reaction would be as follows:

$$C_{2.22}H_{1.52} + 0.56\ H_2O \rightarrow 0.56\ CO + 0.75\ H_2 + 0.75\ C_{2.22}H_{1.52}$$

The total reaction for manure pyrolysis would then be

$$C_{4.35}H_{6.36}O_{2.31}N_{0.24}S_{0.04} + 0.56\ H_2O \rightarrow 1.76\ CO + 0.55\ CO_2 + 2.01\ H_2 + 0.38\ CH_4 + 0.24\ NH_3 + 0.04\ H_2S + 0.75\ C_{2.22}H_{1.52}$$

Heat of combustion data were used to calculate the heat of reaction for pyrolysis. Heats of combustion for manure and char were estimated using the Dulong formula (3) which is

$$\Delta H_c = 14{,}544\ (\text{wt \% C}) + 62{,}028\ \left(\text{wt \% H} - \frac{\text{wt \% O}}{8}\right) + 4050\ (\text{wt \% S}).$$

The heat capacity for manure was estimated from the heat capacity for cellulose given by Stamm (4) which can be expressed as

$$C_p = 0.266 + 0.00116\ T \qquad (T \text{ in } ^\circ C).$$

Char can be represented as coal with zero volatile content for calculating heat capacity using the equation (5)

$$C_p = 0.17 + 0.00011\ T \qquad (T \text{ in } ^\circ F).$$

Heat capacities for ash and sand can be obtained from the equation

$$C_p = 0.18 + 0.00006\ T \qquad (T \text{ in } ^\circ F)$$

given in Perry's Handbook (6). Table III gives the calculated values for ΔH_c and $\bar{C}_p$ along with values for other compounds used to calculate the heats of reaction. Tables IV and V show the calculations for the heats of reaction for pyrolysis and the steam-char reaction respectively. The overall heat of reaction for pyrolysis can be obtained by adding the two reactions and was found to be 828 Btu/lb DAFM at 1500°F.

The calculated heat of reaction for the pyrolysis reaction was considerably higher than values calculated from experimental data by Burton (9) for municipal solid waste. He reported a value 423 Btu/lb dry ash-free MSW as the heat of pyrolysis at 77°F (compared to 993 Btu/lb DAFM). However, in his calculation procedure he assumed that heat losses from the experimental fluidized bed were constant in a temperature range from 1435°F to 1840°F. Based on his data, a reduction in heat losses at the lower temperature by less than 10% would account for the discrepancy.

The total heat load for the reactor was found by adding the heat required to raise the temperature of the feed to 1500°F to the heat of reaction at 1500°F. Table VI shows the feed heating requirements for feed entering the reactor at 160°F. The total heat load for the pyrolysis reactor was found to be 2,348 Btu/lb DAFM.

The heat available from char combustion at 1750°F was obtained in a similar manner. The calculations are shown in Table VII. The heat of combustion at 1750°F was found to be 3,444 Btu/lb DAFM assuming all the char would be burned.

In making the overall process energy balance, the following assumptions were made:

1. Pyrolysis gas would be cooled from 1500 to 600°F by preheating combustion air.
2. Enough excess air would be used to maintain the

combustion temperature at 1750 °F.

3. Temperature of the dried manure feed would be 160 °F.

4. Heat losses would be 2% for both pyrolysis and combustion.

Table VIII shows the overall energy balance calculations. The air rate required to maintain the combustion temperature at 1750 °F was found to be 3.097 lb/lb DAFM or about 17% in excess of the stoichiometric requirement.

A final balance was made to determine the feasibility of utilizing the combustion gas for drying the incoming feed. The combustion gas moisture content would be about 5 wt %. From Combustion Engineering (10) the wet bulb temperature was estimated to be 175 °F. The dryer outlet moisture content would have to be approximately 0.40 lb H_2O/lb dry gas giving an outlet gas temperature of 300 °F. Such an operation would be feasible.

The sand circulation rate can be found from the reactor heat load assuming the sand is cooled from 1750 °F to 1500 °F in the reactor.

$$\Delta H_{sand} = \int_{1500}^{1750} (0.18 + 0.00006\ T)\ dT = 69.4 \text{ Btu/lb sand}$$

$$\text{Sand Rate} = \frac{1.02\ (2348) \text{ Btu/lb DAFM}}{69.4 \text{ Btu/lb sand}}$$

$$= 34.5 \text{ lb sand/lb DAFM}$$

REFERENCES

1. Appel, H.R., Y.C. Fu, S. Friedman, P.M. Yavorsky, and I. Wender, "Converting Organic Wastes to Oil," Agricultural Engineering, 53, 3, p. 17, 1972.

2. "Solid Waste: A New Natural Resource," (Staff of the Fluidized Bed Gasification Project), Dept. of Chemical Engineering, West Virginia University, Morgantown, W. Va., 1971.

3. Himmelblau, D.M., Basic Principles and Calculations in Chemical Engineering, 2nd Ed., Prentice-Hall Inc., Englewood Cliffs, N.J., 1967, p. 286.

4. Stamm, Alfred J., Wood and Cellulose Science, The Ronald Press Co., New York, 1964, p. 284.

5. "Estimation of Coal and Gas Properties for Gasification Design Calculations," Office of Coal Research, U.S. Dept. of Interior, Research and Development report no. 22, 1971, p. 46.

6. Perry's Chemical Engineers' Handbook, Fourth Ed., McGraw-Hill, New York, 1963, p. 3-217.

7. Himmelblau, op. cit., p. 449ff.

8. Hougen, O.A., K.M. Watson and R.A. Ragatz, Chemical Process Principles, vol. 1, 2nd Ed., John Wiley and Sons, New York, 1954, p. 258.

9. Burton, Robert S., "Fluid Bed Gasification of Solid Waste Materials," Master's Report, West Virginia University, Morgantown, W. Va., 1972.

10. Combustion Engineering, ed. by Otto de Lorenzi, Combustion Engineering, Inc., New York, 1957, p. 29-5.

Table I

Manure Composition, wt%

Component	As Used	Dry, ash-free
Carbon	18.9	52.2
Hydrogen	2.3	6.4
Oxygen	13.3	36.9
Nitrogen	1.2	3.3
Sulfur	0.4	1.2
Ash	13.9	-
Water	50.0	-

Table II

Pyrolysis Gas Composition

Component	Vol. %
CO_2	16.3
CO	35.5
CH_4	11.1
H_2	37.1

Table III

Thermochemical Data for Pyrolysis Reaction

Component	ΔH_c^o (Btu/lb)[1]	$\bar{C}_p$(Btu/lb · °F)[2]
DAFM	-8,750	0.75
Char	-17,060	0.34
CO	-4,348	0.27
CO_2	-	0.26
H_2	-60,997	3.50
CH_4	-23,895	0.85
NH_3	-9,665	0.67
H_2S	-7,114	0.28
O_2	-	0.25
$H_2O(g)$	-	0.50
Ash	-	0.23

[1]DAFM and char by Dulong formula (3) and rest from Himmelblau (7).

