1 of 2

Conceptual Design of a Fluidized Bed Combustor For Volume Reduction of Waste

Topical Report

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June 1992

Work Performed Under Contract No.: DE-FC21-87MC24207

For
U.S. Department of Energy
Office of Fossil Energy
Morgantown Energy Technology Center
Morgantown, West Virginia

By
West Virginia University
National Research Center for Coal and Energy
Morgantown, West Virginia

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EXECUTIVE SUMMARY

A group of research engineers was assembled to investigate novel fluidized bed combustion systems for incinerating low-level radioactive wastes. The goal of this project was to significantly reduce the volume of these wastes in an efficient and environmentally safe manner. The process is to be limited by a maximum temperature of 600°C and the waste product was to contain a minimum of waste material produced by the process itself (refractory, absorbents, catalysts, etc.). The approach presented in this study is to evaluate; 1) a modification of the existing system, 2) a hybrid reactor concept, and 3) a fast circulating reactor concept. Carbon dioxide and oxygen were to be the only gases introduced into the reactors.

The results of this study indicate that the existing system may be more efficient with gas and solids recirculation; however, two beds are required. The hybrid reactor concept involves a bubbling bed and a fast bed in one reactor. Although a simple operation is envisioned for this system, there are several technical questions which must be addressed in order to optimize the system for a final analysis. The fast circulating reactor concept also appears to represent a simple system to operate; however, it also presents several technical questions which must be addressed before a thorough evaluation of this concept may be completed.

In conclusion, this report represents a first evaluation of new concepts for significant volume reduction of low-level radioactive wastes. The technical issues required for a complete evaluation of these concepts are presented. A future research effort is outlined the result of which should significantly increase our knowledge of these issues.

CHAPTER 1 INTRODUCTION

1.1 BACKGROUND

This report describes a preliminary design study to develop a fluidized-bed combustion system for the incineration of low-level radioactive wastes (radwaste). This study includes a re-evaluation of the original system at the Rocky Flats USDOE Facility and suggested modifications to this system. Two new concepts employing more modern technology are then presented. These two new concepts are: 1) hybrid beds and 2) fast fluidized beds. Recommendations are then represented for a more comprehensive continuation of this effort.

The Department of Energy's Rocky Flats Facility has accumulated mixed hazardous and radioactive wastes that would require a decade to burn at a feed rate of 1.2 Million Btu/hr. The waste consists of plastics, air filters, coveralls, machining oils, carbon tetrachloride, etc. It is necessary to reduce the volume of these wastes so that other permanent disposal techniques can be used for containment of the waste. Fluidized bed combustion is the best demonstrated available technology for so treating these wastes.

Fluidized beds are desirable because they provide intimate contact between the solid (waste) and the gas (oxygen) necessary for combustion. Heat and mass transfer coefficients in fluid beds are high, and mixing is good so that hot and cold regions in the combustor are eliminated. Solids addition and removal are easily achieved, and the combustor embodies no moving parts. Bubbling fluidized beds have been widely employed, but turbulent, fast or circulating beds are now usually favored. This report identifies the advantages and disadvantages of each system for the waste combustion and the development of each conceptual design of the bed.

1.2 SCOPE OF WORK

The design of a complete system for the volumetric reduction of radwaste, is a very complicated problem. The objective of the WVU contribution to the complete system was limited to the determination of a suitable reactor for the overall system; therefore, the WVU effort was restricted to the boundaries as shown in Figure 1.1. The obvious goals for such a system are to minimize the use of chlorine sorbents, catalysts, and off-gases per unit of radwaste. References [1,2] were the basis for the initial efforts of this study which describes the performance of the existing reactor at the Rocky Flats Facility. This reference provided the necessary data for the average composition of the radwaste and it also provided information which allowed for estimates for the overall flow conditions at various locations in the existing reactors.

Although this study is not concerned with the exhaust gas clean-up, the off-gases should not cause any technical or economic hardships. The gases chosen for fluidization are oxygen and carbon dioxide. The exhaust gases will then be carbon dioxide, with a small amount of water vapor, and a trace of nitrogen. In addition, the use of CO₂ as a fluidizing gas allows for lower velocities in the reactor and, subsequently, reduced erosion problems.

1.3 DESIGN RESTRICTIONS

The basic concerns of any system to reduce the volume of radwaste are:

- 1) safety
- 2) reliability
- 3) the minimization of the amount of disposable waste, and
- 4) the production of a solid waste, such that the radioactive material may be readily extracted.

The specific constraints assigned to this study were:

- 1) a one to two million Btu/hr heat generation
- 2) a small diameter reactor (approx. 6 inches)
- 3) an operating pressure less than one atmosphere pressure
- 4) a readily maintainable system
- 5) minimum erosion
- 6) seven to fifteen year lifetime
- 7) to minimize solid refuse from radwaste
- 8) an all metal structure to avoid radioactive-contaminated-refractory disposal problems
- 9) an operating temperature under 600°C
- 10) that natural gas not be used.

The first constraint is essentially on the feed rate, which then implies that the minimum feed rate of radwaste will be about 82 kg/hr.

Constraints 2 and 3 relate to safety. A reactor vessel with a diameter of less than six-inches will preclude any accumulation of radioactive material which may form a critical mass. For the low level of radioactivity in the radwaste to be consumed at Rocky Flats, the six-inch criterion is not applicable in this study. An operating pressure of less than one atmosphere will insure that, if a leak occurred, the flow would be into the reactor.

Constraints 4, 5, and 6 relate to reliability. The system should be readily maintainable and all potential problem locations are to be observable and cleanable. A minimum of erosion is to be encountered such that no unscheduled shut-downs are encountered and that problem sections may readily be replaced on a scheduled routine. The operating conditions are to be selected so that the reactor will essentially have a useful life of seven to fifteen years.

Constraints 7, 8, 9, and 10 are related to the waste streams from the reactor. Obviously, the initial radwaste material is to be reduced to a significantly smaller volume. The use of any reactor material which will deteriorate with time, and be contaminated, is to be avoided. Also, restricting the operating temperature of the system to under 600°C allows for easy extraction of the radioactive particles from the solid residue.

1.4 FLUIDIZED-BED SYSTEMS

Several fluidization systems and flow regimes will be discussed in this report. It is, therefore, appropriate to present a short introduction to the terminology and fluidization regimes employed. Figure 1.2(a) illustrates a basic fluidization system with a bubbling bed. Fluidizing gases enter the system, flow into a plenum chamber, through a distributor plate, and up through a bed of material. Particles erupting from the bed, and the fluidizing gases, then flow through the freeboard. The particles reaching the top of the freeboard region then flow into the cyclone where most of the particles are separated from the gases.

The characteristics of the fluidized bed are controlled by the superficial velocity, particle characteristics, bed geometry, etc. The superficial velocity is the gas velocity across the bed if no bed material were present. The flow regimes as a function of superficial velocity (u_0) are shown in Figure 1.2(b). For u_0 less than u_{mf} , the particles in the bed support each other and the bed is a fixed bed. When u_{mf} is equal to u_0 , then the particles begin to float in the gas flow and the pressure drop corresponds to the weight of the bed material. This is referred to as particulate fluidization, Figure 1.3. Increasing the superficial velocity, one eventually reaches u_{mb} where bubbles begin to appear. Again increasing u_0 causes the bubble activity to grow. Eventually, the bubbles become so large that they essentially cover the cross sectional area of the bed. This condition is denoted as slugging and begins when u_0 equals u_{ms} . Continuing to increase u_0 eventually leads to u_k , the minimum turbulent flow regime velocity. The velocity u_0 is now so large that a significant portion of the bed material is in the freeboard region. The turbulent flow regime corresponds to clusters of bed materials floating in the gas flow. As u_0 is increased further, more of the bed material flows through the freeboard region and out of the system. This condition is referred to as the fast fluidization regime. Increasing u_0 further and decreasing the feed rate eventually changes the system to part of a pneumatic transport system. These flow regimes as shown in Figure 1.3.

The values of u_{mf} , u_{mb} , u_{ms} , and u_k depend on the particle size and density in the bed, as well as the bed geometry. Figure 1.4 describes a method for classifying particles. Particles in the C range are very small and very difficult to fluidize. The A

range particles are easy to fluidize. The B range particles are also easy to fluidize and u_{mb} is essentially equal to u_{mf} . The D range particles are not as easy to fluidize and frequently u_{mf} is essentially equal to u_{ms} . One frequently finds, in the literature, that D particles should be fluidized in a spouted bed which is essentially a bed in which a jet of gas issues up through the center of the bed. This strong action insures good circulation (movement) of the large bed particles.

The terminal velocity (u_t) of a particle is that minimum velocity which would carry a particle up and away in the freeboard region. Unfortunately, the flow is very complicated in the freeboard region and subsequently, u_t is only a parameter used in attempting to describe the flow above the bed.

Figures 1.5 and 1.6 attempt to organize the information on flow regimes relative to a dimensionless velocity and particle diameter (including gas and solid density effects). These diagrams are for beds which have no internals, which is the condition encountered in this study. The existing primary reactor is designed to be a bubbling bed with a $u^* = 0.66$ and $d_p^* = 15.4$ (based on the Cr_2O_3). The existing bed is a mixture of particles of the same size.

The design of a new fluidized bed reactor for this study basically involves the following rationale. Starting with the flow rate of the radwaste, one may calculate the desired oxygen flow rates and locate the oxygen injection points. With the selection of the average bed particle size, an appropriate fluidization velocity may be selected. Then, with a prescribed reactor diameter, the mass flow rate of fluidizing gas may be estimated. This flow rate must be compared with the required calculated flow rates of oxygen. If there is a problem of incompatibility, then the flow rate of the radwaste must be reduced, the reactor diameter increased, or the particle size may be changed (within limits). Thus, the designs proposed may not contain all the constraints initially imposed in this study.

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- [1] Atkinson, C., et al "Conceptual Design of a Fluidized Bed Combustor for Volume Reduction of Waste" Proposal submitted to USDOE by EWRC - WVU under the WVU/DOE - METC Cooperative Agreement, DE-FC21-87MC 24207.
- [2] Meile, L.J., et al "Rocky Flats Plant Fluidized Bed Incinerator" RFP - 3249, UC -70 Nuclear Waste Management, DOW/TIC - 4500, March 1982.
- [3] Geldart, D. (ed) Gas Fluidization Technology, Wiley, Chichester, 1986.
- [4] Kunii, D. and Levenspiel O., Fluidization Engineering, Butterworth-Heinemann, Boston, 1991.

Figure 1.1 Boundaries of the Present Study

(a) Basic Fluidization System

(b) Bed Characteristics Variation with Superficial Velocity

Figure 1.2 Fluidization Concepts

Figure 1.3 Visual Features of Principal Flow Regimes

Figure 1.4 Simplified Classification of Powders [3]

Figure 1.5 Flow Regimes for a Fluidized Bed

Figure 1.6 Turbulent Flow Regime Map

CHAPTER 2 BASIC INFORMATION

2.1 INTRODUCTION

In order to develop new design concepts, one must have knowledge and information about the substances with which one must deal. In this chapter, a description of the typical fuel will be presented. The available literature on the rate-controlling chemical reaction will then be described. Afterwards, a short discussion of catalysts will be described. Since the radwaste contains undesirable species which may shorten the life of the reactor, a discussion of sorbents will also be presented. Finally, selected data for heat transfer will be given.

2.2 DESCRIPTION OF RADWASTE

While a general description of the radwaste (fuel) may be found in Reference [1], it is inadequate for describing combustion reactions. Consequently, the description of the radwaste was reformulated in the following manner:

Latex:

25%, 10,000 Btu/lb_m, 0% Ash

PVC: Polyvinyl Chloride: $(-\text{CH}_2\text{CHCl}-)_x$,
1.0 %, 13,500 (Btu/lb), Ash 0.0 (%),
Density 1360-1580 (kg/m³).

Polyethylene: $(-\text{CH}_2\text{CH}_2-)_x$,
4.5 %, 20,000 (Btu/lb), Ash 0.0 (%).

Polyurethane Resins: $[(\text{CH}_2)_6\overset{\text{H}}{\underset{|}{\text{N}}}-\overset{\text{O}}{\underset{||}{\text{C}}}-\text{O}(\text{CH}_2)_{144}\overset{\text{O}}{\underset{||}{\text{C}}}-\overset{\text{H}}{\underset{|}{\text{N}}}]_x$,

0.5 %, 13,000 (Btu/lb), Ash 0.0 %.

Water: (H₂O), 5.0 (%), -1,000 (Btu/lb).

Miscellaneous: 5.0 %, Ash 100.0 %.

Wood: 59.5 % (includes Paper, Cloth, Leather)

(dry basis %)

C 50.18%; H 5.06%;

O 43.00%; N 0.30%

S 0.15%; Ash 1.31%.

High Heating Value 8,770 (Btu/lb).

Ultimate Radwaste properties:

(Dry Ash Free Basic)

C 63.60%; H 7.50%;

O 27.90%; N 0.20%;

S 0.07%; Cl 0.73%

Moisture 5.0%

Ash 1.31%.

Heating Value: 7,779 Btu/lb or 18,094 kJ/kg

2.3 COMBUSTION

The combustion processes associated with hydrocarbons in the presence of catalysts and sorbents at relatively low temperatures are very difficult to describe. After a thorough review of the combustion literature, it was apparent that insufficient information was available for modeling the combustion (oxidation) processes. A useful engineering approach to a chemical reaction problem is to determine the slowest reaction in the overall process. If this reaction rate determining reaction can be determined, then an estimate of its chemical kinetics data may be made. Subsequently, the reaction rate for the rate determining reaction provides an order of magnitude rate of reaction for the overall reaction. Unfortunately, the partial oxidation of the hydrocarbons is difficult to estimate; especially in the presence of catalysts and sorbents. Thus one cannot ascertain the rate limiting reaction in the partial oxidation of the hydrocarbons. However, the completion of the combustion (oxidation) process involves the reaction

which may be the rate limiting reaction in the afterburner or fast sections of the systems considered in this study. Useful information has been found for this

The combustion of a hydrocarbon fuel in an oxygen-lean environment will lead to the formation of a carbon-monoxide-rich gas exhaust stream from the pyrolysis/lean combustion stage. This will lead to the production of carbon dioxide in later reactions when additional oxygen is present. Very small amounts of water vapor (about 20 ppm) will catalyze the oxidation of carbon monoxide and greatly enhance the chemical reaction

kinetics. The CO-O₂ oxidation reaction is generally acknowledged to proceed in the following manner:

The most important step in the oxidation process is

The OH (hydroxyl) radical determines the CO oxidation kinetics, but, as it is always regenerated in the chain, the actual initial concentration of hydrogen-containing catalyst (i.e., H₂O) need not be large.

The rate at which the oxidation reaction occurs is given by

$$\frac{-d[\text{OH}]}{dt} = \frac{-d[\text{CO}]}{dt} = \frac{d[\text{H}]}{dt} = \frac{d[\text{CO}_2]}{dt} = k[\text{CO}][\text{OH}] \quad (2.3.1)$$

However the hydroxyl radical is not a stable species, and its concentration must be estimated.

The temperature dependence of the rate constant k can be expressed in an Arrhenius form as

$$k = A \exp (E_a/RT) \quad (2.3.2)$$

where A is the pre-exponential factor and E_a is the activation energy. Baulch & Drysdale, [2] have reported the Arrhenius rate constant as $\log (k/\text{cm}^3\text{mol}^{-1}\text{s}^{-1}) = 10.83 + 3.94 \times 10^{-4}T$ which implies $k = 1.49 \times 10^{11} \text{ cm}^3 \text{ mol}^{-1}\text{s}^{-1}$ at 870K

(corresponding roughly to 600°C). The molar density of CO in a gas stream of 10% CO (by volume) at 870K and 100 kPa is approximately $1.4 \times 10^{-6} \text{ mol/cm}^3$. Now as

$$\frac{-d[\text{CO}]}{dt} = k[\text{CO}][\text{OH}], \quad (2.3.3)$$

the concentration of CO as a function of time in the oxidation reaction can be given as

$$[\text{CO}] = [\text{CO}]^0 \exp (-k[\text{OH}] t) \quad (2.3.4)$$

where $[\text{CO}]^0$ is the initial carbon monoxide concentration. For a hydroxyl concentration of 1 ppm, this implies an 88% conversion of CO in one second, and as the hydroxyl concentration rises to 10 ppm, the 88% conversion rate is achieved in 0.1 second. Thus the carbon monoxide consumption rate is strongly dependent on the

hydroxyl concentration, which in turn is linked to the presence of water vapor. Due to the large amount of bonded water and absorbed moisture in the combustion feedstock, the water vapor concentration will be considerably higher than those amounts considered here. Thus, residence times on the order of tenths of a second are considered to be quite adequate for the total oxidation of carbon monoxide to carbon dioxide.

2.4 DISCUSSION OF CATALYSTS

2.4.1 Introduction

The radwaste solid materials mostly consist of clothes, paper, plastics etc., as indicated by the data provided by Rocky Flats, [1]. These materials are the mixtures of polymers consisting of cellulose, polyolefines, rubbers, and other reactive hydrocarbons with or without sulfur or chlorides (or HCl). Accordingly, when these mixtures are burned (oxidized) with oxygen at 550 - 600°C, the main chemical reactions are quite similar to the thermal or catalytic cracking of polymers or heavy hydrocarbon oil (so-called atmospheric residue oils in oil refinery plants, which is called "Atmos-Residue"). On the other hand, the radwaste liquid fuel oils are similar to diesel fuel. Therefore, both solid and liquid radwastes can be considered as approximately the same kind of heavy hydrocarbons; i.e. "Atmos-Residue oil". In terms of the feedstock property, the radwaste materials are similar to the feedstock for fluid catalytic cracking ("FCC") process in oil industry. Thermally, the radwaste reactor temperature of 550°C is also close to that in the FCC for the coking process in the oil industry.