[2]DAFM as cellulose from Stamm (4), char from coal with no volatiles (5), ash as sand from Perry's Handbook (6) and rest from Hougen et al. (8).

Table IV

Calculation of Heat of Pyrolysis Reaction

Component	X(lb/lb DAFM)	$X\Delta H_c^\circ$ (Btu)	$X(\bar{C}_p)\Delta T$ (Btu)
Reactants			
DAFM	1.000	-8,750	1067
Products			
CO	0.336	-1,461	129
CO_2	0.242	-	90
H_2	0.025	-1,525	125
CH_4	0.061	-1,458	74
NH_3	0.041	- 396	30
H_2S	0.013	- 92	5
Char	0.282	-4,811	136
Σ Products		-9,743	589

$$\Delta H^\circ_{Rx} = \Sigma \Delta H^\circ_c \text{ react.} - \Sigma \Delta H^\circ_c \text{ prod.} = -8,750 + 9,743 = 993 \text{ Btu/lb DAFM}$$

$$\Delta H^{1500}_{Rx} = \Delta H^\circ_{Rx} + \Delta H_{prod.} - \Delta H_{react.} = 993 + 589 - 1,067 = 515 \text{ Btu/lb DAFM}$$

Table V

Calculation of Heat of Steam-Char Reaction

Component	X(lb/lb DAFM)	$X\Delta H_c^\circ$ (Btu)	$X(\bar{C}_p)\Delta T$ (Btu)
Reactants			
Char	0.071	-1,211	34
H_2O	0.101	-	175*
Σ reactants		-1,211	209
Products			
CO	0.157	- 683	60
H_2	0.015	- 915	75
Σ products		-1,598	135

$$\Delta H_{Rx}^\circ = \Sigma\, \Delta H_c^\circ \text{ react.} - \Sigma\, \Delta H_c^\circ \text{ prod.} = -1{,}211 + 1{,}598 = 387 \text{ Btu/lb DAFM}$$

$$\Delta H_{Rx}^{1500} = \Delta H_{Rx}^\circ + \Delta H_{\text{prod.}} - \Delta H_{\text{react.}} = 387 + 135 - 209 = 313 \text{ Btu/lb DAFM}$$

* Includes heat of vaporization

Table VI

Feed Heating Requirements

Component	X(lb/lb DAFM)	ΔH (Btu/lb)	Heat Load (Btu/lb DAFM)
DAFM	1.000	1,067	1,067
Ash	0.385	484	186
H_2O	0.155	1,721	267
		Total	1,520

Table VII

Heat from Char Combustion

Component	X(lb/lb DAFM)	$X\Delta H_c^\circ$ (Btu)	$\bar{C}_p$ (Btu/lb·°F)	$X(\bar{C}_p)\Delta T$ (Btu)
Reactants				
Char	0.211	-3,600	0.37	131
Air	2.655	-	0.26	1,155
Σ reactants		-3,600		1,286
Products				
CO_2	0.732	-	0.27	331
H_2O	0.103	-	0.51	193*
N_2	2.032	-	0.27	918
Σ products				1,442

$$\Delta H_c^{1750} = \Delta H_c^\circ + \Delta H_{prod.} - \Delta H_{react.} = -3{,}600 + 1{,}442 - 1{,}286$$
$$= 3{,}444 \text{ Btu/lb DAFM}$$

* Includes heat of vaporization

Table VIII

Process Energy Balance

Heat Available

Char Combustion: 0.98(3,444) = 3,375 Btu/lb DAFM

Cool Pyrolysis Gas to 600°F:

Component	lb/lb DAFM	$\bar{C}_p$ (Btu/lb·°F)	Btu/lb DAFM
CO	0.493	0.28	122
CO_2	0.242	0.28	62
H_2	0.040	3.53	127
CH_4	0.061	0.98	54
NH_3	0.041	0.74	27
H_2S	0.013	0.30	4
H_2O	0.054	0.53	26

Total Heat Available = 3,797 Btu/lb DAFM

Heat Required

Pyrolysis Reactor Heat Load: 1.02(2,348) = 2,395 Btu/lb DAFM

Heat Char and Ash for Combustion:

Char ΔH = (0.211 lb/lb DAFM)(0.53 Btu/lb·°F)(250°F)

= 28 Btu/lb DAFM

Ash ΔH = (0.385 lb/lb DAFM)(0.28 Btu/lb·°F)(250°F)

= 27 Btu/lb DAFM

$$\text{Air Rate} = \frac{3{,}797 - (2{,}395 + 28 + 27)\ \text{Btu/lb DAFM}}{(0.26\ \text{Btu/lb}\cdot{}^\circ\text{F})(1673\,{}^\circ\text{F})} = 3.097$$

lb/lb DAFM

APPENDIX B

EQUIPMENT SIZING AND COST ESTIMATION

Equipment costs for the 500 T/D conceptual design plant were estimated using the procedure outlined by Guthrie (1). For some items only the capacity was required, but for others additional sizing information was needed. The total equipment cost was then used to determine fixed and total capital investments. Where necessary costs were updated to mid-1974 values using the Marshall and Stevens index. In all cases except the dryers the escalation is from 1968 with a factor of 1.355. Below are given estimations used for the major equipment items.

<u>Dryers</u>

$$\text{Water removal rate} = \frac{10^6\ \text{lb}}{24\ \text{hr}} = 40{,}000\ \text{lb/hr}$$

From Grzelak (2) the 1965 cost of flash drying units to remove 20,000 lb/hr of water was $175,000. The 1974 cost for two units would be

	$2(\frac{370}{244})$ (175,000)	=	$531,000.
Field material and labor (M & L)	0.5(531,000)	=	265,000
Indirect costs	0.38(796,000)	=	302,000
Bare module cost			$1,098,000

<u>Pyrolysis Reactor</u>

Procedures outlined in Kunii and Levenspiel (3) were used for the fluidized bed design as follows:

$\mu = 0.04$ cp $\qquad \rho_g = 8.83 \times 10^{-4}$ g/cm^3 $\qquad \rho_s = 2.64$ g/cm^3

$\bar{d}_p = 0.04$ cm $\qquad \phi_s = 0.86$ $\qquad \epsilon_{mf} = 0.41$

$$U_{mf} = \frac{(\phi_s\ d_p)^2\ (\rho_s - \rho_g)\ g\ (\epsilon_{mf})^3}{150\ \mu\ (1 - \epsilon_{mf})} = 6\ \text{cm/sec}$$

$$U_t = \left(\frac{4\ (\rho_s - \rho_g)^2\ g^2}{225\ \rho_g\ \mu}\right)\ \bar{d}_p = 139\ \text{cm/sec}$$
$$(\text{for } \bar{d}_p = 0.02\ \text{cm})$$

$d_b = 15$ cm $\quad U_{br} = 81$ cm/sec $\quad$ gas production rate = 143 ft^3/sec

assume at inlet $U_o = 30$ cm/sec

assume reactor diameter = 9 ft

recycle rate = $63.6 \ ft^2 \left(\frac{30}{30.48}\right) = 63 \ ft^3/sec$

gas exit $U_o = \frac{206}{63.6}(30.48) = 99$ cm/sec

$$\delta = \frac{U_o - U_{mf}}{U_b} = \frac{93}{174} = 0.53$$

assume $\frac{L}{D} = 0.7 \quad V_t = 0.7(9)(63.6) = 400 \ ft^3$

$V_{dp} = V_t \ (1 - \delta) = 188 \ ft^3$

sand circulation rate = $\frac{34.5 \ (722,400)}{24(3600) \ (86)} = 3.35 \ ft^3/sec$

$t_{sand} = \frac{188}{3.35} = 56$ sec

TDH = 2.3(9) = 20.7 ft

reactor height = 20.7 + .7(9) = 27 ft

add an additional 5 ft for gas inlet

total reactor height = 32 ft

Costs were estimated using data from Guthrie (1) for pressure vessels with 18 inches of firebrick lining.

Shell	1.355(23,000)	=	$31,200
Field M & L	2.03(31,200)	=	63,300
Lining M & L	1.355(30,100)	=	40,800
Indirect costs	1.2(31,200)	=	37,400
Bare module cost			$172,700

Combustion Reactor

Conditions in the combustion reactor were similar to the pyrolysis reactor in regard to the physical properties so the exit gas velocity was set at 100 cm/sec. The combustion gas rate was 430 ft^3/sec giving a column diameter of 13 ft.

assume L/D = 0.75

TDH = 2(13) = 26 ft

total reactor height = 5 + 26 + .75(13) = 41 ft.

Again 18 inches of firebrick lining were used to give an overall shell diameter of 16 ft.

Shell	1.355(36,000)	=$ 48,800
Field M & L	2.03(48,800)	= 99,100
Lining M & L	1.355(48,700)	= 66,000
Indirect costs	1.2(48,800)	= 58,600
Bare module cost		$272,500

Dust Collection Equipment

Pyrolysis gas rate = 15,000 ft^3/min

Combustion gas rate = 26,000 ft^3/min

Information from Guthrie (1) was used to estimate costs for cyclones to clean both gas streams. In addition, a louvered collector was estimated for the combustion gas stream from data in Popper (4).

Cyclone (pyrolysis)	1.355(6,300)	=$ 8,500
Cyclone (combustion)	1.355(10,200)	= 13,800
Louvered collector	1.355(16,900)	= 22,900
Field M & L	0.69(45,200)	= 31,200
Indirect costs	0.65(45,200)	= 29,400
Bare module cost		=$105,800

Compressors

Horsepower requirements for the recycle and pyrolysis gas compressors were estimated from Perry's Handbook (5). The air compressor was estimated on a unit cost basis from Guthrie (1).

Recycle flow = 3780 ft^3/min

$$HHP = 0.0154(3780)(65)(0.0414) = 160$$

$$BHP = \frac{160}{.97} + 10 = 175$$

Equipment cost	1.355(40,000)	= $54,200
Field M & L	1.21 (54,200)	= 65,600
Indirect costs	0.898(54,200)	= 48,700

Pyrolysis gas flow = 2700 ft^3/min

$$HHP = \frac{0.01(2)(2700)(65)}{0.65} (\sqrt{1.22} - 1) = 564$$

$BHP = \frac{654}{.97} + 25 = 610$

Equipment cost	1.355(100,000)	=	135,500
Field M & L	1.21(135,500)	=	164,000
Indirect costs	0.898(135,500)	=	121,700

Air flow = 24,420 ft^3/min

Multiplier = 18 Unit cost = $2,900

Equipment cost	1.355(18)(2,900)	=	70,700
Field M & L	0.6(70,700)	=	42,400
Indirect costs	0.62(70,700)	=	43,800
Bare module cost			$746,600

Heat Exchanger

Q = 12 MM Btu/hr $\Delta T_{1m} = 736\,°F$

assume $U_o = 5$ Btu/hr·ft^2·°F

$$\text{Area} = \frac{12 \times 10^6}{5(736)} = 3410\ ft^2$$

The heat exchanger cost was estimated from Perry's Handbook (6) at $1.15/ft^2.

Equipment cost	1.355(3,900)	=	$ 5,300
Field M & L	1.4(5,300)	=	7,400
Indirect costs	0.92(5,300)	=	4,900
Bare module cost			$20,600

Pyrolysis Gas Quench and Scrubbing

Tower dimensions were found to be

Quench	5 ft	dia	x	15 ft	
Scrubber	11 ft	dia	x	28 ft	
Desorber	12 ft	dia	x	25 ft	

with a total of 3230 ft^3 of 2-inch Raschig rings required for packing.

Quench shell cost	1.355(5,000)	=	$ 6,700
Scrubber shell cost	1.355(24,000)	=	32,500
Desorber shell cost	1.355(19,000)	=	25,800
Field M & L	2.03(65,000)	=	132,000

Packing	1.355(7.5)(3230)	=	32,800
Indirect costs	1.2(65,000)	=	78,000
Bare module cost			$307,800

Pumps

Scrubbing system rate = 11,000 gpm

ΔP = 300 psi

Equipment cost	1.355(3)(25,000)	=	$101,600
Field M & L	1.412(101,600)	=	143,500
Indirect costs	0.972(101,600)	=	98,800

Quench pump = 450 gpm

ΔP = 75 psi

Equipment cost	1.355(2,000)	=	2,700
Field M & L	1.412(2,700)	=	3,800
Indirect costs	0.972(2,700)	=	2,600
Bare module cost			$353,000

Conveying Equipment

For manure feed converying a screw converyor 1 foot diameter by 100 feet long was assumed, and for ash cooling three hollow flight screw converyors 16 inches diameter by 20 feet long were assumed. It was also assumed the cost of the hollow flight conveyor would be 1.5 times a screw conveyor the same size.

Screw conveyor	1.355(40)(270)	=	$14,600
Cooling conveyor	1.355(2.5)(6)(300)	=	6,100
Field M & L	0.59(20,700)	=	12,200
Indirect costs	0.62(20,700)	=	12,800
Bare module cost			$45,700

Storage Bins

Only enough dried manure storage to allow steady feeding of the pyrolysis reactor was provided. This was estimated to be ½ day's hold-up. Storage for ash was estimated on the basis of 2 days' production. A total of 4 bins each 15 feet diameter by

20 feet high would required.

Storage bins	1.355(4)(300)	=	$1,600
Field M & L	(negligible)		
Indirect costs	0.4(1,600)	=	600
Bare module cost			$2,200

Buildings and Structure

Cost for buildings and structures were estimated from low to average cost data presented by Guthrie (1).