The fundamental difference between the radwaste burning process and the FCC process is that the radwaste burning process uses a large quantity of oxygen while the FCC process does not use any oxygen. In view of the oxygen usage for the purpose of partial oxidation, the radwaste burning process is similar to the partial oxidation cracking process for the production of olefine from heavy Atmos-Residue (such as BASF's Olefine production process by partial oxidation).

It is apparent the fundamental characteristics of radwaste material burning at 550°C can well be evaluated in the absence of sorbents, [3]. It is apparent that an oxidation catalyst is required. Chromia (Cr_2O_3) is considered to be a very logical and important catalyst for these oxidation processes.

2.4.2 Catalyst Selection

A wide range of materials are conceivable, as oxidation catalysts, but chromia (Cr_2O_3) is one of the most representative and effective compounds. The ionic value of Cr^{+x} can be changed in the range of +3 → +5, which is one of the most typical

behaviors of a transition metal. (Therefore chromium oxides are frequently described as Cr_xO_y .) To further enhance the activity of the Cr_2O_3 , the surface of Alumina (Al_2O_3) particles is impregnated with Cr_2O_3 . This widely used "Chromia-Alumina" catalyst generally consists of 5% - 20% (by weight) Cr_2O_3 and 80%-95% (by weight) Al_2O_3 . This "Chromia-Alumina" catalyst was, and is recommended, for the oxidation catalyst for the radwaste burning in fluidized beds.

Since the mechanical strength of "Chromia-Alumina" catalyst is not high, the catalyst size becomes very small (50 - 100 μm) during use in fluidized beds. Even though efforts to make a physically strong catalyst of "Chromia-Alumina" have been attempted in the past, using relatively coarse "Chromia-Alumina" catalyst particles in fluidized bed reactors has proven to be very difficult. Chromia (Cr_2O_3) must be handled with great care since it is a known carcinogen.

The chemical reaction of the regeneration of "Chromia-Alumina" catalyst depends upon two basic physical parameters. One is to burn out the carbon deposit on the surface of the catalyst and the other is to maintain the special microstructure of Cr_2O_3 on the surface of Al_2O_3 [3]. Therefore, a small amount of catalyst should be circulated to regenerators to maintain the activity of "Chromia-Alumina" catalyst.

2.5 SELECTION OF SORBENT FOR CHLORINE AND SULFUR

As an absorbent of Cl_2 , HCl , and SO_x in process gas, various materials such as CaCO_3 (CaO), BaCO_3 (BaO), Na_2CO_3 (Na_2O), and K_2CO_3 (K_2O) etc. can be considered. Since the reaction temperature is 550°C, CaO is not appropriate. BaCO_3 (BaO) and K_2CO_3 (K_2O) are cost prohibitive. Therefore Na_2CO_3 (Na_2O) is found to be the most promising.

In the temperature range of 550°C - 600°C the Na_2CO_3 (Na_2O) is very reactive with Cl_2 , HCl and SO_x . The vaporization loss in the form of NaCl is also very low at 550 - 600°C, which is favorable. Economically, Na_2CO_3 (Soda Ash) is more expensive than $\text{Na}_2\text{CO}_3\text{-NaHCO}_3$ (Trona). Since the chemical reactivity of Soda Ash and Trona is almost the same, the use of Trona would be the most appropriate. Another advantage of Na_2CO_3 (Na_2O) is that Na_2CO_3 (Na_2O) is also a good catalyst to prevent the formation of carbon at the cracking reaction.

The sulfur will be removed from the gas as Na_2SO_3 (or Na_2SO_4), and the chlorine is fixed as NaCl .

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- [2] Baulch, D.L. and Drysdale D.D., "Combustion and Flame", 23 (215), 1974.

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CHAPTER 3 EVALUATION AND OPTIMIZATION OF THE ORIGINAL SYSTEM

3.1 DESCRIPTION OF THE ORIGINAL SYSTEM

3.1.1 Introduction

The original fluidized bed incineration (FBI) system at Rocky Flats consisted of a two stage combustion system, as shown in Figure 3.1 and described in Reference [1]. This system operated at temperatures below 600°C and below atmospheric pressure. The primary reactor employed a flameless combustion (pyrolysis) process to produce a rich hydrocarbon and carbon monoxide off-gas. This off-gas was then fully combusted (oxidized) in an afterburner with 100% excess oxygen.

The primary reactor was a fluidized bed operating with a lean oxygen content in the fluidizing gas. The partial oxidation of a radwaste was accomplished by using a catalyst, (Cr_2O_3), and the combined fluidized bed material and lean oxygen supply allowed for the operation at 600°C without the formation of hot spots in the reactor. In situ neutralization of the acidic gases formed (mainly chlorides) was accomplished by using a sorbent (Na_2CO_3) in a reducing atmosphere. The primary reactor bed initially started operating with a bed material mixture of 50% catalyst (Cr_2O_3) and 50% sorbent (Na_2CO_3). The particle size of the two materials ranged from 500 to 1000 μm .

Ash and bed material were continuously removed from the gas stream by a cyclone between the two reactors. The gases entering the afterburner appeared to be free of material leading to carbon and/or tar deposits. However, if the oxygen content of the fluidizing gases to the primary reactor was too low, carbon and tar deposits would occur downstream of the primary reactor.

The reactive gases entering the afterburner were combusted in a fluidized bed with an estimated 100% excess oxygen. The afterburner was a fluidized bed of catalyst (Cr_2O_3), with a particle size range of 500 to 1000 μm . The afterburner was water-cooled in order to maintain a maximum temperature under 600°C. The off-gases from the afterburner were passed through a cyclone, a sintered metal filter, a gas cooler, and finally through HEPA filters.

The temperature control for the primary reactor was achieved by varying the amount of oxygen in the fluidizing gases. The temperature control of the afterburner was achieved by varying the feed rate of radwaste into the primary reactor.

3.1.2 Overall Performance

The original FBI system consumed an average of 88 kg/hr of low-level solid radioactive waste or 34 kg/hr of combustible liquids. The oxygen (as air) consumption was approximately 180

kg/hr. A good overall weight reduction ratio of 5:1 to 10:1 was demonstrated by this system. The final system off-gas contained 9.0% oxygen, and 0.0075 to 0.050% carbon monoxide, by mass. The heat released was about 1,600 MJ/hr. Of the thermal energy produced in the afterburner, 45% was removed by the water cooled walls and 55% was removed by the downstream water-cooled heat exchanger.

3.1.3 System Operation

3.1.3.1 Normal operating conditions

Presorted solid waste was fed by a screw feeder through a glove box system to the primary reactor after a two-stage shredding operation. Air and nitrogen were introduced into the reactor, which was filled with 250 kg of sodium carbonate sorbent and 250 kg of chromic oxide catalyst, both in the 500-1000 μm size range. The superficial gas flow velocity was approximately 0.6 m/s, and the gas was fed through distributors with partially shielded holes to prevent solids backflow. The reactor was operated at 550° C well below the 630° C requirement with a slightly negative pressure maintained by downstream gas ejectors. The sodium carbonate-sodium chloride formed an eutectic mixture at this temperature. The primary reactor temperature was controlled by adjusting the relative amounts of nitrogen and air while maintaining the same overall fluidization gas flow rate. Off-gas from the primary reactor passed through a cyclone separator and then entered the afterburner. Additional combustion air (corresponding to 100% excess O_2) was introduced separately through a series of pipe distributors to avoid open flames. The nominally 0.5 m deep bed was charged with about 1350 kg of oxidation catalyst (Cr_2O_3) and was cooled by a water spray on the outside of the afterburner reactor vessel. The off-gas was then passed through a cyclone separator in series with a bank of sintered metal filters. After particulate removal, the off-gas was cooled to approximately 50° C before passing through a bank of HEPA filters. The cooled gas was drawn through an air ejector which maintained a negative pressure throughout the system and was driven by air in a 1:1.2 drive air to process gas flow rate. Ash was removed from the system in the form of fly ash entrained in the off-gas. The fly ash was removed from the two cyclone separators and the sintered metal filters via cycling dumping valves into a water-cooled screw conveyor.

3.1.3.2 Emergency shut-down

No specific information is available on the existing system shutdown procedure but a credible scenario would be the following: in the event of an emergency, the solid or liquid feed to the primary reactor is immediately stopped, the air feed

to both the primary reactor and afterburner is shut off, and only the nitrogen fluidizing gas is allowed to flow to quench the pyrolysis and oxidation reactions. The external cooling water to the afterburner is increased to cool the bed to manageable temperatures. Once temperatures are sufficiently low, the beds are allowed to slump or the solid material is removed by increasing the superficial gas velocities and elutriating the particles into the cyclones.

3.1.4 Start-up and Shut-down

Initial loading of the solids in the two fluidized beds is achieved by pneumatic conveying. The primary reactor was heated with fluidizing gas that has been preheated by two 80 kW electric heaters. After the bed temperature reached 300 - 350° C, the waste feed was started and the heaters shut off. The afterburner was heated to 300° C by a 90 kW air heater, and a methanol-burning, honeycomb platinum catalytic heater was then used to heat the combustion gas stream to 600° C after which the waste feed was started. The exothermic pyrolysis and oxidation reactions were sufficient to maintain the necessary operating temperatures in both beds. Shutdown would involve stopping the waste feed, reducing the oxygen flow rate until the temperature drops, and then slumping the beds after a sufficient cool-down period. Bed material was removed by increasing the (cold) fluidization velocities in the beds and removing the transported particulate matter from the cyclone separators.

3.2 SUGGESTED MODIFICATIONS

While the existing system has been shown to operate reliably and predictably with a wide range of feedstock fuel, it is concluded [1] that modification in a few areas would improve the overall process. The following features of the present system are considered to warrant attention:

(i) The existing system is a "once-through" system as there is no provision for a solids recycle. This results in the fact that about 38% of the FBI residue is either catalyst or sorbent. In addition, the catalyst and sorbent lost are in the finer particle size ranges that have the greatest chemical activity. By recycling the solids that are released from the two beds back to each bed, the consumption of catalyst and sorbent and the total volume of the FBI residue will be drastically reduced. In addition, solids recycling will reduce the effect of variation of the fluidization hydrodynamics on the overall system performance and increase the overall solids residence time. The retention of fines in the primary reactor and the afterburner will also have the effect of reducing the overall rate of particle attrition, through their "lubricating" effect on the larger particle collisions.

(ii) The existing system uses catalyst and sorbent material as the fluidizing medium in the primary reactor. It is anticipated that using sand or even silica spheres as a bed medium will reduce the attrition of catalyst and sorbent, extend catalyst and sorbent life and reduce the overall production of fines in the system.

(iii) The third major modification to consider is the conversion from an air/nitrogen gas system to a carbon-dioxide/oxygen fluidizing gas mixture. The use of recycled CO_2 as a fluidizing gas and O_2 variation for temperature control will reduce the total volume of off-gas that requires further treatment.

3.3 MODIFIED SYSTEM DESIGN

3.3.1 Option A

The original system has apparently functioned well; therefore, suggestions for the modification of the system were rather modest. The obvious change to consider was the recirculation of solids and gases. The recirculation of solids would lead to more complete utilization of catalysts and sorbents and a reduction of carbon in the residue. The recirculation of product gas for fluidization would reduce the consumption of gas required by the system. In addition, the use of sand as a useful component of the bed material was considered.

The advantages of mixing solid and liquid radwaste are such that it is assumed in a modified system, the solid and liquid wastes will be injected simultaneously. Hopefully, the mix of solid and liquid wastes would be such that the heating value of the radwaste feed will not have a great variation. Also, the make-up catalyst and sorbent should be considered as part of the radwaste feed.

As stipulated for this study, CO_2 was to be the fluidizing gas, instead of air and nitrogen. Also stipulated for this study was an evaluation of this distributor plate design.

The proposed modification of the original system is shown in Figure 3.2. The use of CO_2 instead of N_2 as a carrier gas has advantages in treating the off-gases from this system. The use of CO_2 leads to a simplification in the recirculation of the gases through the afterburner, as the gas in the afterburner system is essentially CO_2 . The utilization of CO_2 as the fluidizing gas in both the primary reactor and the afterburner will allow for lower fluidization velocities and still maintain the same fluidization characteristics. Subsequently, the erosion problems may be reduced, but the reduced velocities may not have a significant effect on elutriation. The attrition of in-bed particles may also be reduced. The flow characteristics around the modified existing system are described in Table 3.2.

The solids elutriated from the primary reactor contain

catalyst, sorbent, and refuse containing carbon. Thus it is appropriate to recycle these solids back to the primary reactor. The solids elutriated from the afterburner should essentially be catalyst which should be recycled to the afterburner. Note that a two-stage cyclone system is recommended. The first cyclone is to remove the larger particles from the gas stream and recycle them back to the reactors while the second cyclone is to remove the finer particles. The second cyclone thus removes the fine spent sorbent, finely attrited catalyst, and the ash as waste and produces a relatively clean gas stream for the downstream ejectors. Although it may be desirable to keep fines in a bubbling bed, it is believed an equilibrium quantity of fines will remain in the two fluidized beds.

According to [1], most of the solid refuse in the original system exited the system by way of the first cyclone. Rough estimates of the freeboard height indicated that an additional two feet (.066 meter) in height may significantly reduce the solids loading in the cyclone.

The addition of sand to the beds should not cause problems unless the sand accelerates the attrition process. Since, the catalyst and sorbent apparently did an excellent job in breaking-down the hydrocarbons which would lead to carbon and/or tar deposits downstream of the reactor, an experimental study must be performed to determine the amount of sand that may be used in the primary bed, as a function of oxygen concentration, which will produce an acceptable off-gas.

The modified system should essentially operate in the same manner as the existing system. However, the liquid and solid radwaste would be mixed with an experimentally determined amount of catalyst and sorbent. If sand is to be employed as a bed material component, then the required sand replacement would also be added to the radwaste feed. The amount of O_2 fed into the primary reactor will control the temperature in that reactor. The radwaste-feed rate will be controlled in order to maintain the desired temperature in the afterburner.

Start-up and shut-down procedures are not affected by the proposed modifications. The emergency shut-down should also remain the same.

The distributor plate in the existing system represents a component that may be improved. The existing system is shown in Figure 3.2, reference [1]. This configuration consists basically of two 45° plates leading down to a relatively large, flat, rectangular area. Apparently the withdrawal screw is placed over this area. It appears that a relatively large "dead" region may exist in this area; and subsequently, residue containing carbon may be removed, along with non-combustibles, catalyst, and sorbent.

A preliminary design of a new distributor plate is shown in Figure 3.3 and Figure 3.4. There are three or more rings which have holes with protective covers. The holes in each ring are

such that the gas flow is essentially uniform, and the pressure drop between the plenum chamber and the height of the bed at the hold location is uniform. At the center of the distributor plate is a drainage port as shown in Figure 3.3. The drainage port may be emptied intermittently. The waste stream may then go to a separator which may separate the catalyst and sorbent ($d_p < 1000\mu\text{m}$) from the larger tramp metal or other non-combustibles. The smaller catalyst and sorbent particles may then be recycled back to the primary reactor. The preliminary design of the new distributor system shows a rather steep (45°) funnel shape. This geometry was selected so that the primary reactor could be readily drained. The geometry of the holes in each ring is such that the gas experiences the same pressure drop across the bed.

3.3.2 Option B

A major goal of this study is to develop a system to minimize the flow of off-gases that must be treated downstream of the incineration system. Recycling the off-gases significantly reduces the net amount of off-gases from the incinerator. However, high pressure CO_2 is required to drive the injectors which control the recirculation of off-gases. As a potential solution to this problem, it is suggested that an additional stream of gas from the settling chamber be used for the injectors. This gas would first be filtered and then sent to a compressor. The compressed gases would then flow to an accumulator and be distributed as necessary. This system, Option B, is shown in Figure 3.5 and the flow quantities are listed in Tables 3.3. An obvious extension of option B is to also recycle off-gases to the primary reactor. However, a significant amount of water vapor would be recycled. Since concern has been expressed about the water vapor content in the primary reactor, these extensions have not been considered in detail.

3.4 CONCLUSIONS

The original incineration system, developed for low-level radwaste, produced good results. The system was simple and easy to control. The off-gases contained essentially no carbon monoxide. The mass reduction ratio varied from 5:1 to 10:1. The once through flow of gases led to 2351 kg/hr of off-gas per 82 kg/hr of radwaste (28.67 kg/kg). The ratio of O_2 in the off-gas to radwaste was 2.58 kg/kg.

The modified systems recirculate a portion of the off-gases, which reduces the amount of fluidizing gas and oxygen required by the incineration system. The recirculation of solids reduces the amount of make-up catalyst and sorbent required and reduces the amount of solid waste produced by the system. For option A, the amount of off-gas is 1124 kg/hr and ratio of off-gas per radwaste is 13.7 kg/kg. The ratio of O_2 in the off-gas to radwaste is

0.35 kg/kg.