Compressor house			
Shell	1.355(3.25)(3750)	=	$16,500
Services	1.355(3.15)(3750)	=	16,000
Control house			
Shell	1.355(3.75)(900)	=	4,600
Services	1.355(10.00)(900)	=	12,200
Administrative offices			
Shell	1.355(4.30)(1500)	=	8,700
Services	1.355(13.00)(1500)	=	26,400
Shop area			
Shell	1.355(2.50)(1500)	=	5,100
Services	1.355(12.00)(1500)	=	24,400
Manure storage			
Shell	1.355(2.75)(8000)	=	29,800
Services	1.355(1.00)(8000)	=	10,800
Structure			
Shell	1.355(0.45)(10^5)	=	61,000
Indirect costs	0.030(215,500)	=	64,700
Bare module cost			$280,200

Site Development

Site development was estimated as 10% of equipment costs plus acquisition of 25 acres at $500/acre and survey fees at 10% of the land cost.

Site development	$111,400
Land acquisition	12,500
Survey fees	1,300
Bare module cost	$125,200

Offsite Facilities

The following offsite facilities would be required:

Water		$ 4,100
Instrument air		27,100
Flare		110,900
Fire protection		17,100
Fuel gas system		16,900
Power distribution		143,100
Yard lighting		25,500
Payloaders		62,100
Indirect costs	0.34(406,800) =	138,300
Bare module cost		$545,100

Fixed and Total Capital Investments

Fixed and total capital investment costs were estimated by adding contingency costs and other fees to the bare module costs as follows:

Bare module costs	
Processing equipment	$3,124,900
Buildings	280,200
Site development	125,200
Offsite facilities	545,100
Total	4,075,400
20% contingency	815,100
	4,890,500
5% contractors fees	244,500
	5,135,000
10% contingency	513,500
Fixed capital investment	5,648,500
Working capital (15%)	847,300
Total capital investment	6,495,800

REFERENCES

1. Guthrie, K.M., "Capital Cost Estimating," Chemical Engineering, 76 (March 24, 1969), p. 114ff.

2. Grzelak, R.F., "C-E Raymond Flash Drying Systems," Cost Engineering, July 1965, p. 4.

3. Kunii, D. and O. Levenspiel, Fluidization Engineering, John Wiley and Sons, Inc., New York, 1969.

4. Popper, H., ed., Modern Cost Engineering Techniques, McGraw-Hill, New York, 1970, p. 138.

5. Perry's Chemical Engineers' Handbook, Fourth Edition, McGraw-Hill, New York, 1963, p. 6-16.

6. Ibid., p. 11-14.

APPENDIX C

FEEDLOT CAPACITIES IN SOUTHWESTERN KANSAS

Region 1

50-mile radius around Garden City

Nearest Town	Number of Cattle
Cimmaron	12,200
Dighton	19,000
Dodge City	91,000
Garden City	147,600
Holcomb	11,000
Ingalls	43,000
Lakin	43,000
Leoti	88,800
Montezuma	27,000
Satanta	24,300
Scott City	81,500
Sublette	59,200
Syracuse	17,000
Ulysses	57,500
Total	722,100

Region 2

50-mile radius around Dodge City

Nearest Town	Number of Cattle
Ashland	11,000
Cimmaron	12,200
Dodge City	91,000
Fowler	6,200
Garden City	147,600
Ingalls	43,000
Jetmore	5,500
Kinsley	14,500
Meade	10,500
Montezuma	27,000
Sublette	59,200
Total	427,700

Region 3

50-mile radius around Liberal

Nearest Town	Number of Cattle
Fowler	6,200
Liberal	45,000
Meade	10,500
Montezuma	27,000
Satanta	24,300
Sublette	59,200
Ulysses	57,500
Adams, OK	25,000
Beaver, OK	2,000
Guymon, OK	77,500
Hooker, OK	56,800
Turpin, OK	17,000
Hitchland, TX	32,000
Perryton, TX	90,000
Total	530,000

APPENDIX D

EXPERIMENTAL DATA

RUN NO. 5

Coarse sand wt., 18,179 g
Test material, 1,327 g of fine sand (171 g of -65 + 100 mesh sand)
Air flow rate, 55% ΔP(cm Hg): max., 16 avg., 10
Bed ht.(in): static, 16 max., 52 min., 24
Elutriation data for -65 + 100 mesh sand:

Time(min)	Wt.(g)	Time(min)	Wt.(g)
0.75	3.4	8	6.1
1.50	2.7	14	11.5
2.25	1.8	30	34.9
3.25	2.2	60	40.4
5	4.2	90	20.7

RUN NO. 10

Coarse sand wt., 18,000 g
Test material, 2,000 g of -100 + 150 mesh sand
Air flow rate, 55% ΔP(cm Hg): max., 14 avg., 9
Bed ht. (in): static, 15 max., 43 min.,24
Elutriation data:

Time(min)	Wt.(g)	Time(min)	Wt.(g)	Time(min)	Wt.(g)
1	30.7	8	65.6	21	130.1
2	42.8	9	64.6	25	139.2
3	56.9	10	65.0	30	141.8
4	69.4	12	114.9	37	146.7
5	75.6	15	154.1	47	133.3
6	74.6	18	144.5	64	124.4
7	72.8				

Bed wt. at end, 18,110 g

RUN NO. 12

Coarse sand wt., 18,787 g
Test material, 1,213 g of -150 + 250 mesh sand
Air flow rate, 55%   ΔP(cmHg): max., 11   avg., 9
Bed ht.(in): static, 15   max., 43   min., 24
Elutriation data:

Time(min)	Wt(g)	Time(min)	Wt.(g)	Time(min)	Wt.(g)
1	103.1				
2	86.3	6	112.1	11	114.0
3	100.2	7	106.0	14	98.7
4	119.3	8	94.7	20	58.8
5	115.0	9	74.7	60	29.3

Bed wt. at end, 18,875 g

RUN NO. 13

Coarse sand wt., 18,642 g
Test material, 1,358 g of -250 + 325 mesh sand
Air flow rate, 55%   ΔP(cmHg): max., 11   avg., 8
Bed ht.(in): static, 14.5
Elutriation data:

Time(min)	Wt.(g)
1	146.6
5.25	923.8
7	59.4
31	20.2

Bed wt. at end, 18,354 g

RUN NO. 14

Coarse sand wt., 18,850 g
Test material, 1,150 g of -250 + 325 mesh sand
Air flow rate, 55%   ΔP(cmHg): max.,   avg., 8
Bed ht.(in): static, 15
Elutriation data:

Time(min)	Wt.(g)	Time(min)	Wt.(g)	Time(min)	Wt.(g)
0.75	108.6	2.5	100.8	5	109.3
1.5	125.4	3	126.1	7	77.8
2	151.0	4	144.2	30	36.5

Bed wt. at end, 18,852 g

RUN NO. 15

Coarse sand wt., 18,788 g
Test material, 1,212 g of -150 + 250 mesh sand
Air flow rate, 55% ΔP(cmHg): max., avg., 10
Bed ht.(in): static, 15 max., 48
Elutriation data:

Time(min)	Wt.(g)	Time(min)	Wt.(g)	Time(min)	Wt.(g)
1	51.2	6	91.0	12	99.6
2	113.8	7	90.0	15	115.3
3	95.2	8	70.4	30	110.5
4	130.2	10	126.6	60	12.6
5	98.9				

Bed wt. at end, 18,798 g

RUN NO. 16

Coarse sand wt., 18,798 g
Test material, 1,205 g of -150 + 250 mesh sand
Air flow rate, 55% ΔP(cmHg): max., avg., 10
Bed ht.(in): static, 15 max., 24 min., 21
Elutriation data:

Time(min)	Wt.(g)	Time(min)	Wt.(g)	Time(min)	Wt.(g)
1	50.4	4	105.8	8	114.8
2	170.8	4.75	133.0	13	87.8
2.75	172.3	6	139.3	31	34.9
3.5	168.6				

Bed wt. at end, 18,765 g
Note: 1 static mixer 12" above distributor

RUN NO. 18

Coarse sand wt., 19,020 g
Test material, 980 g of -250 + 325 mesh sand
Air flow rate, 55% ΔP(cmHg): max., avg., 10
Bed ht.(in): static, 15 max., 33 min., 22
Elutriation data:

Time(min)	Wt.(g)	Time(min)	Wt.(g)	Time(min)	Wt.(g)
0.5	163.1	2	124.1	8	50.9
1	202.6	2.75	121.1	30	26.2
1.5	159.6	4	107.8		

Bed wt. at end, 18,960 g
Note: 1 static mixer 12" above distributor

RUN NO. 19

Coarse sand wt., 18,822 g
Test material, 1,178 g of -150 + 250 mesh sand
Air flow rate, 55% ΔP(cmHg): max., avg., 9
Bed ht.(in): static, 16 max., 30 min., 20
Elutriation data:

Time(min)	Wt.(g)	Time(min)	Wt.(g)	Time(min)	Wt.(g)
1	125.4	5	133.4	11	126.1
2	143.3	6	109.7	15	54.1
3	140.1	8	149.0	46	75.4
4	142.1				

Bed wt. at end, 18,784 g
Note: 1 static mixer 6" above distributor

RUN NO. 20

Coarse sand wt., 18,801 g
Test material, 1,199 g of -150 + 250 mesh sand
Air flow rate, 55% ΔP(cmHg): max., avg., 8
Bed ht.(in): static, 16 max., 26 min., 21
Elutriation data:

Time(min)	Wt.(g)	Time(min)	Wt.(g)	Time(min)	Wt.(g)
1	102.6	3.5	117.2	8	128.2
2	152.4	4.5	146.7	12	126.8
2.75	141.4	6	153.8	30	68.1

Bed wt. at end, 18,761 g
Note: 2 static mixers 6" and 18" above distributor

RUN NO. 21

Coarse sand wt., 18,992 g
Test material, 1,008 g of -325 mesh sand
Air flow rate, 55%
Bed ht.(in): static, 15
Elutriation data:

Time(min)	Wt.(g)	Time(min)	Wt.(g)	Time(min)	Wt.(g)
0.75	317.9	2	41.6	5	31.4
1	149.4	2.5	25.2	34	20.5
1.5	113.4				

Bed wt. at end, 19,043 g

RUN NO. 23

Coarse sand wt., 18,000 g
Test material, 1,832 g ash including material >10 mesh
Air flow rate, 55%
Bed ht.(in):
Elutriation data:

Time(min)	Wt.(g)	Time(min)	Wt.(g)	Time(min)	Wt.(g)
0.5	248.7	2	59.7	10	63.1
1	233.1	3	70.5	30	75.4
1.5	111.6	6	58.4		

Bed wt. at end, 18,516

RUN NO. 24

Coarse sand wt., 18,495 g
Test material, 1,505 g ash (all <10 mesh)
Air flow rate, 55%
Bed ht.(in): static, 16
Elutriation data:

Time(min)	Wt.(g)	Time(min)	Wt.(g)	Time(min)	Wt.(g)
0.25	206.9	1.5	143.8	7	41.2
0.5	156.7	2.5	124.1	37	74.1
1	286.9	4	57.0		

RUN NO. 25

Coarse sand wt., 18,000 g
Test material, 1,954 g ground manure
Air flow rate, 55%
Bed ht.(in):
Elutriation data:

Time(min)	Wt.(g)	Time(min)	Wt.(g)	Time(min)	Wt.(g)
0.3	119.8	4	42.7	9.5	26.5
0.75	101.5	4.75	28.4	12	33.2
1.25	66.4	5.75	31.8	20	67.9
1.75	28.8	6.75	27.1	39	66.7
3	67.4	8	29.1		

Bed wt. at end, 19,105 g

RUN NO. 26

Coarse sand wt., 19,275 g
Test material, 725 g elutriated manure
Air flow rate, 55%
Bed ht.(in): static, 15
Elutriation data:

Time(min)	Wt.(g)	Time(min)	Wt.(g)	Time(min)	Wt.(g)
0.5	145.3	3	99.8	7	47.4
1	118.6	4	52.5	30	97.2
2	106.8	5	33.9		

RUN NO. 27

Continuous feeding of ground manure (<¼ in)
Coarse sand wt., 20,000 g   Static bed ht., 16 in.
Air flow rate, 55%   Feeder speed set to give 440 g/min

Time(min)	Elutriation rate (g/min)
6	90
9	91

Length of time for run, 11 min
Bed wt. at end of run, 22,972 g   Bed ht., 24 in.
Total elutriated, 1,021 g
Avg. feed rate, 363 g/min.

RUN NO. 29

Continuous feeding of ground manure (<¼ in)
Coarse sand wt., 20,000 g (including some large manure particles)
Static bed ht., 20 in   Air flow rate, 55%
Feeder speed set to give 200 g/min
Wt. of manure in feeder, 4,635 g
Elutriation data:

Time(min)	Wt.(g)	Time(min)	Wt.(g)	Time(min)	Wt.(g)
Start[a]	20.2	5	82.0	10	74.2
1	43.2	6	75.6	11	78.0
2	52.2	7	88.4	12[b]	69.0
3	74.2	8	85.0	30	350.4
4	76.6	9	75.0		

[a]From manure mixed with coarse sand at start.
[b]Feeder turned off.

Bed wt. at end of run, 21,679 g   Left in feeder, 1,666 g
Bed ht. at end of run, 26 in.
Total elutriated, 894 + 350 = 1,244 g
Avg. feed rate, 247 g/min

RUN NO. 33

Continuous feeding of elutriated manure
Coarse sand wt., 20,000 g   Static be ht., 16 in.
Air flow rate, 55%   Feeder speed set to give 153 g/min
Wt. of manure in feeder, 3,670 g
Elutriation data:

Time(min)	Wt.(g)	Time(min)	Wt.(g)
1	59	6	130
2	104	7	143
3	124	8	133
4	126	9	157
5	130	10	165

Bed wt. at end of run, 20,310 g   Bed ht., 17 in.
Wt. left in feeder, 2,032 g   Total elutriated, 1,271 g
Avg. feed rate, 164 g/min.