Employing the option B, then the amount of gases recycled is increased, and the ratio of off-gas per radwaste is 10.8 kg/kg. The ratio of O₂ in the off-gas to radwaste is 0.26 kg/kg. The off-gases are again restricted to that part of the system associated with the afterburner.

The modified systems, options A and B, retain the simple concepts of temperature control and safety of the original system. Because there is no specified steam or off-gas production, the two reactors are not strictly connected, thus there are no significant stability problems. Although the compressor may cause some concern about the reliability of the option B system, stand-by compressed CO₂ could be available to convert the system from option B to option A. The modified systems would also produce a very low level of carbon monoxide. There appear to be no items requiring research which would cause these modifications to be found unacceptable. The basic problems would concern the cyclones and how to select the materials for recycling and for waste disposal.

Unfortunately, it was difficult to estimate temperatures, pressures, and heat transfer throughout these systems based on the data in [1].

References

- [1] Meile, L.J., et al "Rocky Flats Plant Fluidized Bed Incinerator" RFP - 3249, UC - 70 Nuclear Waste Management, DOW/TIC - 4500, March 1982.

Existing System

[illegible]

Notes 1) All mass flow rates are in kg/hr . 3) Afterburner $U_o = 0.6 \text{ m/sec}$

Modified System
Option B

Selected Process Data

[illegible]

Notes 1) All mass flow rates are in kg/hr • * Avoids condensation
2) T and P are in °C and kPa

Figure 3.1 Schematic of Existing System

Figure 3.2 Schematic of Proposed Modification of the Existing System - Option A

Figure 3.4 Proposed Distributor Plate Design

Figure 3.5 Schematic of Proposed Modification of the Existing System - Option B

CHAPTER 4

PROPOSED HYBRID REACTOR CONCEPT

4.1 DESIGN OBJECTIVES

The objective is to develop a fluidized hybrid bed reactor for optimum performance. This means minimum off-gas quantities, minimum operating and maintenance costs, and simple and reliable pressure and temperature control. The radwaste input is variable but averages 1.5 million BTU/hr and consists of 88 kg/hr of combustible fuel. The carbon comprises 63.6% by weight along with hydrogen (7.5%) and oxygen (23.9%).

Inside the first bubbling bed reactor, the 56 kg/hr carbon is converted to carbon monoxide requiring 75 kg/hr of oxygen. The 7 kg/hr hydrogen is oxidized to H_2O requiring 53 kg/hr of oxygen. Since the fuel supplies 25 kg/hr of oxygen the net oxygen flow required to the first reactor is 103 kg/hr.

In the second fast bed reactor, the 131 kg/hr of carbon monoxide is oxidized to carbon dioxide requiring an additional 75 kg/hr of oxygen. To assure complete combustion, excess oxygen in the amount of an additional 15 kg/hr is supplied to the second fast reactor. This excess oxygen is discharged together with the combustion products and recycled carbon dioxide in the off-gas. Thus the total stoichiometric oxygen required by the fuel at minimum capacity is $75 + 53 - 25 + 75 = 178$ kg/hr, and the oxygen supplied is $178 + 15 = 193$ kg/hr. When the 5 kg/hr H_2O is included in the combustion process parameters, then one deals with $5 + 56 + 25 + 7 = 93$ kg/hr feed and $93 + (75 + 53 - 25 + 75) = 271$ kg/hr combustion products in the off-gas plus 15 kg/hr of excess oxygen in the off-gas. Although there is only 15 kg/hr of excess oxygen in the off-gas, there is 36 kg/hr of excess oxygen leaving at the outlet of the second fast bed reactor. The reason for this high free oxygen content is the high internal recirculation rate which produces a much higher mass flow rate inside the fast bed reactor compared to the off-gas flow rate. The overall process contains $36/178 = 20\%$ excess oxygen at the outlet of the second bed fast reactor. The remaining component in the off-gas is the recycled CO_2 which is pressurized and used as ejector driving fluid. A large fluidization flow rate is obtained by pressurizing and recirculating both gases and solids, using ejectors driven by compressed carbon-dioxide. This procedure mimics and makes dual use of the carbon-dioxide which must be recycled for fluidization purposes. This gas is separated by condensation or freezing and then be compressed for recycling. Obviously designing the system so as to minimize the recycled ejector driving gas CO_2 will minimize installation and operational costs.

4.2 PRELIMINARY CALCULATIONS

The proposed hybrid reactor must meet a number of design criteria. Firstly, the solids must be oxidized largely to CO in the first stage of the reactor, and CO must be oxidized fully to CO₂ using excess oxygen in the second stage. The reactors must have sufficient volume for the necessary residence time to achieve the reaction. This volume can be inferred from prior empirical evidence or through reaction kinetics calculation. Secondly, the design must obey heat and mass balances. Thirdly, the hydrodynamic properties of the reactor must be examined. Both reactor geometry and gas throughput will affect the regime and quality of fluidization.

It was assumed that the bed material would consist of a blend of both sodium carbonate (or substitute) and chromic oxide and alumina mix (or substitute) catalysts, since in a single vessel hybrid reactor solids cannot readily be separated by species. The material was assumed to have a density of 2500 kg/m³ and be 500 to 1000 μm in size. For purposes of fluid bed regime estimation, the fluidizing gas was taken to be CO₂ at 600° C and atmospheric pressure.

Dimensionless diagrams are available, Figure 1.5 illustrates a regime map in terms of U* vs. d*,

$$U^* = U_0 \left[\frac{\rho_g^2}{\mu (\rho_s - \rho_g) g} \right]^{1/3} \quad (4.2.1)$$

$$d^* = d \left[\frac{\rho_g (\rho_s - \rho_g) g}{\mu^2} \right] \quad (4.2.2)$$

For this operation, take

$$\begin{aligned} \rho_s &= 2500 \text{ kg/m}^3 \\ \rho_g &= 0.62 \text{ kg/m}^3 \\ \mu_g &= 3.58 * 10^{-5} \text{ kg/msec} \\ g &= 9.81 \text{ m/sec}^2 \end{aligned}$$

from the Handbook of Chemistry and Physics, [1], using linear interpolation.

Hence,

$$U^* = U \left[\frac{(0.62)^2}{3.58 * 10^{-5} * 2500 * 9.81} \right]^{1/3} \quad (4.2.3)$$

so that

$$U^* = 0.76U$$

and

$$d^* = d \left[\frac{0.62 * 2500 * 9.81}{(3.58 * 10^{-5})^2} \right]^{1/3} \quad (4.2.4)$$

or

$$d^* = 2.28 * 10^4 d \quad (4.2.5)$$

From Figures 1.5 and 1.6 taking particle sizes of 500 μm ($d^* = 11.40$) to 1 mm ($d^* = 22.80$), one concludes that a reasonable range of superficial velocities to maintain a bubbling bed is 0.53 m/sec to 2.50 m/sec. For a high velocity turbulent bed with 500 to 1000 μm particles to operate, velocities of 2.6 to 8.6 m/sec would be typical, although higher velocities would lead only to a reacting pneumatic transport flow.

As a check on these diagrams, the approach of Turton et al. [2], was used to find the terminal velocities, assuming the particles were rough rather than smooth spheres. For the terminal velocity,

$$\frac{1}{U_t^*} = \left(\frac{18}{d^{*2}} \right) + \left(\frac{1}{d^*} \right)^{0.5}, \quad (4.2.6)$$

so that for 500 μm spherical particles $U_t = 5.48$ m/sec. Rough particles will have terminal velocities up to 30% lower. Minimum fluidizing velocities were found using the Ergun equation as approximated by Wen and Yu as presented by Geldart [3].

$$U_{mf} = \frac{\mu}{\rho_g d} \{ (1,135 + 0.0408 Ar)^{1/2} - 33.7 \} \quad (4.2.7)$$

where

$$Ar = \frac{\rho_g d^3 (\rho_s - \rho_g) g}{\mu^2} \quad (4.2.8)$$

For a 500 μm particle, $U_{\text{mf}} = 0.10$ m/sec, and for a 1000 μm particle, $U_{\text{mf}} = 0.38$ m/sec. Minimum bubbling should occur at slightly higher fluidizing velocities following some bed expansion.

The basic hybrid design calls for a fluid bed acting as a gasifier and pyrolyzer in the lower part, and an excess oxygen afterburner in the upper part. Since circulation is common in most beds and mixing and channeling are difficult to predict, a staged bed system was chosen, with the upper bed being a fast or turbulent bed and the lower bed a bubbling bed. There would be one distributor at the base of the lower bed and additional gas injection at the base of the upper bed, so that solids mixing between beds would be minimal. Higher velocities needed for fast fluidization in the upper bed are attained through reduction in cross-section and the injection of additional gas.

4.3 PROPOSED CONFIGURATION

The proposed hybrid reactor configuration is based on Rocky Flats' experience using two reactors, one with only 40% of stoichiometric oxygen content for pyrolysis and hydrochloric acid neutralization and the other for complete combustion using 100% excess oxygen, [4]. Based on their success, we propose a similar 0.69 m diameter, 1.8 m high, primary bubbling bed reactor with a fluidization velocity of about 0.6 m/s. Stacked on top of this bed is a 0.46 m diameter second reactor (fast bed) with a fluidization velocity of 4m/sec and a minimum height of 1.8 m. In this way, the lower bed is in a bubbling mode, while the upper bed is in a high velocity fluidization mode. Solids are discharged along with the gas at the top of the upper bed.

Each bed is provided with a solids feed from its own cyclone. The ejectors are designed to maintain a cyclone pressure equal to that at the base of the bed to which the solids are returned. This pressure equalization permits the solids to be returned by gravity without significant gas transfer between the beds and cyclones. The gas recirculation rate is controlled separately by another ejector. Cooling of the fast bed gas recirculation loop provides fine control of the bed temperature and also improves the ejector performance. Ejector driving pressures are controlled by throttling to achieve the desired recirculation rates, pressure levels and fluidization velocities at each station. Solids composition is the same throughout the two-bed system.

The design incorporates a high recirculation rate for all solids in an attempt to achieve complete combustion, neutralization of the off-gas and good temperature uniformity. The gas recirculation rates are adjusted to achieve the desired fluidization velocity. The solids recirculation rates depend on the particle size and fluidization velocity. The maximum pressure in the reactor is chosen slightly below atmospheric.

The maximum pressure in the system occurs at the first bubbling bed injector (denoted as station (1), Figure 4.1). For simplicity this peak value is set equal to ambient pressure or 0 kPa vacuum. All other station pressures are computed relative to station (1) and their vacuum level is reported in (kPa) pressure units.

To minimize the mass flow rate of CO₂ needing to be recycled and discharged, this high pressure gas source is not wasted on direct fluidization injection into the reactor, but only used as driving fluid for the ejectors which pressurize the recirculated gases. The oxygen is also available at high pressure and when possible should also be used as an ejector driving fluid. However some must be injected directly into the reactors for oxygen level control to regulate the temperature.

4.4 PROPOSED PRESSURE LEVELS AND EJECTOR PRESSURE RISE

In Figure 4.1 is shown a schematic of the hybrid system. The bubbling (27" diameter) reactor has the fast (18" diameter) reactor stacked above. Two cyclones in series are shown with valves permitting the first cyclone to recirculate outlet gases and solids mainly through the fast reactor and the second cyclone to recirculate outlet gases and solids mainly through the bubbling reactor. All pressure levels tabulated are in -kPa vacuum relative to station (1), which is at the highest system pressure level. Station (1) is upstream of the injector of the bubbling bed reactor. The calculated pressure levels at all stations are shown in Fig. 4.2, including the pressure rise needed in each ejector. Note that none of the station pressure levels has a vacuum level in excess of 25 kPa.

All ejectors are driven by either compressed carbon dioxide or oxygen at an assumed pressure level of 1000 kPa (10 atmospheres) and thus all ejectors are designed to operate on a primary to secondary pressure ratio of 10. The high pressure selected reduces the required mass flow rate of the recycled CO₂. The ejectors supply pressures are throttled to maintain the desired pressure levels for optimum recirculation. The required driving jet supply pressure depends on the pressure rise ΔP needed to offset the pressure drop through all other system components.

- a) First bubbling reactor has an estimated bed pressure drop of 10 kPa (about 1 m water) and a distributor pressure drop of 10 kPa.
- b) Second fast reactor has an estimated bed pressure drop of 5 kPa and distributor pressure drop of 3 kPa.
- c) Both cyclones are designed for a vortex nozzle pressure drop of 1 kPa, and a thus their gas and solids outlet pressure is 1kPa lower than at the inlet.
- d) The first cyclone bottom outlet pressure is maintained equal to that at the base of the fast bed reactor so that solids

- will freely return by gravity with negligible gas transfer. This requires a 6 Kpa pressure rise ejector at the inlet to the first cyclone.
- e) The second cyclone bottom outlet pressure is maintained equal to that at the base of the bubbling reactor so that solids will freely return by gravity with negligible gas transfer. This requires a 11 kPa pressure rise ejector at the inlet to the second cyclone. Note that the cyclone nozzle pressure drop was estimated to be 1 kPa.
 - f) Most of the fluidization gas needed by the bubbling bed reactor is taken from the outlet of the second cyclone. It must therefore be compressed 10 kPa in an ejector placed at the 1st bed injector inlet. This is downstream of the water cooling section in the return loop, for more efficient ejector action.
 - g) Almost half of the fluidization gas required by the fast reactor is taken from the outlet of the first cyclone and must be compressed by 3 kPa to overcome the injector pressure loss.
 - h) The remaining oxygen needed to control the fast bed O_2 level at 100% excess oxygen is provided by direct injection.

4.5 PROPOSED RECIRCULATION FRACTIONS NEEDED FOR O_2 LEVEL CONTROL

The 93 kg/hr of radwaste fuel supplied to the first reactor requires 50 kg/hr of oxygen for 40% of stoichiometric combustion. The remaining 60% or (75 kg/hr of oxygen) is needed for stoichiometric burning. However in the second (fast) reactor the oxygen level is controlled to 100% excess so that 150 kg/hr oxygen flow rate is required. Thus 75 kg/hr excess oxygen discharged in the off-gas coming from the second cyclone. The composition of the off-gas is the same as that in the first reactor recirculation loop.

The first loop recirculation rate is limited such that the first reactor does not receive more than the desired 50 kg/hr oxygen flow rate. The design shown uses a 23 kg/hr oxygen level in the recirculation loop and has the remaining 27 kg/hr of oxygen supplied by direct injection. The system mass outflow rate must equal the mass inflow rate, thus the maximum pressure in the system can be controlled by varying the off-gas flow rate. The off-gas consists of all the gaseous combustion products generated by the 93 kg/hr radwaste fuel including 5 kg/hr moisture and 125 kg/hr stoichiometric oxygen for a total of 218 kg/hr combustion products. In addition there is the 75 kg/hr excess oxygen used and the 618 kg/hr recycled carbon-dioxide needed as primary flow to drive the ejectors. Note that the mass fractions in the recirculation loop to the bubbling reactor are the same as that of the off-gas. The amount of mass flow needed to support a 0.6 m/sec fluidization velocity can be calculated as follows: In the first reactor with cross sectional area 0.37 m^2

and gas density of 0.6 kg/m^3 the flowrate must be 472 kg/hr .

In the second (fast) reactor with a fluidization velocity of 4 m/s and gas, density of 0.6 kg/m^3 and cross-sectional area of 0.164 m^2 the flow must be 1408 kg/hr .

The computed mass balance shown in Table 4.1 is based on an ejector nozzle supply pressure of 10^4 kPa using either recycled CO_2 or required combustion O_2 . The mass flow rates of major gas components such as: fuel, oxygen, combustion products including fuel, recycled CO_2 and total flow rates are shown in Fig 4.3.

For recirculation rate control purposes, it is best to throttle ejector (13) jet supply pressure. Each cyclone exit pressure should be controlled to equal that at the base of the respective bed to which the solids are to be returned. Such a procedure reduces gas leakage through the solids return loop and allows using a larger tube. The design velocity in the bubbling bed is 0.6 m/s at a total flow rate of 472 kg/hr , and in the fast bed at 4 m/s and a total flow rate 1408 kg/hr . Return loop 1 has 276 kg/hr and loop 2 has 763 kg/hr . The first (bubbling) bed has an equivalence ratio of 2.5 with only 50 kg/hr of oxygen supplied. This oxygen supply is adjusted to control the bubbling bed temperature. The second fast bed has an oxygen flow rate of 186 kg/hr resulting in an equivalence ratio 0.4, to assure complete combustion. Note that this results in 75 kg/hr excess oxygen escaping with the off-gas. The required total oxygen supply rate is 200 kg/hr of which 125 kg/hr is needed for complete combustion and 75 kg/hr is the excess oxygen.

The total CO_2 to be recirculated through the ejectors is 618 kg/hr . By driving one of the ejectors with oxygen, one can minimize the CO_2 content in the off-gas which must be recycled. The externally recycled CO_2 is compressed and thus serves a dual purpose as an ejector driving fluid and as a fluidization fluid. The total off-gas flow rate is 911 kg/hr (consisting of: 75 kg/hr excess O_2 , 218 kg/hr combustion products and 618 kg/hr CO_2 to be recirculated). The recirculation rates are controlled regulating the various pressure levels. The bubbling bed temperature is controlled by varying the bubbling bed oxygen injector supply. While the fast bed temperature control will require active external cooling. The radwaste feed rate may also be used as a temperature control mechanism. In the absence of exterior insulation the wall radiation losses can be as high as 10^6 BTU/hr . The purpose of designing the incinerator with two return loops, each with a high gaseous recirculation rate, is for the following reasons:

- 1) achieve complete combustion
- 2) maintain the system temperature as uniform as possible
- 3) minimize the amount of CO_2 required to be recycled.