RUN NO. 34

Continuous feeding of ash
Coarse sand wt., 20,000 g   Static bed ht., 16 in
Air flow rate, 55%   Feeder speed set to give 103 g/min
Wt. of ash in feeder, 6,925 g
Elutriation data:

Time(min)	Wt.(g)	Time(min)	Wt.(g)
1	74	7	113
2	106	8	119
3	103	9	119
4	110	10	121
5	121	11	123
6	122	(2nd cyclone)	8

Bed wt. at end of run, 20,035 g
Wt. left in feeder, 5,512 g   Total elutriated, 1,239 g
Avg. feed rate, 128 g/min

APPENDIX E

ELUTRIATION CONSTANT DETERMINATION

Example: Run 10 data

Calculate $\frac{W(d_p)}{W_o(d_p)}$ where

$$W(d_p) = W_o(d_p) - (\text{amount elutriated through time t})$$

Time(min)	$\frac{W(d_p)}{W_o(d_p)}$	Time(min)	$\frac{W(d_p)}{W_o(d_p)}$	Time(min)	$\frac{W(d_p)}{W_o(d_p)}$
1	0.985	8	0.756	21	0.419
2	0.963	9	0.724	25	0.350
3	0.935	10	0.691	30	0.279
4	0.900	12	0.634	37	0.205
5	0.862	15	0.557	47	0.139
6	0.825	18	0.484	64	0.077
7	0.789				

Plot log $\frac{W(d_p)}{W_o(d_p)}$ vs t as in Figure 1.

Determine slope of curve as shown to give

$$K = \frac{\ln(.638)}{11.7 - 1.7} = 0.0449/\text{min.}$$

$$K^* = \frac{(.0449/\text{min})(20{,}000\ \text{g})}{(60\ \text{sec/min})(324\ \text{cm}^2)} = 0.0462\ \text{g/cm}^2\cdot\text{sec}$$

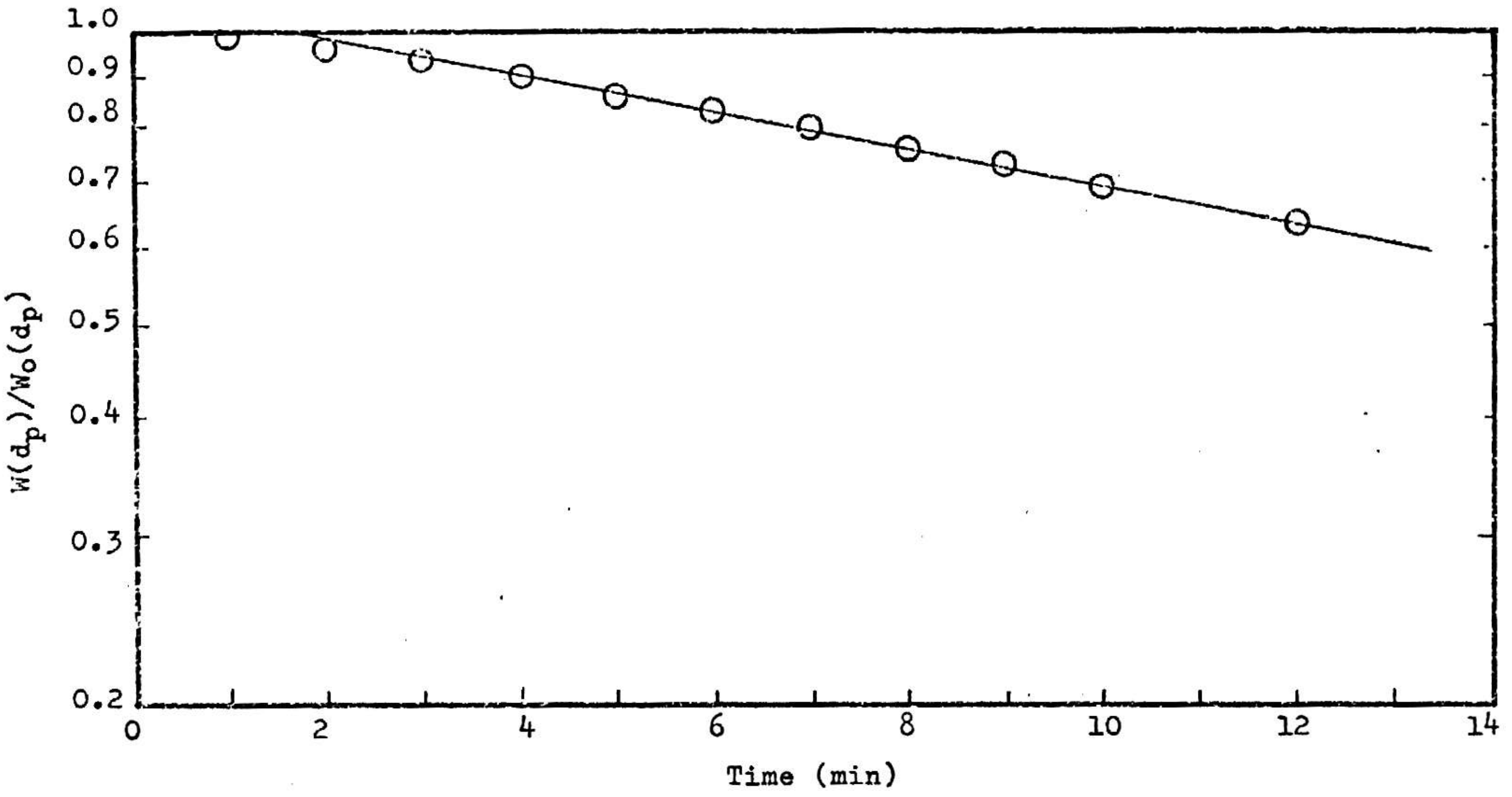


Fig. 1. Data of Run 10 Plotted to Determine Value of K.

APPENDIX F

CALCULATIONS FOR ESTIMATED RESULTS FOR RUN 33

1. Calculate feed rate ($F_o(d_p)$) for each particle size using distribution for elutriated manure given in Table II, Chapter IV and assuming average feed rate of 164 g/min.

2. Estimate terminal velocity (U_t) from Figure 7, Chapter IV for each particle size.

3. Calculate the abscissa for the Wen and Hashinger correlation given in Figure 9, Chapter IV, then calculate K* from the value of the ordinate.

4. Calculate K from

$$K = \frac{60\ K^*\ A}{W}$$

5. Calculate the amount of each particle size collected in the bed during the run as

$$W(d_p) = \frac{F_o(d_p)}{K} \quad (1 - e^{-Kt})$$

6. Calculate the amount elutriated as

$$E\ (d_p) = F_o(d_p) \cdot t - W(d_p).$$

Calculated values for Run 33 are given in Table I.