4.6 SOLIDS MOVEMENT

For a full understanding of reactor operation, the solids

movement in the system must be modeled. This will govern the choice of cyclones and is necessary in determining reactor height and two-phase mixture density in the fast section of the hybrid reactor.

The following fluid bed model is proposed. The bubbling bed section will throw solids into the freeboard of the bed where the vessel is still 0.69 m in diameter. Section 4.2 has shown that the fluidizing velocity of 0.6 m/sec is far lower than the terminal velocity of a 500 μm particle in this bed, so that all the 500 μm particles should fall back into the bed after a sufficient transport disengagement height (TDH) (Zenz, [5]; Geldart, [3]). If the freeboard is lower than this height, 500 μm solids will be carried into the fast bed section. It is assumed in this analysis that all particles entering the fast section will remain entrained and will be expelled from the top of the fast section and sent to the cyclone. The diagram of Zenz [5] may be used to show that the TDH for 500 μm particles in the lower bed is 3.31 m (for 1000 μm particles it is 2.54 m). Since true composition of the bed, including fines, is not known, the calculation proceeds assuming 500 μm particle size.

Wen and Chen [6] offer a scheme for predicting the particle ejection rate E_o into the bed freeboard. The rate at which 500 μm solids enter the upper bed would decline exponentially up to the TDH after which (in theory), no solids of 500 μm would leave the lower bed. In practice fines will always leave the bed and may carry some larger particles over with them.

If the freeboard height in the lower bed is far lower than the TDH, then one may assume as an approximation that all particles ejected from the surface of the lower bed pass through to the upper (fast) bed. The ejection rate of solids can be found from the equation of Wen and Chen. Hence the entrainment flux of particles is $E_o = 1.85 \text{ kg/m}^2\text{s}$, so that the mass rate $m_o = 1.23 \text{ kg/s}$. Since the gas mass flow rate in the fast bed is 0.39 kg/s (=1408 kg/hr), this represents a solids-to-gas mass flow rate of 3.14. However, the in-situ solids-to-gas mass ratio will be far higher due to slip between the phases. Results of Canada et al. [7] for 650 μm particles suggest that the upper bed will be turbulent since $U > 0.65U_t$. Few data are available (except for fine cracking catalyst) to predict the precise solids holdup in the bed, but even in the limit of fast fluid beds the volumetric solids fraction varies from 0.2 to 0.4 at the base to 0.02 to 0.05 at the exit. Solids residence time in this bed would therefore be on the order of a minute. In turbulent beds the solids fraction is yet higher, leading to the conclusions that

- (a) adequate catalyst will be present in the upper bed.
- (b) excessive height of the upper bed should be avoided to avoid high pressure drop.

4.7 EJECTOR EQUATIONS RELATING MASS FRACTIONS TO COMPRESSION RATIO

A brief description of ejectors is presented in Appendix B which serves as an introduction to the following material. For preliminary design of this incinerator it is more convenient to use some of the empirical equations suggested in these reports. These were obtained by curve fit comparison with several large scale commercially available ejectors. These are all iso-energetic air-to-air or steam-to-steam ejectors with an equation for mass flow ratio m_i required (m_i = ejector outlet over secondary entrained mass flow rate). The required input data are:

- a) the primary flow to secondary flow total pressure ratio R (this is 10 in the present as the ejector primary or driving fluid is assumed to be available at 1000 kPa (10 atmospheres) and the component pressures are all at about 1 bar.
- b) The desired compression ratio (r) is given by pressure at the outlet divided by pressure at the inlet. In this incinerator this is approximately $r = 1 + \Delta P/10^5$ where ΔP is in Pa. The empirical equation has two coefficients C which depend on the geometry such as: constant area ejector with $C_1 = 0.830$ and $C_2 = 0.753$. For the second throat type ejector find $C_1 = 0.912$ and $C_2 = 0.809$. The second throat ejector is more efficient than the constant area ejector but less suitable for varying flow rates such as occur in the incinerator type operation.

The required iso-energetic mass flow ratio is given by:

$$1/m_i = \{1 - \ln(r)/[C_i\{\ln(R)\}^{C_2}]\}^{2.6} \quad (4.7.1)$$

For non-iso-energetic ejectors, the differences in molecular weight and total temperature modify the required mass flow ratio m by the ratio of the speed of sound of both the primary driving fluid a_p and that of the entrained secondary fluid a_s , or:

The actual mass flow ratio m required is:

$$m = 1 + (m_i - 1)(a_s/a_p) \quad \text{and} \quad a = (\gamma R_{\text{universal}} T/M)^{1/2} \quad (4.7.2)$$

where M is molecular weight.

For the proposed hybrid incinerator assume the compressed O_2 and CO_2 to be supplied at 1000 kPa thus $R=10$ and due to heat transfer the primary flow temperature T_p is about half that of the entrained gas and thus $a_s/a_p = 1.4$.

Example calculations for constant area ejector with $R=10$:

$$\begin{array}{llll} \Delta P = 0.25 \times 10^4 \text{ Pa} & \text{or} & r = 1.025 & m_i = 1.042 \quad \text{and} \quad m = 1.059 \\ \Delta P = 0.5 \times 10^4 \text{ Pa} & \text{or} & r = 1.05 & m_i = 1.086 \quad \text{and} \quad m = 1.120 \end{array}$$

$$\begin{array}{llll} \Delta P = 0.6 \times 10^4 \text{ Pa} & \text{or } r = 1.06 & m_i = 1.104 & \text{and } m = 1.146 \\ \Delta P = 1.1 \times 10^4 \text{ Pa} & \text{or } r = 1.11 & m_i = 1.198 & \text{and } m = 1.277 \end{array}$$

4.8 ENERGY CONSIDERATIONS

4.8.1 Background

It is necessary to evaluate the heat released by the combustion in both the lower and upper reactors and to compare this with the ability to transfer heat through the bed walls. Temperature control is essential to maintain the system below 600° C, yet maintain a sufficiently high temperature of 575° C used in the heat transfer calculations below.

We will assume that 88 kg/hr of fuel is fed to the combustor. For total combustion, using a heating value of 7,779 BTU/lb., this represents a heat release of 1.5 million BTU/hr.

4.8.2 Heat Transfer Coefficients

Bed-to-surface heat transfer coefficients cannot be predicted with absolute certainty, but for the lower (bubbling) bed, one may employ the dimensionally inconsistent equation of Zadbrosky [8] using SI units

$$h_{\max} = 35.8 \rho_g^{0.2} k_g^{0.6} d_p^{-0.36} \quad (4.8.1)$$

where k_g is the gas thermal conductivity and $h_{\max}$ is the highest value of heat transfer coefficient expected. Taking $\rho_g = 2500 \text{ kg/m}^3$, $d_p = 0.0005 \text{ m}$ (ie 500 μm) and $k_g = 0.038 \text{ W/mK}$ (for CO_2), then that $h_{\max} = 371 \text{ W/m}^2\text{K}$. Bock and Schweinzer, [9], measured values for h ranging from 100 to 160 $\text{W/m}^2\text{K}$, in a bed of 970 μm particles, while Carlomagno et al., [10], confirm values of around 250 $\text{W/m}^2\text{K}$ for 500 μm particles.

There are little data on large particle high velocity beds, but Guigon et al., [11], have shown coefficients of 200-300 $\text{W/m}^2\text{K}$ for fast circulating beds of 260 μm particles. A conservative approach would be to use 200 $\text{W/m}^2\text{K}$ for fast circulating beds of 260 μm particles. The approach adopted here is, therefore, to use 200 $\text{W/m}^2\text{K}$ for both beds in the hybrid design.

4.8.3 Heat Production

The total heat produced must be divided between the lower and upper sections for preliminary design purposes. As limited by oxygen delivery, 40% of the oxidation occurs in the lower vessel, so that the remaining 60% must occur in the upper turbulent bed.

Calculation shows that 88 kg of feed consists of 55.9 kg C, 24.6 kg O, and 6.5 kg H. One kilogram of feed consists of other heteroatoms. Complete oxidation to CO_2 and H_2O requires 176.4 kg

of additional oxygen. Forty percent of this quantity is therefore 70.6 kg O₂, added to the first bed. This is capable of partially oxidizing (ie to CO and H₂O) 69.3% of the bed feed of 87 kg combustibles.

It is therefore assumed that the upper reactor will produce heat from:

- (a) the combustion of 30.7% of the feed, or 463,000 BTU/hr
- (b) the combustion of the CO produced by partial combustion of 69.3% of the material in the bubbling bed. This corresponds to 90 kg/hr of CO, with a heat release of 4,600 BTU/lb of CO, yielding a total of 910,000 BTU/lb.

One concludes that the vast bulk of the heat, namely 1,370,000 BTU/hr of 1,500,000 BTU/hr is released in the upper bed. The lower bed therefore supplies only 130,000 BTU/hr of heat generation.

4.8.4 Heat Loss - Lower Bed

Let us assume that the lower bed loses heat to the walls primarily by convection ($h = 200 \text{ W/m}^2\text{K}$) and that the uninsulated vessel walls (Temperature T_w) in turn radiate this heat energy to the surroundings which are at a significantly lower temperature. Therefore,

$$\epsilon \sigma T_w^4 = h (575 - T_w) \quad (4.8.2)$$

where $\sigma = 5.67 \times 10^{-8} \text{ W/m}^2\text{K}^4 \quad (4.8.3)$

and $\epsilon = 0.6 \quad (4.8.4)$

so that $T_w = 511^\circ \text{ C}$ and heat flux is $12,800 \text{ W/m}^2$. For a cylindrical vessel of 0.69 m diameter and 1.8 m height, neglecting end areas, the total possible heat loss is 49,900 W, or 170,370 BTU/hr, which is in excess of the heat generated. Insulation will be required for thermal control of this bed.

In addition, cold solids are fed to this bed at a rate of 88 kg/hr and water at 5 kg/hr. To bring these feeds from ambient conditions to 575° C requires an energy delivery of 26,888 W (91,770 BTU/hr) and 4,964 W (16,941 BTU/hr) respectively.

Since nearly all of the energy released by combustion in the lower bed is required to heat the incoming solids, the bed will need to be insulated. In addition, the feed gases and returned catalyst will need to be heated to near operating temperature, and an auxiliary fuel may be required if a less energetic solid is fed. Temperature control can be accomplished by regulating gas and returned catalyst feed temperatures.

4.8.5 Heat Loss - Upper Bed:

Following the same argument used for the lower bed, one

finds that for $T_w = 487^\circ\text{C}$, heat loss from the bed if the walls are uninsulated is 33,280 W (156,200 BTU/hr) which is insufficient to dissipate the heat of combustion.

However, this bed is fed with recycle gas and oxygen. There is a potential to absorb about 150,000 W by feeding these gases at 27°C temperature and causing the bed to heat them to operating temperature.

By utilizing cool gas feed to the upper bed and by permitting radiation from the uninsulated vessel, approximately 50% of the heat generated may be dissipated. The remainder must be removed by adding a cooling system to the bed, or by recycling some of the solids from the cyclone to the upper bed rather than the lower bed. If the solids leave the bed at 575°C , are externally cooled, and returned at, say 200°C , then a solids flow rate of 2,208 kg/hr (0.61 kg/s) will be needed to remove the remaining heat of reaction (220,800 W). This assumes that the solids are 50% alumina, 50% sodium carbonate for specific heat of (0.96 kJ/kgK).

A further recommendation is that the percentage of stoichiometric oxygen supplied to the lower bed be raised to 50%, shifting more heat release to the lower bed while still maintaining reducing conditions in that bed.

4.9 CONCLUSIONS

A hybrid reactor, consisting of a fast, oxygen-rich bed, on top of a bubbling, fuel rich bed has been proposed. Cyclones are used to separate the transported solids from the off-gas. By using ejectors in the design, acceptable pressure and mass balances are achieved, and the beds can be fluidized primarily by the off-gas. This initial study certainly indicates that the hybrid bed concept is feasible and reliable.

Adjustments to the mass flow rates or pressure levels inside the system can be readily achieved. As long as one satisfies gaseous mass flow rate continuity at each junction, one can recompute the new system local flow rates.

4.10 DISCUSSION

Preliminary design calculations have demonstrated the feasibility of the hybrid reactor concept, but detailed analysis with experimental research will be required for a precise, optimized design. Firstly, the reactor overall sizing requires a knowledge of the reaction rates of the radwaste feed into the bubbling bed, and of the combustible gases and oxygen catalyzed in the upper bed. Without such data, there is a danger that the bed will be oversized in the interests of maintaining a conservative design. Secondly, the equilibrium solids size distribution remains unknown, but this will have a profound effect on the system hydrodynamics. A monosize material of 500

μm was assumed in the calculations, but the actual bed inventory will be determined by a balance between the attrition rate solids makeup rate, and solids removal rate, coupled with bed elutriation and cyclone performance. A finer inventory will increase the solids flow rate in the upper bed and cyclones. Thirdly, the stable balance of pressures, particularly between the hybrid bed and the cyclone/return pipe combination requires further examination, preferably by using an experimental cold model.

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Table 4.1 Selected Process Data for Hybrid System

Location	M _{O₂}	M _{CO}	M _{H₂O}	M _{total}	M _{CO₂}	T	P	Additional Remarks
1	103	0	21	459	335	600	0	
2	103	0	21	552	335	600	-10	93 kg/hr fuel added
3	0	131	86	552	335	600	-20	
4	111	0	157	1408	1150	600	-25	
5	36	0	157	1618	1425	600	-19	
6	20	0	86	1205	1099	600	-9	
7	15	0	65	904	824	600	-10	
8	5	0	21	301	275	600	-10	
9	10	0	71	729	648	600	-10	
10	0	0	0	60	60	400	+10Atm	Pressures adjustable
11	0	0	0	32	32	400	+10Atm	for ejector driving
12	0	0	0	210	210	400	+10Atm	fluid.
13	0	0	0	316	316	400	+10Atm	
14	98	0	0	98	0	400	0	
15	95	0	0	95	0	400	0	

Note
 1) All mass flow rates are in kg/hr
 2) T and P are in °C and kPa gage

Figure 4.1 Schematic of Proposed Hybrid System

Figure 4.2 Vacuum Levels (-kPa) Inside Components and Ejector Pressure Rise ΔP in kPa

Figure 4.3 Gas Component Mass Flow Rates in kg/hr

CHAPTER 5 PROPOSED FAST CIRCULATING REACTOR CONCEPT

5.1 INTRODUCTION

This chapter presents an investigation of the fast circulating reactor (FCR) concept as a solution to the radwaste volume reduction problem. Because the FCR concept is different than the existing system, a discussion of potential problems will be described. A general description of the proposed FCR concept will then be presented along with a discussion of solutions to potential problems. Mass and energy analyses are presented for the optimum FCR concept.

The basic problem for the combustion of the radwaste is the heavy hydrocarbons, which have a tendency to produce significant carbon deposits at the desired operating temperatures. Apparently, the presence of the catalyst and sorbent in the present system has a rather strong mitigating effect on the carbon deposition process. An additional problem is that the solid radwaste material injected into the reactor will be rather large, flat, particles relative to the bed material. While in the present system, these particles represent a relatively small portion of the bed material, in the FCR concept, these particles may represent a significant portion of the material in the distributor region. Finally, the existing system has two distinct reaction chambers, the primary reactor and the afterburner. An FCR system contains one reaction chamber which must create a distributed heat generation zone to maintain temperature limitations and produce a minimum of intermediate products of combustion.

5.2 PROPOSED FAST CIRCULATING REACTOR DESIGN

The proposed FCR design is shown in Figure 5.1. The radwaste is injected in the FCR at location 8 and mixes with the gases from locations 1 and 2. Solid material dropping down through the FCR will form a bed of material on a spouted-bed type distributor plate design. This design should allow for good fluidization of the larger flat particles until they are reduced to a particle size suitable for entrainment. If the liquid radwaste is not mixed with the solid radwaste, it is recommended that the liquid be sprayed onto the bed, as shown in Figure 5.1.

In order to "stretch-out" the reaction zone and avoid creating hot spots ($T > 600^{\circ}\text{C}$), oxygen is introduced into the FCR at several locations, 1, 2, 3, and 4. The oxygen introduced at these locations is not only to control the combustion process, but also the carbon deposits on the internal surfaces of the FCR system. Thus the oxygen feed locations are very important design parameters. It should be noted that CO_2 gas is also injected with the O_2 to provide for the correct amount of gases for the desired

fluidization characteristics.

The solids circulating in this system consist of radwaste, catalyst, and sorbent. The catalyst, Cr_2O_3 , is necessary for a rapid chemical reaction at the relatively low temperatures ($T < 600^\circ\text{C}$) in this system. Again the sorbent Na_2CO_3 is necessary to remove chlorine and sulfur compounds. It should be noted that during sustained operation, there may be a build up of carbon on the surfaces of the catalyst. Consequently, the catalyst may require regeneration periodically. The Na_2CO_3 will convert to NaCl and Na_2SO_3 , which should be removed from the system. The attrition of the sodium compounds should provide additional surface area for additional absorption of chlorine and sulfur.

5.3 BASIC DESIGN

The basic design geometry constraint was a six-inch diameter for the FCR. However, this dimension together with the feed rate condition for the radwaste leads to an undesirable design. The required oxygen flow leads to very high velocities which may cause unacceptable residence times and erosion. Thus calculations were also completed for a ten-inch and twelve-inch diameter bed. Superficial velocities of 8-50 m/sec have been calculated.

The solids to gas mass flow rates are to be controlled in the range of from 2 to 5, which is common in gas-solid two phase flow systems. The catalyst and sorbent particles will be in size range of 50-100 μm . This size range has been used successfully in systems with stable operation and no significant problems with erosion and attrition. The spent sorbent, radwaste ash (fines), and catalyst fines are removed at location 31, whereas, the heavier radwaste inerts are removed at 6.

5.4 MASS AND HEAT BALANCES OF PROPOSED PROCESSES - CASE STUDY

5.4.1 Data Used for the Calculation

The calculation was carried out based upon the overall process described in Figure 5.1

The waste fuels are solid and liquid phases. In the following calculation, only the case of waste solid (mixtures of plastics, clothes, paper, etc.) is considered. (Similar calculation can be done for the case of waste liquid).

The chemical and heating values were assumed according to the data of current system, they are shown in Table 5.1.

5.4.2 Material Balance

The material balance is calculated based on the following approach. The approach is described below, together with the assumptions and the description of the combustion process.

- 1) Combustion Temperature: 550°C
- 2) Feed Rate of Fuels (ash free): 87 kg/hr
- 3) Heat Release Rate (required): 1,600,000 kJ/hr
or 382,000 kcal/hr
- 4) As the oxidation catalyst, Cr₂O₃ will be used.
- 5) The combustion products were assumed as CO₂ and H₂O. In fact, at relatively low combustion temperature, the C-CO-CO₂-O₂ Chemical equilibrium reaches mostly to CO₂. CO/CO₂ ratio becomes much higher when temperature is above 1,000°C. Therefore, we assume the combustion products are CO₂ and H₂O substantially.
- 6) The mass flow of the process is shown in Figure 5.2. The fuel being fed at the rate of W_F (Kg/hr) will be burned by the feed gas containing oxygen (10%, 15%, 18% or 21%) in V_{GF} (V_{GF} = V_j+V_g+V_{s1}+V_{s2}), where the excessive oxygen ratio (λ) will be changed in the range of 1.8, 1.2 and 1.0. In this study, the oxygen concentration is the same at each injection point. However, this need not be the case in future studies of this concept. The quantity λ is defined as:

$$\lambda = \frac{V_{O_2, \text{Actual}}}{V_{O_2, \text{Theor.}}} \quad (5.4.2.1)$$

For the fuel molecule of C₃H₄O (defined in Table 5.1) and feed rate of 87 kg/hr, the stoichiometric equation (in the unit of Kmole/hr) can be written as:

$$4.66 \text{ C} + 3.11 \text{ H}_2 + 0.78 \text{ O}_2 + (5.43*\lambda) \text{ O}_2 + [5.43*\lambda*(1-\text{Conc.}_{O_2})/\text{Conc.}_{O_2}] \text{ CO}_2 \\ = 4.66 \text{ CO}_2 + 3.11 \text{ H}_2\text{O} + [5.43*(\lambda-1)] \text{ O}_2 + [5.43*\lambda*(1-\text{Conc.}_{O_2})/\text{Conc.}_{O_2}] \text{ CO}_2 \quad (5.4.2.2)$$

where Conc._{O₂} is the concentration of O₂ (volume fraction of O₂) in input gas. The theoretical (stoichiometric) oxygen volume flow rate per mass of radwaste is:

$$V_{O_2, \text{Theor.}} = [C]*22.4/12 + ([H]/2 - [O]/16)*(22.4/2) \quad (5.4.2.3)$$

where [C], [H] and [O] indicate weight percentage of C, H and O in fuel. For the fuel shown in Table 5.1:

$$V_{O_2, \text{Theor.}} = 1.40 \text{ Nm}^3/\text{kg-fuel}$$

The total feed gas volume flow rate V_{GF} (Nm³/hr) is:

$$V_{GF} = W_F * V_{O_2, \text{Theor.}} * \lambda / \text{Conc.}_{O_2} \quad (5.4.2.4)$$

Let the total feed gas mass flow rate be W_{GF} (kg/hr) then:

$$W_{GF} = V_{GF} * \rho_{GF} \quad (\text{Kg/hr}) \quad (5.4.2.5)$$

where ρ_{GF} is the bulk density of the feed gas. The feed gas V_{GF} is fed to the fluidized bed/through V_j , V_g , V_{s_1} and V_{s_2} .

7) The circulation of solid particles is tentatively considered as 35,000 kg/hr for a 12 inch diameter bed [3].

The solid/gas mass flow rate (G_s/G_g) in the bed is assumed in the range of 1 - 10, where G_s and G_g have the unit of kg/m²sec. The 90% of the carried-over particles from the bed will be collected in the first cyclone and the remaining particles in gas will be collected in the second cyclone.

8) The exit gas flow rate, V_{GE} (Nm³/hr) can be described as

$$V_{GE} = ([C] * \frac{22.4}{12} + [H] * \frac{22.4}{2}) * W_F + (1 - \text{Conc.}_{O_2}) * V_{GF} + (\text{Conc.}_{O_2} * \frac{\lambda - 1}{\lambda}) * V_{GF}$$

(5.4.2)

and $W_{GH} = \rho_{GH} * V_{GE}$ (kg/hr)

where ρ_{GH} is the density of exit gas (kg/Nm³)

9) The combustion conditions were chosen for several cases. The excessive oxygen ratios were $\lambda = 1.0, 1.2$ and 1.8 . The O_2 concentrations were, $\text{Conc.}_{O_2} = 10\%, 15\%, 18\%$ or 21% . Based upon the approach described above, the material balance calculations were accomplished for several cases, as shown in Table 5.2. Variables V_{GF} , W_{GF} , V_{GE} , W_{GE} , λ and Conc._{O_2} are described earlier. The gas velocity U_0 with 6", 10" and 12" fluidized bed is shown in Table 5.3.

5.4.3 Heat Balance

Based upon the material balance shown in Table 5.2 and technical data used for heat balance calculations, in Table 5.4, the heat balance was calculated for four cases, as shown in Table 5.5.

The schematic diagram for the heat balance is shown in Figure 5.3, together with some heat transfer data. To obtain a stable operation, the heat generated by the chemical reactions must be equal to the energy absorbed by the flow and heat lost due to radiation and convection at the desired temperatures. The difference between the heat generated and the heat disposed is the heat unbalance ΔQ . This term must be zero in the actual final system design. For this preliminary concept study, the ΔQ was calculated to determine how close each set of selected conditions was to $\Delta Q = \text{zero}$. For the most representative cases, a heat balance was calculated as shown in Table 5.4.

5.4.4 Fluidization Conditions and Selection of Bed Diameters

For cases of A1, B1 and C2, the gas velocity in fluidized bed is shown in Table 5.6. Since gas flows V_{s_1} and V_{s_2} are injected at the middle and top of the fluidized bed, the gas

velocity are different at the lower portion, middle portion and the top of the fluidized bed. Assume $V_{s1} = V_{s2} = 0.25 \cdot V_{up}$, the gas velocity in lower, upper and top of the fluidized bed are shown in Table 5.6.

The size of the recycling particles (mixture of Cr_2O_3 , Na_2O , ash and recycling carbon) is assumed to be in the range of 50 - 80 μm . This particle range corresponds to "FCC" catalyst, which is widely utilized in oil industry. This particle range has been demonstrated to produce a very stabilized fluidization process and has a minimum of problems of attrition and erosion. These fluidization conditions facilitate the control of the circulation rate of solid particles. The circulation rate (35,000 kg/hr for a 12 inch bed) of Cr_2O_3 , Na_2O and carbon can be accomplished.

The terminal velocity, U_t , is in the range of 0.2 - 0.4 m/sec for the expected bed material. The choking phenomenon will never occur in this design because the terminal velocity of particles is negligible to the gas velocity.

5.5 PRESSURE AND TEMPERATURE IN THE VARIOUS LOCATIONS

The pressure and temperature of various locations, for example, P_0 , P_1 , P_2 , P_3 & P_4 , and T_0 , T_1 , T_2 , T_{c1} & T_{c2} , are shown in Figure 5.4.

5.6 PROCESS CONTROL

5.6.1 Combustion Temperature

The combustion temperature in the fluidized bed should be kept at 550 - 600°C. The temperature can be controlled by:

- a) radwaste feed rate.
- b) the O_2 concentration, $Conc_{O_2}$.
- c) the O_2 ratio λ .

5.6.2 Heat Balance

The heat balance ($\Delta Q=0$) can be controlled by:

- a) Q_2 is determined by the volume flow rate and temperature of the inlet gas.
- b) Q_3 , the heat loss by radiation and convection. With proper insulation of the cyclone and downcomer, the heat loss can be reduced.

5.6.3 Fluidization Condition

When the fluidization particle size is in the range of 50 - 80 μ m, the fluidization condition is generally stable for a wide range of gas velocities (1-15 m/sec). There would be negligible attrition and erosion problems, if the velocity is under 15 m/sec, as experienced in the oil industry (FCC process).

5.6.4 Pressure

The pressure can be maintained by the combination of ejector(s) and blowers.

5.7 OPERATING STRATEGY - EMERGENCY SHUT DOWN, START UP AND ORDINARY SHUT DOWN

5.7.1 Emergency Shut-down

- a) Shut down O₂ feed.
- b) Stop the flow of the fuel.
- c) If necessary, stop or reduce the flow of all the inert fluidizing gas (CO₂).
- d) Keep the bed material in the reactor while feeding a small amount of inert gas (CO₂) through the system.
- e) After sufficient cooling, the bed material can be withdrawn by opening the pipe used for gas jet injection.

5.7.2 Start-up

- a) Feed char or coke powder into the reactor.
- b) Gradually heat-up to 550°C under packed bed conditions and then fluidize char powder at low gas velocity (< 0.3m/sec).
- c) Gradually begin the flow of radwaste into system.

5.7.3 Regular Shut-down

- a) Stop the flow of the O₂ gas.
- b) Stop the flow of the fuel.
- c) Feed in only a very small amount of inert gas (CO₂)
- d) Unit can be maintained under these conditions for few days and can easily be restarted by just feeding appropriate amount of O₂ and inert gas (CO₂).

5.8 CONCLUSION

Based upon the consideration and process design on the fast circulating reactor for radwaste burning process, the following conclusions were obtained.

1. The FCR should use fine particles such as 30-100 μ m, which are almost identical to FCC catalyst particles. The FCC process is one of the most reliable fluidized bed reactor. Therefore, since the design of the system is based on the experience from the FCC process, the risk is minimized.
2. The quantities: gas velocity (U_0), oxygen ratio (λ), oxygen concentration Conc._{O₂} and system geometry are important in developing the mass and heat balances of the system.
3. From the series of mass and heat balance calculations of various operating cases, some appropriate operating conditions were identified as follows:
 - 1) Excess oxygen ratio $\lambda=1.2-1.0$
 - 2) Oxygen concentration at the bottom of the reactor Conc._{O₂}=18-21%
 - 3) Gas velocity $U_0=4-9$ m/sec when 12" I.D. circular equivalent FB is used.

Note that the following assumptions were made:

- a) The system is not thermally insulated, and the heat loss (mostly radiation) is to the surrounding environment.
 - b) The reaction rates of combustion are assumed to be very high.
4. Based upon the proceeding considerations, a flow sheet was prepared as shown in Figure 5.1.
 5. The usage of a spouted fluidized bed is highly recommended. This system can improve the internal circulation flow of solid particles in the fluidized bed. Tramp material can also be easily withdrawn from the system by the jet gas opening. Another advantage is that the liquid radwaste (oils) can efficiently be burned by this fluidization method.
 6. The chromia-alumina catalyst will be very useful to burn the radwaste at low temperature (500-600°C). The regenerator should be used to maintain the reactivity of this catalyst.
 7. The reactivity of absorbent (for Cl₂ or HCl) should also be maintained by using the regenerator. The NaCl should be removed from the surface of Na₂O (Na₂CO₃).

5.9 NOMENCLATURE

A_T	Total surface area for heat release
$[C]$	Weight percentage of C in fuel
$Conc_{O_2}$	The concentration of oxygen (volume fraction of O_2 in input gas).
G_g	Gas mass flow rate in fluidized bed (kg/m^2sec)
G_s	Solid mass flow rate in fluidized bed (kg/m^2sec)
h	Overall heat transfer coefficient ($kcal/(m^2*^{\circ}C*hr)$)
h_1	Heat transfer coefficient for convection ($kcal/(m^2*^{\circ}C*hr)$)
h_r	Heat transfer coefficient for radiation ($kcal/(m^2*^{\circ}C*hr)$)
$[H]$	Weight percentage of H in fuel
$[O]$	Weight percentage of O in fuel
P_o	Pressure of input gas (kKPa)
P_1	Pressure at the bottom of the fluidized bed (kPa)
P_2	Pressure at the top of the fluidized bed (kPa)
P_3	Pressure at the transfer line between two cyclones (kPa)
P_4	Pressure of exit gas (kPa)
Q_1	Heat required to heat up the fuel (kcal/hr)
Q_2	Heat required to heat up the input gas (kcal/hr)
Q_4	Heat loss due to radiation and convection (kcal/hr)
Q_R	Reaction heat (kcal/hr)
T_o	Temperature of input gas ($^{\circ}C$)
T_1	Temperature at the bottom of the fluidized bed ($^{\circ}C$)
T_2	Temperature at the top of the fluidized bed ($^{\circ}C$)
T_{c1}	Temperature at the cyclone #1 ($^{\circ}C$)
T_{c2}	Temperature at the cyclone #2 ($^{\circ}C$)
U_o	Gas velocity
V_g	Feeding gas volume flow rate at the bottom of the fluidized bed through the funnel shaped gas distributor (Nm^3/hr)
V_{GII}	Exit gas volume flow rate (Nm^3/hr)
V_{GIF}	Total feeding and recycled gas volume flow rate (Nm^3/hr)
V_j	Feeding gas jet volume rate at the bottom center of the fluidized bed (Nm^3/hr)
$V_{O_2, Theor.}$	Oxygen volume required per 1 kg fuel, theoretical value ($Nm^3/kg-fuel$)
$V_{O_2, Actual}$	Oxygen volume required per 1 kg fuel, actual value used for the system ($Nm^3/Kg-fuel$)
V_{s1}	Feeding gas volume flow rate at the middle of the fluidized bed (Nm^3/hr)
V_{s2}	Feeding gas volume flow rate at the top of the fluidized bed (Nm^3/hr)
$W_{circulate1}$	Particle circulation rate through the cyclone #1 (kg/hr)
$W_{circulate2}$	Particle circulation rate through the cyclone #2 (kg/hr)
$W_{circulateT}$	Total particle circulation rate (kg/hr)
W_F	Fuel mass flow rate (kg/hr)
W_g	Feed gas mass flow rate at the bottom of the fluidized bed through the funnel shaped gas distributor (kg/hr)
W_{GII}	Exit gas mass flow rate (kg/hr)
W_{GIF}	Total feeding and recycled gas mass flow rate (kg/hr)

W_j Feeding gas jet mass flow rate at the bottom center of the fluidized bed (kg/hr)
 W_{s_1} Feeding gas mass flow rate at the middle of the fluidized bed (kg/hr)
 W_{s_2} Feeding gas mass flow rate at the top of the fluidized bed (kg/hr)
 λ Excessive oxygen ratio
 ρ_{OP} Bulk density of the feed gas (kg/Nm³)
 ρ_{OE} Bulk density of the exit gas (kg/Nm³)
 ΔQ Difference between reaction heat and total heat loss (kcal/hr)
 ΔT Temperature difference (°C)

References

- [1] Kono, H. et al., Hydrocarbon Processing, p. 121 - 126, Nov. 1970.
- [2] Kono, H., Hydrocarbon Processing, p. 123 - 129, Jan. 1980.
- [3] Yerushalmi, J. et al., Chapter 7 in Fluidization edited by Davidson et al., second edition, Academic Press, London, 1985.

Table 5.1. Fuel Properties

Fuel Molecule	C_3H_4O
Percentage by Weight:	
[C]	64.4%
[H]	7.1%
[O]	28.5%
High Heating Value H_h	≈ 4400 kcal

Table 5.2. Mass Balance for Various Cases

Case	λ	Conc. O_2	W_F (kg/hr)	V_{GF} (Nm ³ /hr)	W_{GF} (kg/hr)	V_{GE} (Nm ³ /hr)	W_{GE} (kg/hr)
A1	1.8	21%	87	1,044	1,934	1,096	2,021
A2	1.8	15%	87	1,462	2,754	1,514	2,841
A3	1.8	10%	87	2,193	4,190	2,245	4,277
B1	1.2	21%	87	696	1,289	748	1,376
B2	1.2	15%	87	975	1,836	1,027	1,923
B3	1.2	10%	87	1,462	2,793	1,514	2,880
C1	1.0	15%	87	812	1,530	864	1,617
C2	1.0	18%	87	677	1,264	729	1,351

Table 5.3. Gas Velocity U_0 in Fluidized Bed

Case	Exit Gas (Nm ³ /hr)	Gas Velocity U_0 (m/sec) at 550°C & 94KPa *		
		6" Bed	10" Bed	12" Bed
A1	1,096	53.54	19.27	13.39
A2	1,514	73.94	26.62	18.48
A3	2,245	109.63	39.47	27.41
B1	748	36.54	13.16	9.14
B2	1,027	50.14	18.05	12.53
B3	1,514	73.93	26.62	18.48
C1	864	42.21	15.19	10.55
C2	729	35.60	12.82	8.90

Note: * 550°C and 94 KPa are temperature and pressure at the top of the bed.