Table I

Calculated Values for Run 33

d_p (cm)	$F_o(d_p)$ (g/min)	u_t (cm/sec)	Abscissa (x 10^{-5})	K* (g/cm^2·sec)	K (min^{-1})	$W(d_p)$ (g)	$E(d_p)$ (g)
0.0478	18	71.8	0.89	0.0385	0.0374	146	34
0.0376	26	56.3	1.14	0.0575	0.0559	171	89
0.0240	25	33.2	1.10	0.122	0.119	136	114
0.0170	15	20.8	0.87	0.200	0.194	84	66
0.0120	12	12.6	0.62	0.325	0.316	75	45
0.0083	23	7.2	0.41	0.53	0.515	165	65
0.0048	45	3.1	0.20	1.08	1.05	378	72
	164					1155	485

APPENDIX G

CALCULATIONS FOR ESTIMATED ASH BUILD-UP

1. Estimate viscosity of pyrolysis gas at reactor conditions from

$$\mu = \frac{\sum_i y_i \mu_i (M_i)^{\frac{1}{2}}}{\sum_i y_i (M_i)^{\frac{1}{2}}}$$

where y_i = mole fraction of component i,

μ_i = viscosity of component i at conditions and

M_i = molecular weight of component i

as given by Maxwell (1). Viscosity estimated as 0.04 cp.

2. Estimate pyrolysis gas density (assuming perfect gas) as 0.0004 g/cm^3.
3. Calculate terminal velocity as a function of particle size from procedure in Kunii and Levenspiel (2). The function is shown in Figure 1.
4. Estimate K* using extrapolation procedure described in Chapter IV. Values of K* as a function of d_p are shown in Figure 2.
5. Estimate an average K using particle size distribution for elutriated ash given in Table II, Chapter IV. The estimation is shown in Table I.
6. Estimate average K for combustion reactor assuming gas has same physical properties as pyrolysis gas and combustion reactor is 12" diameter.

$$K = 0.052\left(\frac{20,000}{324}\right)\left(\frac{730}{25,000}\right)$$

$$K = 0.094 \text{ min}^{-1}$$

7. Use data from Figure 3 to estimate F (solids overflow rate from pyrolysis reactor) and x (quantity of ash circulating

with sand from combustion reactor) based on the assumption that char is consumed instantaneously in the combustion reactor at the rate of 48 g/min. The calculation uses the following equation given in Kunii and Levenspiel (5):

$$F_1 = \sum_{d_p} \frac{F_o(d_p)}{1 + \frac{W\,K(d_p)}{F_1}}$$

where F_1 = solids overflow rate,

$F_o(d_p)$ = feed rate for size d_p,

W = total bed weight and

$K(d_p)$ = elutriation constant for size d_p.

Balances around each fluidized bed give

$$F_{1a} = \frac{136 + x}{1 + \frac{20{,}000(.052)}{F_{1a}}} + 7858 - x$$

$$7858 = \frac{88 + x}{1 + \frac{25{,}000(.094)}{7858}} + 7858 - x.$$

Solution of the simultaneous equations results in

F_{1a} = 7,944 g/min

x = 294 g/min.

The pyrolysis reactor cyclone load is then

7858 + 136 - 294 = 50 g/min.

and the percent of ash in the solids circulating from the combustion reactor is 3.7%.

REFERENCES

1. Maxwell, J.B., Data Book on Hydrocarbons, D. Van Nostrand, Princeton, N.J., 1950, p. 157.

2. Kunii, D. and O. Levenspiel, Fluidization Engineering, John Wiley and Sons, New York, 1969, p. 76ff.

3. Wen, C.Y. and R.F. Hashinger, "Elutriation of Solid Particles from a Dense-Phase Fluidized Bed," A.I.Ch.E. Journal, 6, p.220.

4. Yagi, S. and T. Aochi in Fluidization Engineering, by D. Kunii and O. Levenspiel, John Wiley and Sons, New York, 1969, p.315.

5. Kunii, D. and O. Levenspiel, op.cit., p. 333.

Table I

Estimation of Average K Value

d_p(cm)	K (min^{-1})	Wt. fraction	$\frac{\text{Wt. fraction}}{K}$
0.0240	0.0185	0.10	5.41
0.0170	0.0306	0.07	2.29
0.0120	0.0496	0.13	2.62
0.0083	0.0758	0.64	8.44
0.0048	0.141	0.06	0.43
			19.19