Table 5.4. Technical Data Used for Heat Balance Calculation

Temperature in furnace	550°C
Temperature of input gas	30°C
Temperature of environment	30°C
Temperature of recycled particles	550°C
Temperature of fuel	30°C
Heat loss rate due to convection $h_1 \cdot \Delta T$	4,000 kcal/(m ² hr)
Heat loss rate due to radiation $h_2 \cdot \Delta T$	18,000 kcal/(m ² hr)
Specific heat of fuel	0.2 kcal/(kg*°C)
Specific heat of gas (input or exit)	0.14 kcal/(kg*°C)
Size of the fluidized bed:	
Diameter	6" (0.1524 m) or 10" (0.2540 m) or 12" (0.3048 m)
Length	10 m
Surface area for heat release	
6" Bed	4.79 m ²
10" Bed	7.98 m ²
12" Bed	9.58 m ²
Diameter of the downcomer	10"
Total surface area for heat release (including cyclone & downcomer) A_T	12.8 m ²

Table 5.5 Heat Balance for Selected Cases

Case	Q_R kcal/hr	Q_1 kcal/hr	Q_2 kcal/hr	Q_3 kcal/hr	ΔQ kcal/hr	Comment
A1	382,800	9,048	140,765	281,600	-48,613	insulation
B1	382,800	9,048	93,843	281,600	-1,691	balanced
C1	382,800	9,048	111,384	281,600	-19,232	insulation
C2	382,800	9,048	92,028	281,600	124	balanced

Table 5.6. Gas Velocity U_0 in Fluidized Bed by Location

Case	Exit Gas (Nm ³ /hr)	Bed Portion	Gas Velocity U_0 (m/sec) at 550°C		
			6" Bed	10" Bed	12" Bed
A1	1,096	lower	26.77	9.64	6.69
A1	1,096	upper	40.16	14.46	10.04
A1	1,096	top	53.54	19.27	13.39
B1	748	lower	18.27	6.58	4.57
B1	748	upper	27.41	9.87	6.85
B1	748	top	36.54	13.16	9.14
C2	729	lower	17.80	6.41	4.45
C2	729	upper	26.70	9.61	6.67
C2	729	top	35.60	12.82	8.90

Figure 5.1. Proposed Circulating FB System

Note:

$$W_{\text{circulateT}} = W_{\text{circulate1}} + W_{\text{circulate2}} \\ = 5,000 \text{ Kg/hr}$$

$$V_{\text{GF}} = V_j + V_g + V_{s1} + V_{s2}$$

$$W_{\text{GF}} = W_j + W_g + W_{s1} + W_{s2}$$

$$V_{\text{GF}} = \text{Total feed and recycled gas volume flow rate} \quad \text{Nm}^3/\text{hr}$$

$$W_{\text{GF}} = \text{Total feed and recycled gas mass flow rate} \quad \text{Kg/hr}$$

$$V_{\text{GE}} = \text{Exit gas volume flow rate} \quad \text{Nm}^3/\text{hr}$$

$$W_{\text{GE}} = \text{Exit gas mass flow rate} \quad \text{Kg/hr}$$

$$W_F = \text{Fuel mass flow rate} \quad \text{Kg/hr}$$

Figure 5.2. Simplified Mass Flow for Schematic Flow Sheet

Note:

$$\Delta Q = Q_R - (Q_1 + Q_2 + Q_3)$$

$$Q_3 = (h \cdot \Delta T) \cdot A_T$$

$$h \cdot \Delta T = h_1 \cdot \Delta T + h_2 \cdot \Delta T$$

whereby the ambient temperature (30°C) is used
for heat balance calculations

- Q_R reaction heat
- Q_1 heat required to heatup the fuel
- Q_2 heat required to heatup the input gas
- Q_3 heat loss due to radiation and convection
- h overall heat transfer coefficient
- h_1 heat transfer coefficient for convection
- h_2 heat transfer coefficient for radiation

Figure 5.3. Simplified Heat Balance for Schematic Flow Sheet

Pressure (KPa)	Temperature (°C)
$P_0 = 100$	$T_0 = 100$
$P_1 = 99$	$T_1 = 550$
$P_2 = 94$	$T_2 = 550$
$P_3 = 92$	$T_{c1} = 550$
$P_4 = 90$	$T_{c2} = 550$

Figure 5.4. Simplified Pressure and Temperature Diagram of the Process

CHAPTER 6 SUMMARY AND RECOMMENDATIONS

6.1 INTRODUCTION

The following summary and recommendations relate to the preliminary design study to develop an improved combustion system for the incineration of low-level radioactive wastes. Three different configurations are evaluated. The first involved improvements to the original system, while the remaining two are a hybrid fluidized bed system and a fast fluidized bed system.

6.2 SUMMARY

6.2.1 Existing System

The proposed modified system is an adaptation of the original combustion system that includes a primary reactor followed by an afterburner. Modifications to the original two reactor system were designed to improve overall combustion efficiency, reduce carbon and oxidant catalyst in the waste residue, reduce the amount of gas required, and permit the collection and sampling of combustion off gases before release to the atmosphere. Specific changes include:

1. Addition of solids recycle dip legs to the reactor cyclones to increase the solids residence time in the reactors, thereby facilitating improved carbon combustion and reduced sorbent and catalyst consumption.
2. Use of pure oxygen in place of air as the prime combustion gas to eliminate nitrogen in the off-gas.
3. Use of carbon dioxide as a reactor fluidizing gas. The use of carbon dioxide reduces the number of gaseous species and facilitates the off-gas environmental characterization before release to the atmosphere.
4. Mixing of solid and liquid waste to provide a more uniform heating value radwaste feed to reduce wide fluctuations in reactor combustion temperature. Make-up catalyst and sorbent will be supplied with the radwaste feed.
5. Addition of sand to both reactors may provide for a more uniform bed and to further increase the residence time of fine solids.

6. Use of a mixing chamber ejector to provide a positive motive force for movement of entrained solids from the primary reactor and to reduce the overall pressure drop of the system.
7. A new gas distributor to improve the fluidization characteristics of the primary reactor.

The basic characteristics of the original system, simple design and ease of control, are retained in the modified existing system. Improved performance is anticipated with the suggested modifications.

6.2.2 Hybrid Fluidized Bed

The proposed hybrid fluidized bed concept is based on the earlier Rocky Flats' experience using two reactors. In this hybrid concept, the two reactors are stacked as one but each section has its own geometry. The lower part of the reactor is larger in diameter, a bubbling bed, while the upper part of the reactor is smaller in diameter, a fast fluidized bed. Solids, elutriated from the lower bed, pass through the upper bed and are captured by cyclones for return to the system. Many of the modifications proposed for the existing system (previously described in 6.1.1) are incorporated in the hybrid system. The geometry and arrangement of support facilities offer some significant advantages to conventional designs. These include:

1. Ability to extend particle retention time either part of the reactor. The lower part of the reactor is ideal for the gasification of the radwaste and the upper reactor is employed to complete the combustion and optimize sorbent and catalyst usage. The bubbling bed provides for a good combustible solids residence time and for good mixing, while the upper bed provides adequate residence time for the conversion of all carbon containing species to carbon dioxide.
2. Use of two recycle cyclones provides the opportunity to classify solids elutriated from the reactor and may permit the separation and recycle of catalyst independent of reacted sorbent.
3. All of the remaining features for the modified system are utilized, i.e.: oxygen combustion gas, carbon dioxide fluidizing gas, mixing solid and liquid radwaste, mixing ejectors, and gas distribution plate.

4. An advantage over a separate two bed system is that solids and gases are conveyed from the lower to the upper reactor efficiently and with little pressure loss. Solids circulation assists in achieving temperature stability.

The hybrid system design is basic and simple. It offers a wide range of control options and can be easily constructed. Operation at acceptable temperatures can be achieved by insulating the lower (bubbling) bed and by cooling the gases and solids fed to the upper (fast) bed.

6.2.3 Fast Fluidized Bed

The second new system concept is the fast fluidized bed or fast recirculating bed. The configuration for this system is similar to the hybrid reactor except the reactor is of uniform dimension for the lower and upper zones. Further, since the velocities in the fast fluidized bed are to be much higher than in either zone of the hybrid reactor, the overall height of the fast fluidized bed reactor will be greater than for the hybrid reactor. Some of the other features for this system include:

1. The increased height of the reactor accommodates staged combustion along the height of the reactor using an oxidant mixture controller. This provides operating flexibility to control reactor temperature, solids, and liquid flow.
2. The higher reactor velocities offer expanded capabilities to feed solid and liquid waste.
3. As in the hybrid system, use of the two recycle cyclones provides the opportunity to classify solids elutriated from the reactor and may permit the separation and recycle of catalyst independent of reacted sorbent.
4. All of the remaining features for the modified system are utilized, i.e.: oxygen combustion gas, carbon dioxide fluidizing gas, and mixing ejectors.

The concept and principles of fast fluidized beds have been commercially utilized for many years in fluid catalytic cracking processes for transportation fuel production. These units have a very good operating record.

6.3 EVALUATION OF DESIGN CONSTRAINTS

6.3.1 Introduction

This section responds to the specific performance requirements of the radwaste combustion system, as described in a communication from S. Stiefvater, dated February 4, 1992. These needs may be summarized as:

- Maximum CO₂ production per mass of radwaste
- Minimum O₂ in the off-gas stream.
- Reduce catalyst and sorbent in solid refuse and
- Minimize bed diameter.

In addition, it was stated in discussions with Rocky Flats personnel that no CO was to be present in the off-gas stream. The questions concerning the off-gas composition are all interrelated and are discussed in section 6.2.2. Reduction of catalyst and sorbent in the solid refuse from the system will be discussed in section 6.3.2. It was concluded that the minimum bed diameter requirement is no longer significant during the April 24, 1992 meeting in Colorado. The removal of tramp metal was discussed during the January 22, 1992 meeting in West Virginia. Subsequently, a discussion of this problem, in the final report, was requested.

6.3.2 Off-Gas Stream Composition

The need to maximize the CO₂ concentration in the off-gas stream may also be stated as the carbon in the solid refuse flow must be a minimum and that the carbon monoxide in the off-gas must also be a minimum (zero). This need essentially dictates that all the carbon and carbon monoxide in the system be in intimate contact with sufficient oxygen under suitable reaction conditions (temperature, and catalyst). Boilers, and other combustion systems, frequently operate with excess oxygen (excess air) in order to consume all of the fuel. Subsequently, if the fuel composition was fixed, excess oxygen for complete combustion would be required. Also as the fuel composition varies from barrel to barrel sufficient excess oxygen must be present for a rapid increase in carbon content in the fuel feed. This leads to an additional requirement of significant excess O₂ in the system. If there is a rapid decrease in fuel carbon to the system, then an increase of oxygen in the off-gas will obviously occur.

At this time, it is very difficult to identify which proposed system would have a minimum oxygen flow in the off-gas. Experimental models of each system would be required to determine the minimum acceptable excess O₂ required. These models could then be modified to further minimize the O₂ requirements. The system model which produced the final minimum oxygen in the off-gas would then be the system of choice. The oxygen recirculated with the recirculating gases reduces the oxygen demand of the

entire system; however, the recirculation rates for each system could be adjusted so that the effect would be the same for all systems.

6.3.3 Reduction of Catalyst and Sorbent in Solid Refuse

The reduction in catalyst and sorbent consumption may be significantly reduced by recirculating the solids in the gas stream. The regeneration of the catalyst may also significantly reduce its consumption. The larger particles are removed from the gas stream and are recirculated into the fluidized bed. These larger particles should consist of catalyst sorbents, ash, and the unconsumed carbon. A second cyclone, downstream of the first cyclone, would remove fines from the gas stream. These fines would consist of attrited catalyst and sorbent as well as powdered ash. Ideally, there would be no carbon released by the second cyclone, which goes to the solid waste stream.

The attrition rates for the catalyst and sorbent are generally unknown. For the two modifications of the original system, these rates are low. For small particles in a fast bed, the attrition rates are also low. The attrition rate for particles moving between a bubbling bed and a fast bed has not been investigated. Again it is difficult to ascertain specific numbers for the catalyst and sorbent consumption. Obviously, the consumption rate will be significantly reduced by recirculating a large portion of the solids.

6.3.4 Tramp metal

The radwaste material being fed in the reactor may contain small metal objects. The disposal of an accumulation of this tramp metal is shown in Figure 3.4 of the text. Also, for the spouted bed distributor plate, the liquid spray pipe may be extracted and the tramp metal then removed. Therefore, it appears that removing tramp metal does not represent a problem.

6.4 COMPARISON OF SYSTEMS

In order to compare one system against another system, a certain degree of confidence must be developed in the analysis. Also, an advantage of one system over another system must be such that the approach may not be used in all the other systems. That is for a system to have an advantage over other systems, that which provides the advantage must be unique to that system. Consequently, the reduction in specific oxygen consumption due to recycling would be common to all systems. Also, the amount of specific carbon dioxide released can be reduced to that generated by combustion and to that required by the ejectors.

The specific¹ feed rate of catalyst and sorbent will depend on attrition rates, consumption rate (for sorbent), and the

cyclone removal efficiency. For the lower fluidization velocities of a bubbling bed, the attrition rate for particles of 500 μ m range should not be high. For the high velocities of a fast recirculating bed, the particles are small. Experience has shown that the small particles have a low attrition rate and produce low erosion rates. However, there appears to be no information to determine the steady state particle size distribution in the hybrid system. Ideally, the large particles would remain in the bubbling bed and the small particles would be

¹ The term specific indicates that the following quantity is based on a unit mass of radwaste.

recirculated through the fast bed. It is again difficult to compare the systems other than to say that the hybrid system represents a greater degree of uncertainty than do the others.

The next consideration is to compare the size of each system required to comply with the project goals. The modified systems essentially have the same dimensions of the original system, plus space for a heat exchanger, a settling tank, and a compressor. The two stage combustion allows for simple control of the off-gases from the afterburner, allowing for no CO in the off-gases. The oxygen to the first reactor controls the heat generation rate in that reactor whereas the heat generation rate in the afterburner is controlled by the radwaste feed rate into the first reactor. As there is no fixed external demand rate for steam or hot gases from this system, the two reactors can essentially be controlled independently.

The fast recirculating reactor concept combines the original two-stage combustion process into a single stage reactor operating as a fast recirculating bed. The recirculating bed must be sufficiently tall so that all of the carbon monoxide has been oxidized to carbon dioxide. In order to expand the length of flow (height) so that the combustion zone is not too concentrated, the oxygen is injected at three heights within the system. The uncertainty with this procedure is that there is no information on the mixing and the resulting combustion of a crossflow jet into a fast moving gas-solid stream. Therefore, the possibility exists that there may be CO in the off-gas for certain operating conditions and during transient conditions. A significant temperature control for this system would be to vary the radwaste feed rate.

The hybrid system represents two beds, intimately connected, acting as one bed. The combustion process is in two-stages, similar to the original system. The temperature may again be controlled in the bubbling bed by controlling the amount of oxygen injected into the bed. The temperature in the fast reactor section may be controlled by varying the radwaste feed rate. The comments about the required heights for the fast section are essentially the same as those for the fast

recirculating reactor. However, in the hybrid system, the fast section is in the turbulent flow regime. The hybrid system represents a significant degree of uncertainty. The stability of two intimately connected beds must be investigated. Also, the steady state particle size distribution, for the best set of operating conditions, will have to be determined. The injection of a jet of oxygen containing gases into a solid-gas flow must be investigated. The process will then dictate how tall the fast section must be for complete oxidation of carbon monoxide.

In conclusion, the modified original system represents good performance with the least uncertainty. The hybrid and fast fluidized systems represent (apparently) simple systems with a relatively high degree of uncertainty.

6.5 RECOMMENDATIONS

In order to arrive at the most desirable design concept, it is apparent that an experimental effort is required to develop some basic information. The basic information required concerns combustion, ejector performance with two-phase flow, erosion, attrition of bed particles, and heat transfer. Gas-solids separation in cyclones also represents some uncertainty; however, more is known about these devices and data on actual conditions should evolve from the experimental projects. It is, therefore, recommended that the following experimental program be undertaken to resolve the uncertainty in the proposed designs.

The combustion problems are rather complex. Initially, "chunks" of radwaste are injected into the reactor. These chunks of radwaste are then shredded by the bed material as they begin a pyrolysis process. This process rate depends strongly on the catalyst (and perhaps the sorbent) present and the local temperature. The hydrogen released probably reacts quickly with the oxygen present. The hydrocarbons released are partially converted to carbon monoxide and lighter hydrocarbons due to the presence of the catalyst. In addition, fines containing carbon may also be produced. Information relative to the rate of formation of the pyrolysis products is useful for the modified and hybrid system and essential in the fast recirculating system.

The pyrolysis products then enter a reactor where excess oxygen and the catalyst are present. This is the afterburner in the modified system and represents no problem. In the hybrid and fast recirculating beds, this process involves mixing the oxygen and combustibles in a gas-solid flow. The reaction rates, in the presence of the catalysts, for the production of carbon dioxide, are essential information for these fast beds.

The experimental project for combustion in the hybrid and fast circulating bed requires a cold and hot model of each system. The fast section in the hybrid system has different initial (inlet) conditions than the fast recirculating system; therefore, two models should be constructed. The cold models are

to be used primarily for flow studies and low temperature gas mixing studies. The cold hybrid model may also be employed for bed flow stability problems and for steady state particle size distribution studies. The hot models will ultimately allow one to predict the conditions necessary for all the carbon monoxide to be consumed before leaving the system.

Ejectors play a key role in the systems proposed in this study. However, very little appears to be known about the performance of ejectors when the secondary flow is "dusty". The first concern is the efficiency of the system and the second concern is the potential for erosion. Although the required pressure change across the ejector may not be large, the required primary flow may become significantly greater or the required pressure may become much higher. Both effects would lead to a less efficient system. Therefore, it is recommended that a study be conducted to determine an ejector performance as a function of the solids loading in the secondary stream.

The potential for erosion basically exists in the ejectors, jet exit ports, and at rapid changes in flow areas. The erosion rate may be estimated from data obtained during the combustion studies and the ejector studies. The attrition of particles in the hydrodynamics performance of the hybrid system is an important phenomenon, but must be part of the combustion study for this system. The effect of attrition on sorbent and catalyst performance is also an important problem in the combustion study for both systems. Heat transfer coefficients may be measured as part of the hot combustion studies. Concepts for controlling the heat transfer should be developed based on the experimental results. The concepts developed during the experimental study should be important in start-up, operating, and shut-down for each system.

Although there is available information on cyclones, it is recommended that an attempt be made to use cyclones to segregate catalyst and useful sorbent from sorbent fines and ash. The catalyst may then be regenerated to remove any inhibitors (heavy hydrocarbons) on the surface. Although this concept is in its early stages, a successful regeneration process may have a significant economic impact on the performance of the proposed system.

APPENDIX A
THE CONSULTANTS REPORT

**CONCEPTUAL DESIGN OF A FLUIDIZED BED COMBUSTOR
FOR VOLUME REDUCTION OF WASTE
CONSULTANTS REPORT AND COMMENTS**

APRIL 14, 1992

PREPARED BY: T.E. STRINGFELLOW

APPENDIX A

CONSULTANTS REPORT

GENERAL

As a result of the meetings held on the campus of West Virginia University, on April 9 - 10, 1992, Stringfellow Energy Consultants (SEC) was requested to provide comments on the materials presented at these meetings. The following comments and suggestions are based on the combined 35 years of experience in designing, operating, and testing fluidized bed combustion systems.

One general comment should be noted on the ability to model the combustion system. During the early stages of the meeting a statement was made that the combustion system could not be modeled, due to the relatively low temperature, the catalyst, and sorbents used. We believe that this system can be modeled to show off-gas constituents, ash, and combustion efficiency. A model similar to this is already in use at our office. We would require a proximate, ultimate, ash mineral, sorbent, and catalyst analyses to perform in the model.

Another point is that all three designs can be provided with a close approximation of material balance. Issues such as solids loading, and solids distribution can be calculated. If the designers wish to pursue this approach, SEC would be interested in assisting with this effort.

PROPOSED MODIFIED SYSTEM

The modified system proposed is an adaptation of the existing combustion system that includes a primary reactor, followed by an afterburner. From appearances, the proposed changes would include a mixing chamber eductor, a new air distributor, and the addition of CO₂ and O₂ as the prime fluidizing/combustion gas. Another major change is the recycling of solids back to the primary reactor and afterburner systems.

The use of a compression screw feeder would be utilized as the primary waste feed device. There was no indication of how the liquid waste would be fed, nor was there an indication of how bed make up (catalyst and sorbent) would be introduced. We would recommend that these issues be resolved early in the next phase, perhaps using equipment similar to that shown in sketch #1 and sketch #2.

This design can be modeled using heat and material balance calculations found in the boiler and industrial arena. We would recommend that this approach be adopted as soon as possible. We would further recommend that the cyclones and eductor design be optimized, in order to evaluate the this approach in detail.

The air distributor used in the existing design, did not provide good mixing for the material, and was probably the cause of some problems during combustion. We would not recommend the use of an angle bed distributor. A flat plate arrangement, coupled with nozzles that prevent back leakage of material would be preferred. This approach would also provide a positive influence on the fluidization of material, and provide a directional flow for solid waste. A typical nozzle using this approach is shown in sketch #3.

We would not recommend using a mechanical device to remove solids from the bed. Problems with this type of system would be expected due to material erosion of the device, coupled with differential expansion problems between the cylinder rod and the outside casing. A simple loop seal arrangement, seal pot, or bed cooler would provide the efficient removal of material.

PROPOSED HYBRID DESIGN

The proposed Hybrid design offered by the University, is very similar to the attached (sketches #4 & #5) two patents. The use of a bubbling bed, located under a fast bed arrangement offers some advantages to conventional designs. First, this design offers the ability to extend particle retention time, in either part of the combustor. The lower zone is an ideal application for carbon utilization, and the upper zone further adds to the ability to reduce carbon and optimize any sorbent or catalyst usage.

As stated during the meeting, we have had experience with eductors in a high solids concentration atmosphere. The eductors in these cases could not survive the abrasive conditions. We have seen cases where ceramic lined eductors are used to feed coal and limestone, but the reliability is questioned. We would recommend eliminating both eductors (between 5 & 12, and 6 & 13 of figure 4.1a) and adding a larger eductor between 7 & 8. This will provide the system with the required effluent pressure, and reduce the risk of eductor erosion.

The solid refuse lines from the cyclone should be reduced to only one refuse line from cyclone #2. This probably would never be used, but the ability to drain off waste at this point is practical. The solids loop that will probably be utilized is the drain from the bubbling bed. Again, we would recommend that this drain be designed similar to the bed drain in the modified design.

To further enhance the carbon burn out cycle of this system we would recommend that all solids from both cyclones be introduced into the lower chamber. Although the solids will not stay long in this area, it will add heat to the lower zone and reheat the solids as they travel up the shaft of the combustor.

This design in all is basic and simple, it offers a wide range of control options, and can be constructed easily. The remaining system would probably be the same as the modified system. We again conclude that a model can be developed for this system, and that a close proximity of a heat and material balance can be produced.

PROPOSED CFB SYSTEM

The system described in the WVU CFB system is similar to those currently used in the industrial application of Fluidized Bed Boilers. The obvious changes in this design are the use of CO₂, and the gas mixture controller. This design offers expanded capabilities to reduce solid and liquid waste. Further the design offers operating flexibility to control temperature, solids, and liquid flow.

The only concerns we have at this point is the velocity of the lower zone, and the size of the proposed cyclone downcomer. First the fluidizing velocity of the lower zone must be modeled to prevent hydraulic decrepitation of the solids material. Even if 100 micron material is used, this material can severely fracture if the velocity is too low. In shaft type combustors this phenomenon is very common. Essentially we are using a straight shaft combustor in the application.

Second, the cyclone downcomer is too large and gas leakage back up the cyclone leg is a definite possibility. A small downcomer would provide resistance to flow, and can be easily filled with solids to maintain some type of seal. Further we would remove the valves shown in the downcomer lines, and remove the ash discharge point (19). These are not necessary, and only complicate the operation of this simple system.

The gas mixture controller is one of the more unique features of this design. It has the ability to control percentages of CO₂, O₂, and recycle gases in any area of the combustor. To further enhance this control, we would hope that this controller also is utilized as a flow control device, where the flow to each of the four areas can be controlled based on temperature and CO of a given area.

We do not recommend partial combustion/pyrolysis to be utilized throughout the shaft of this system. We believe that this process should occur in the lower zone of the chamber, and the remaining height be used to attempt complete carbon utilization. This will reduce the load on the cyclones, and improve the overall efficiency of the combustion process. Further using the staged air process better utilization of the catalyst and sorbent would be apparent.

COMMENTS AND RECOMMENDATIONS

All of the proposed systems have their advantages and disadvantages, but the Hybrid design and the CFB designs appear to be simple in design, compared to the Modified version of the existing system. Due to the simplicity of these two designs, we believe that the intended goals, and constraints can be better served by the Hybrid and CFB systems.

Each of the three designs can be physically modeled to provide data, and show each individual viability. The costs associated with modeling is expensive, but in order to achieve a proven technology for this application, models must be used. We can recommend an approach that will reduce the time requirements, and costs associated with testing. We would strongly recommend that the participants in this project consider using a Pyrex (or similar material) type of model that can be utilized in a hot or cold phase of testing.

This approach will be able to show the solids flow aspects of each system, and can provide needed data under hot conditions. Waste material for this application, can be obtained from a number of sources. We can closely construct similar materials for combustion to resemble the analyses of the current wastes at Rocky Flats. The data from these types of tests would be the foundation for presenting the concept to the public, and other concerns.

The cost effective approach mentioned above would provide the investigators with significant data, cut research costs, and provide a compact schedule for eventual implementation. Should the investigators wish to proceed along this concept then modeling would be able to begin, once a final design for each system is completed.

COMPRESSION SCREW/LIQUID FEEDER

SKETCH #1

SKETCH #2

DIRECTIONAL FLUIDIZING
NOZZLE

SKETCH #3

United States Patent (19)
Stringfellow et al.

(11) Patent Number: 4,598,653
(45) Date of Patent: Jul. 8, 1986

[54] COMBUSTION SYSTEM FOR BURNING FUEL HAVING VARIOUS PARTICLE SIZES

[75] Inventors Thomas E. Stringfellow, Aurora
Warrie L. Sage, Littleton, both of Colo.

[73] Assignee Stearns Catalytic World Corporation, Denver, Colo.

[21] Appl. No. 755,981

[22] Filed Jul. 17, 1985

Related U.S. Application Data

[63] Continuation-in-part of Ser. No. 649,558, Sep. 12, 1984, Pat. No. 4,565,119

[51] Int. Cl. F23D 1/00

[52] U.S. Cl. 110/347; 110/245

[58] Field of Search 110/263, 245, 347

122, 4 D

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4,395,248 8/1983 Mehta et al. 110/263
4,475,472 10/1984 Adnan et al. 110/347
4,485,747 12/1984 Pershing et al. 110/347

Primary Examiner—Edward G. Favors
Attorney, Agent or Firm—Mark L. Rodgers, E. Eugene Innis, James C. Simmons

ABSTRACT

The present invention is a process and associated apparatus for obtaining energy from a fuel source having particles varying in size from fines to coarse. The fuel source is introduced into a passageway formed between an upper combustor and a lower combustor. A gas stream is moved upwardly through the passageway such that the fine particles are entrained in the gaseous stream and carried into the upper combustor.

16 Claims, 1 Drawing Figure

United States Patent (19)
Sage et al.

(11) Patent Number: 4,565,139
(45) Date of Patent: Jan. 21, 1986

[54] METHOD AND APPARATUS FOR
OBTAINING ENERGY

[75] Inventors: Warnie L. Sage, Littleton, Thomas E.
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[21] Appl. No. 649,558

[22] Filed Sep. 12, 1984

[51] Int. Cl. F23D 1/00

[52] U.S. Cl. 110/247; 110/244

[58] Field of Search 110/245, 110/263, 122/4 D

110/147, 263, 244, 245,

122/4 D

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4,485,747 12/1984 Pershing et al. 110/147

Primary Examiner—Edward G. Favory
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Innis, James C. Simmons

[57] ABSTRACT

A low Btu, low volatile content, low carbon and high ash fuel is combusted in a combined shaft kiln combustor and fluidized bed combustor to generate heat which is converted to useful energy. The fluidized bed combustor is mounted vertically above the shaft kiln combustor. A venturi is provided in a passageway connecting the shaft kiln combustor and the fluidized bed combustor to produce a gaseous stream moving at a velocity sufficient to separate finer particles from the low Btu, low volatile content, low carbon and high ash fuel so that the coarser particles are burned in the shaft kiln combustor and the finer particles are burned in the fluidized bed combustor.

19 Claims, 1 Drawing Figure

SKETCH #5

APPENDIX B

EJECTOR ANALYSIS

APPENDIX B

EJECTOR ANALYSIS

Ejectors had been utilized in developing the designs of the proposed incineration systems. Thus it was appropriate to briefly describe the performance of these devices.

The purpose of an ejector is to move fluid (secondary flow) into a region of higher pressure. A basic ejector design is shown in Figure B.1. A high pressure (primary) gas is accelerated so that when it interacts with the low pressure (secondary) gas, the low pressure gas will be entrained into a region of high pressure. This device is rather simple, with no moving parts, and its efficiency is acceptable considering the simplicity and reliability of the ejector.

In order to quantify the efficiency of an injector, [B.1], its performance is compared to that of an ideal compression system. The secondary flow is isentropically compressed in a compressor driven by an ideal turbine. The ideal turbine is represented by an isentropic expansion of the primary flow. The exit streams from the compressor and turbine are at the ejector exit pressure. For incompressible flow, the compressor power (see Figure B.1 for terminology),

$$\dot{W}_{C,I} = \dot{m}_s [P_e - P_s] \frac{1}{\rho} \quad (\text{B.1})$$

and for the turbine

$$\dot{W}_{P,I} = \dot{m}_p [P_p - P_e] \frac{1}{\rho}. \quad (\text{B.2})$$

These equations assume that the kinetic energy is negligible where the pressures are measured. The isentropic efficiency for the ejector may be written as:

$$\begin{aligned} \eta_I &= \frac{\dot{W}_{C,I}}{\dot{W}_{T,I}} \\ &= \left(\frac{\dot{M}_s}{\dot{M}_p} \right) \frac{[P_e - P_s]}{[P_p - P_e]} \end{aligned} \quad (\text{B.3})$$

2 of 2

Introducing the dimensionless quantities

$$r = \frac{P_e}{P_s} \quad (B.4)$$

$$R = \frac{P_p}{P_s} \quad (B.5)$$

$$m_i = \frac{\dot{m}_e}{\dot{m}_s} \quad (B.6),$$

then one may write

$$\eta_I = \frac{1}{m_i - 1} \frac{(r - 1)}{(R - r)}. \quad (B.7)$$

The mass flow ratio, m_i , can also be expressed as a function of the area ratio $\alpha (= A_1/A^*)$, R , and r . This allows for an equation of the form

$$\eta_I = f(\alpha, R, r) \quad (B.8)$$

The ejector efficiency is also dependent on all the geometric details; however, the diffuser area ratio $\beta (= A_m/A_e)$ is a significant parameter.

Typical ejector efficiencies [B.2] are shown in Figures B.2 and B.3. The efficiencies are shown as a function of $(r - 1)/(R - 1)$ and α . From the known pressures and geometry, one may determine efficiency which in turn leads to the mass flow ratio. The diffuser geometry also plays an important role in the efficiency. Figure B.2 shows the efficiencies for $\beta = 1$ and Figure B.2 shows the efficiencies for $\beta = 0.25$. Ejectors are shown to have a reasonably high efficiency, if designed for a specific condition. The smaller the β parameter, the higher the possible ejector efficiencies.

Empirical formulas have been developed for the performance of ejectors with compressible flow. The quantity R is either known or may be estimated. The quantity r is based on the pressure drops in the system. The dimensionless flow ratio m_i may be calculated from

$$\frac{1}{m_i} = \{1 - \ln(r)/c_1 [\ln(R)]^{c_2}\}^{2.6}, \quad (B.9)$$

if the flow is isoenergetic.

The constants C_1 and C_2 depend on the ejector geometry. For the optimum value of α , use:

for $\beta = 1$, $C_1 = 0.830$ and $C_2 = 0.753$
 $\beta = 0.25$, $C_1 = 0.912$ and $C_2 = 0.808$.

If the flow is non-isoenergetic, then the actual mass flow ratio is

$$m = 1 + (m_i - 1) \left(\frac{a_s}{a_p} \right)$$

where

$$a = \left[\frac{\alpha RT}{M} \right]^{1/2}.$$

This brief description of ejector performance is only intended to serve as an introduction to the subject. Additional performance data may be found in [B.2 - B.5] and in the literature of those who market ejectors.

References

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$$\alpha = \frac{A^*}{A_1}, \beta = \frac{A_m}{A_e}, A^* \text{ Nozzle Throat Area}$$

Figure B.1 Basic Design and Terminology for an Ejector

$$\frac{(P_{oe} - P_{os})}{(P_{op} - P_{os})} = \frac{r - 1}{R - 1}$$

Figure B.2 Ejector Efficiency for a Constant
Area Flow, $\beta = 1$

Figure B.3 Ejector Efficiency for an Ejector
With a Diffuser, $\beta = 0.25$

APPENDIX C

RESPONSE TO THE TECHNICAL COMMENTS OF THE REVIEWERS

APPENDIX C

RESPONSE TO THE TECHNICAL COMMENTS OF THE REVIEWERS

It is unfortunate that each concept group used a different solid feed rate in their analysis. The values selected were based on either the heat generation condition, the feed rate of combustible material, or the total solid (and liquid) feed rate. Each concept group selected that value which was deemed appropriate for their analysis. The tabulated flow rates may be used to reduce the data to a specific flow rate; that is, flow per unit mass of feed.

The analysis in Chapter 4 employed the Turton, et al. approach in calculating the terminal velocity only to obtain reference values for the two particles of interest (the catalyst and the sorbent). The specification of bubbling and turbulent fluidization velocities were based on Figures 1.5 and 1.6. The analysis in Chapter 5 only involved selecting velocities about or greater than those employed in existing test or transport fluidization systems in the petroleum industry.