$$K_{avg} = \frac{1}{19.19} = 0.052 \text{ min}^{-1}$$

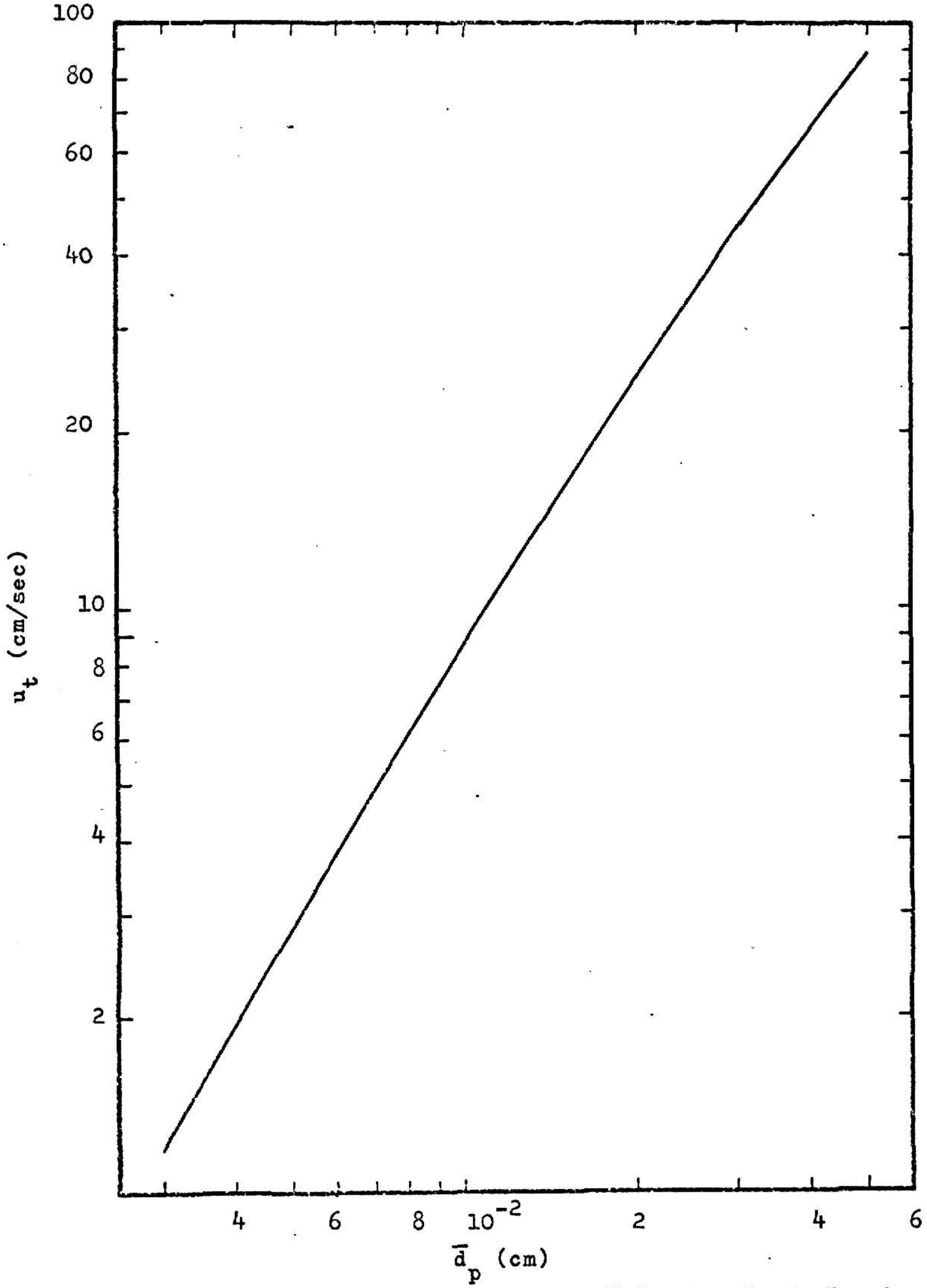


Fig. 1. Terminal Velocity for Ash Under Pyrolysis Reactor Conditions

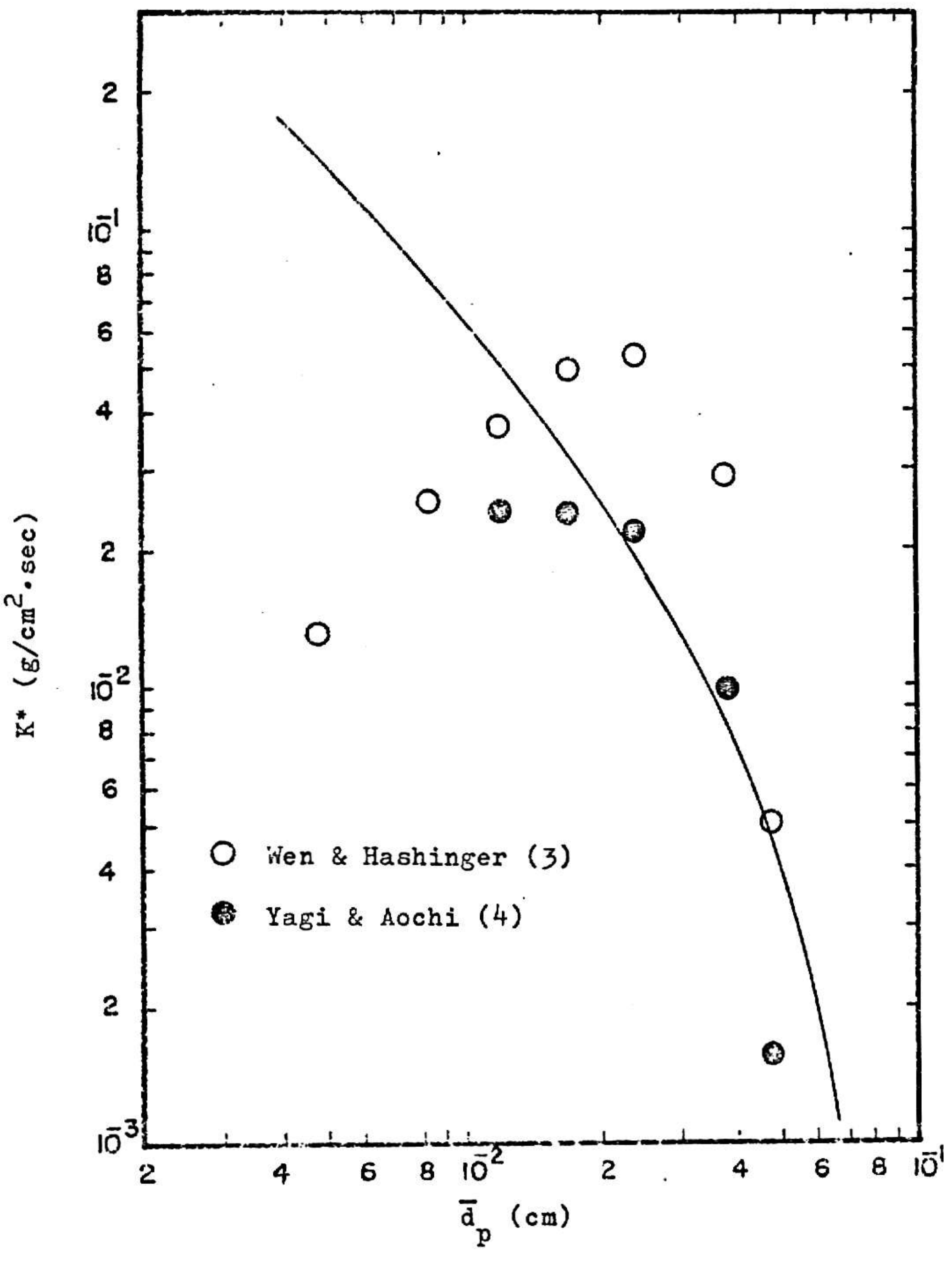


Fig. 2. Estimated Elutriation Constants for Ash under Pyrolysis Reactor Conditions

FLUIDIZED BED PYROLYSIS
OF CATTLE FEEDLOT MANURE

by

CADY ROY ENGLER

B.S., Kansas State University, 1969

AN ABSTRACT OF A MASTER'S THESIS

submitted in partial fulfillment of the

requirements for the degree

MASTER OF SCIENCE

Department of Chemical Engineering

KANSAS STATE UNIVERSITY
Manhattan, Kansas

1974

ABSTRACT

The economic feasibility and potential applications for producing synthesis gas (a mixture of CO and H_2) from cattle feed-lot manure via fluidized bed pyrolysis was studied. The economic analysis was based on a conceptual design for a pyrolysis plant processing 500 tons per day of dry manure. In addition, a fluidized bed simulation unit was developed to provide partial design information for the construction of a pilot scale gasifier.

An estimated 10.7 million standard cubic feet per day of synthesis gas could be produced by the conceptual design plant at a cost, including profit, of $0.85 per thousand standard cubic feet. Sensitivity analysis showed that the cost of producing gas was affected significantly by changes in plant capacity, incoming manure moisture content, and the cost of transporting manure from feedlots to the plant. Changes in chemical composition of the dry, ash-free manure had little effect on gas production costs.

Based on the economic analysis and a survey of feedlot capacities in southwestern Kansas, the potential application of manure pyrolysis was studied. Plants processing up to 4,125 tons per day of dry manure could be supported with the cost of producing synthesis gas about $0.50 per thousand standard cubic feet. The gas could be used as a raw material and energy source for making up to 346,500 tons per year of ammonia or could be used to run electricity generating plants rated at up to 325 megawatts.

The fluidized bed simulation unit was used to study elutriation of sand, manure, and ash particles in addition to observing the general operating characteristics of the bed. Both batch and continuous feeding experiments were conducted. Experimental elutriation rates were higher than predicted from existing corr-elations. The differences were explained by inaccuracies inherent in the correlations. The results were used to estimate ash build-up in the solids for a pilot scale gasifier which was found to be minimal.