The assumption was made in Chapter 4 that all the particles entering the second "fast" bed will remain entrained and will be expelled from the top of the fast section and sent to the cyclone. This assumption was made in order to estimate mass and energy balances in the system analysis. Unfortunately, no significant body of information could be found which would lead to a better assumption. Indeed, the several models available to predict entrainment from a bubbling bed are regarded as approximate. The solids flow through the upper bed was found using a correlation for elutriation above the transport disengaging height (TDH) in the bed. Narrowing the bed above the TDH should not decrease the net elutriation. Some of the rationale behind this assumption was that a significant amount of smaller particles ($<500\text{ }\mu\text{m}$) would be present in the system due to attrition. Thus most of the solids reaching the inlet to the upper bed will be significantly smaller than the $500\text{ }\mu\text{m}$ size. These particles entering the fast section will form a fast bed (significant slip velocities) condition. The larger particles caught in the upward movement into the fast section, it is hoped, will remain in the fast section for a longer time period. The other smaller particles will leave through the exit section cyclone. One may argue that particles predicted to leave the bubbling bed permanently (using an elutriation model) are unlikely to return to the bed once subjected to the higher superficial velocity associated with the upper bed. Clusters forming in the upper bed may fall against the rising gas and are the only apparent means by which solids may return to the upper bed. Fluidization theory is still inchoate in this area. Because this argument (or any other argument) may be supported by experimental observations, a rather thorough research problem has

been recommended to resolve these uncertainties. Attrition studies may have a profound effect of the analysis by determining a more reliable particle size distribution in the bed.

The problem of avoiding "hot spots" in the system has been the subject of many discussions. The oxygen injected into a "fast" section must mix with the reactive mixture such that the heat generated occurs over a large volume, tempered by the flow of solids through that volume. Crossflow jet injection into a fast moving gas-solid mixture is an unexplored problem. This problem is known to be complex. The two-phase flow will probably hinder the penetration of the oxygen into the core of the flow. In addition the strong fluctuating downflow of particles along the wall may cause oxygen to flow in the wrong direction, encountering perhaps another oxygen containing region, which will then certainly create hot spots (although perhaps intermittently). The solution to this problem will involve a unique-creative oxygen injection system which will be solved experimentally. This problem represents a significant effect in the research proposed for the further development of the hybrid and fast circulating bed concepts.

The gas and solid flow rates in the fast circulating fluidized bed require a more elaborate (including experimental) study before acceptable rates may be established. The desire for a fast bed, with its large backmixing of solids (and gases), will lead to a rather uniform temperature distribution in the system. However, this has been established for a fast fluidized combustor with an initial fuel-air mixture and one significant secondary air flow injection. However, the greater the solid suspension, the more difficult it may be to inject oxygen at the various locations. For lack of information, the solids flow rate in Chapter 5 now reflects the condition for "good" fast fluidization as described in reference 5.3. Unfortunately, the restriction of a rather small diameter required due to criticality was lifted after the analysis was completed. A larger riser diameter would allow for smaller superficial gas velocities, which would bring this study more into the realm of established experimental information. Also, since the completion of this study, a special issue devoted to circulating fluidized bed has appeared in the June 1992 issue of Powder Technology. Although a significant amount of information on circulating beds is presented for the operating conditions presented in this report, it appears that the desired research described in Chapter 6 is still necessary before a complete evaluation of the circulating system can be made.

CHAPTER 6 SUMMARY AND RECOMMENDATIONS

6.1 INTRODUCTION

The following summary and recommendations relate to the preliminary design study to develop an improved combustion system for the incineration of low-level radioactive wastes. Three different configurations are evaluated. The first involved improvements to the original system, while the remaining two are a hybrid fluidized bed system and a fast fluidized bed system.

6.2 SUMMARY

6.2.1 Existing System

The proposed modified system is an adaptation of the original combustion system that includes a primary reactor followed by an afterburner. Modifications to the original two reactor system were designed to improve overall combustion efficiency, reduce carbon and oxidant catalyst in the waste residue, reduce the amount of gas required, and permit the collection and sampling of combustion off gases before release to the atmosphere. Specific changes include:

1. Addition of solids recycle dip legs to the reactor cyclones to increase the solids residence time in the reactors, thereby facilitating improved carbon combustion and reduced sorbent and catalyst consumption.
2. Use of pure oxygen in place of air as the prime combustion gas to eliminate nitrogen in the off-gas.
3. Use of carbon dioxide as a reactor fluidizing gas. The use of carbon dioxide reduces the number of gaseous species and facilitates the off-gas environmental characterization before release to the atmosphere.
4. Mixing of solid and liquid waste to provide a more uniform heating value radwaste feed to reduce wide fluctuations in reactor combustion temperature. Make-up catalyst and sorbent will be supplied with the radwaste feed.
5. Addition of sand to both reactors may provide for a more uniform bed and to further increase the residence time of fine solids.

6. Use of a mixing chamber ejector to provide a positive motive force for movement of entrained solids from the primary reactor and to reduce the overall pressure drop of the system.
7. A new gas distributor to improve the fluidization characteristics of the primary reactor.

The basic characteristics of the original system, simple design and ease of control, are retained in the modified existing system. Improved performance is anticipated with the suggested modifications.

6.2.2 Hybrid Fluidized Bed

The proposed hybrid fluidized bed concept is based on the earlier Rocky Flats' experience using two reactors. In this hybrid concept, the two reactors are stacked as one but each section has its own geometry. The lower part of the reactor is larger in diameter, a bubbling bed, while the upper part of the reactor is smaller in diameter, a fast fluidized bed. Solids, elutriated from the lower bed, pass through the upper bed and are captured by cyclones for return to the system. Many of the modifications proposed for the existing system (previously described in 6.1.1) are incorporated in the hybrid system. The geometry and arrangement of support facilities offer some significant advantages to conventional designs. These include:

1. Ability to extend particle retention time either part of the reactor. The lower part of the reactor is ideal for the gasification of the radwaste and the upper reactor is employed to complete the combustion and optimize sorbent and catalyst usage. The bubbling bed provides for a good combustible solids residence time and for good mixing, while the upper bed provides adequate residence time for the conversion of all carbon containing species to carbon dioxide.
2. Use of two recycle cyclones provides the opportunity to classify solids elutriated from the reactor and may permit the separation and recycle of catalyst independent of reacted sorbent.
3. All of the remaining features for the modified system are utilized, i.e.: oxygen combustion gas, carbon dioxide fluidizing gas, mixing solid and liquid radwaste, mixing ejectors, and gas distribution plate.

4. An advantage over a separate two bed system is that solids and gases are conveyed from the lower to the upper reactor efficiently and with little pressure loss. Solids circulation assists in achieving temperature stability.

The hybrid system design is basic and simple. It offers a wide range of control options and can be easily constructed. Operation at acceptable temperatures can be achieved by insulating the lower (bubbling) bed and by cooling the gases and solids fed to the upper (fast) bed.

6.2.3 Fast Fluidized Bed

The second new system concept is the fast fluidized bed or fast recirculating bed. The configuration for this system is similar to the hybrid reactor except the reactor is of uniform dimension for the lower and upper zones. Further, since the velocities in the fast fluidized bed are to be much higher than in either zone of the hybrid reactor, the overall height of the fast fluidized bed reactor will be greater than for the hybrid reactor. Some of the other features for this system include:

1. The increased height of the reactor accommodates staged combustion along the height of the reactor using an oxidant mixture controller. This provides operating flexibility to control reactor temperature, solids, and liquid flow.
2. The higher reactor velocities offer expanded capabilities to feed solid and liquid waste.
3. As in the hybrid system, use of the two recycle cyclones provides the opportunity to classify solids elutriated from the reactor and may permit the separation and recycle of catalyst independent of reacted sorbent.
4. All of the remaining features for the modified system are utilized, i.e.: oxygen combustion gas, carbon dioxide fluidizing gas, and mixing ejectors.

The concept and principles of fast fluidized beds have been commercially utilized for many years in fluid catalytic cracking processes for transportation fuel production. These units have a very good operating record.

6.3 EVALUATION OF DESIGN CONSTRAINTS

6.3.1 Introduction

This section responds to the specific performance requirements of the radwaste combustion system, as described in a communication from S. Stiefvater, dated February 4, 1992. These needs may be summarized as:

- Maximum CO_2 production per mass of radwaste
- Minimum O_2 in the off-gas stream.
- Reduce catalyst and sorbent in solid refuse and
- Minimize bed diameter.

In addition, it was stated in discussions with Rocky Flats personnel that no CO was to be present in the off-gas stream. The questions concerning the off-gas composition are all interrelated and are discussed in section 6.2.2. Reduction of catalyst and sorbent in the solid refuse from the system will be discussed in section 6.3.2. It was concluded that the minimum bed diameter requirement is no longer significant during the April 24, 1992 meeting in Colorado. The removal of tramp metal was discussed during the January 22, 1992 meeting in West Virginia. Subsequently, a discussion of this problem, in the final report, was requested.

6.3.2 Off-Gas Stream Composition

The need to maximize the CO_2 concentration in the off-gas stream may also be stated as the carbon in the solid refuse flow must be a minimum and that the carbon monoxide in the off-gas must also be a minimum (zero). This need essentially dictates that all the carbon and carbon monoxide in the system be in intimate contact with sufficient oxygen under suitable reaction conditions (temperature, and catalyst). Boilers, and other combustion systems, frequently operate with excess oxygen (excess air) in order to consume all of the fuel. Subsequently, if the fuel composition was fixed, excess oxygen for complete combustion would be required. Also as the fuel composition varies from barrel to barrel sufficient excess oxygen must be present for a rapid increase in carbon content in the fuel feed. This leads to an additional requirement of significant excess O_2 in the system. If there is a rapid decrease in fuel carbon to the system, then an increase of oxygen in the off-gas will obviously occur.

At this time, it is very difficult to identify which proposed system would have a minimum oxygen flow in the off-gas. Experimental models of each system would be required to determine the minimum acceptable excess O_2 required. These models could then be modified to further minimize the O_2 requirements. The system model which produced the final minimum oxygen in the off-gas would then be the system of choice. The oxygen recirculated with the recirculating gases reduces the oxygen demand of the

entire system; however, the recirculation rates for each system could be adjusted so that the effect would be the same for all systems.

6.3.3 Reduction of Catalyst and Sorbent in Solid Refuse

The reduction in catalyst and sorbent consumption may be significantly reduced by recirculating the solids in the gas stream. The regeneration of the catalyst may also significantly reduce its consumption. The larger particles are removed from the gas stream and are recirculated into the fluidized bed. These larger particles should consist of catalyst sorbents, ash, and the unconsumed carbon. A second cyclone, downstream of the first cyclone, would remove fines from the gas stream. These fines would consist of attrited catalyst and sorbent as well as powdered ash. Ideally, there would be no carbon released by the second cyclone, which goes to the solid waste stream.

The attrition rates for the catalyst and sorbent are generally unknown. For the two modifications of the original system, these rates are low. For small particles in a fast bed, the attrition rates are also low. The attrition rate for particles moving between a bubbling bed and a fast bed has not been investigated. Again it is difficult to ascertain specific numbers for the catalyst and sorbent consumption. Obviously, the consumption rate will be significantly reduced by recirculating a large portion of the solids.

6.3.4 Tramp metal

The radwaste material being fed in the reactor may contain small metal objects. The disposal of an accumulation of this tramp metal is shown in Figure 3.4 of the text. Also, for the spouted bed distributor plate, the liquid spray pipe may be extracted and the tramp metal then removed. Therefore, it appears that removing tramp metal does not represent a problem.

6.4 COMPARISON OF SYSTEMS

In order to compare one system against another system, a certain degree of confidence must be developed in the analysis. Also, an advantage of one system over another system must be such that the approach may not be used in all the other systems. That is for a system to have an advantage over other systems, that which provides the advantage must be unique to that system. Consequently, the reduction in specific oxygen consumption due to recycling would be common to all systems. Also, the amount of specific carbon dioxide released can be reduced to that generated by combustion and to that required by the ejectors.

The specific¹ feed rate of catalyst and sorbent will depend on attrition rates, consumption rate (for sorbent), and the

cyclone removal efficiency. For the lower fluidization velocities of a bubbling bed, the attrition rate for particles of 500 μ m range should not be high. For the high velocities of a fast recirculating bed, the particles are small. Experience has shown that the small particles have a low attrition rate and produce low erosion rates. However, there appears to be no information to determine the steady state particle size distribution in the hybrid system. Ideally, the large particles would remain in the bubbling bed and the small particles would be

¹ The term specific indicates that the following quantity is based on a unit mass of radwaste.

recirculated through the fast bed. It is again difficult to compare the systems other than to say that the hybrid system represents a greater degree of uncertainty than do the others.

The next consideration is to compare the size of each system required to comply with the project goals. The modified systems essentially have the same dimensions of the original system, plus space for a heat exchanger, a settling tank, and a compressor. The two stage combustion allows for simple control of the off-gases from the afterburner, allowing for no CO in the off-gases. The oxygen to the first reactor controls the heat generation rate in that reactor whereas the heat generation rate in the afterburner is controlled by the radwaste feed rate into the first reactor. As there is no fixed external demand rate for steam or hot gases from this system, the two reactors can essentially be controlled independently.

The fast recirculating reactor concept combines the original two-stage combustion process into a single stage reactor operating as a fast recirculating bed. The recirculating bed must be sufficiently tall so that all of the carbon monoxide has been oxidized to carbon dioxide. In order to expand the length of flow (height) so that the combustion zone is not too concentrated, the oxygen is injected at three heights within the system. The uncertainty with this procedure is that there is no information on the mixing and the resulting combustion of a crossflow jet into a fast moving gas-solid stream. Therefore, the possibility exists that there may be CO in the off-gas for certain operating conditions and during transient conditions. A significant temperature control for this system would be to vary the radwaste feed rate.

The hybrid system represents two beds, intimately connected, acting as one bed. The combustion process is in two-stages, similar to the original system. The temperature may again be controlled in the bubbling bed by controlling the amount of oxygen injected into the bed. The temperature in the fast reactor section may be controlled by varying the radwaste feed rate. The comments about the required heights for the fast section are essentially the same as those for the fast

recirculating reactor. However, in the hybrid system, the fast section is in the turbulent flow regime. The hybrid system represents a significant degree of uncertainty. The stability of two intimately connected beds must be investigated. Also, the steady state particle size distribution, for the best set of operating conditions, will have to be determined. The injection of a jet of oxygen containing gases into a solid-gas flow must be investigated. The process will then dictate how tall the fast section must be for complete oxidation of carbon monoxide.

In conclusion, the modified original system represents good performance with the least uncertainty. The hybrid and fast fluidized systems represent (apparently) simple systems with a relatively high degree of uncertainty.

6.5 RECOMMENDATIONS

In order to arrive at the most desirable design concept, it is apparent that an experimental effort is required to develop some basic information. The basic information required concerns combustion, ejector performance with two-phase flow, erosion, attrition of bed particles, and heat transfer. Gas-solids separation in cyclones also represents some uncertainty; however, more is known about these devices and data on actual conditions should evolve from the experimental projects. It is, therefore, recommended that the following experimental program be undertaken to resolve the uncertainty in the proposed designs.

The combustion problems are rather complex. Initially, "chunks" of radwaste are injected into the reactor. These chunks of radwaste are then shredded by the bed material as they begin a pyrolysis process. This process rate depends strongly on the catalyst (and perhaps the sorbent) present and the local temperature. The hydrogen released probably reacts quickly with the oxygen present. The hydrocarbons released are partially converted to carbon monoxide and lighter hydrocarbons due to the presence of the catalyst. In addition, fines containing carbon may also be produced. Information relative to the rate of formation of the pyrolysis products is useful for the modified and hybrid system and essential in the fast recirculating system.

The pyrolysis products then enter a reactor where excess oxygen and the catalyst are present. This is the afterburner in the modified system and represents no problem. In the hybrid and fast recirculating beds, this process involves mixing the oxygen and combustibles in a gas-solid flow. The reaction rates, in the presence of the catalysts, for the production of carbon dioxide, are essential information for these fast beds.

The experimental project for combustion in the hybrid and fast circulating bed requires a cold and hot model of each system. The fast section in the hybrid system has different initial (inlet) conditions than the fast recirculating system; therefore, two models should be constructed. The cold models are

to be used primarily for flow studies and low temperature gas mixing studies. The cold hybrid model may also be employed for bed flow stability problems and for steady state particle size distribution studies. The hot models will ultimately allow one to predict the conditions necessary for all the carbon monoxide to be consumed before leaving the system.

Ejectors play a key role in the systems proposed in this study. However, very little appears to be known about the performance of ejectors when the secondary flow is "dusty". The first concern is the efficiency of the system and the second concern is the potential for erosion. Although the required pressure change across the ejector may not be large, the required primary flow may become significantly greater or the required pressure may become much higher. Both effects would lead to a less efficient system. Therefore, it is recommended that a study be conducted to determine an ejector performance as a function of the solids loading in the secondary stream.

The potential for erosion basically exists in the ejectors, jet exit ports, and at rapid changes in flow areas. The erosion rate may be estimated from data obtained during the combustion studies and the ejector studies. The attrition of particles in the hydrodynamics performance of the hybrid system is an important phenomenon, but must be part of the combustion study for this system. The effect of attrition on sorbent and catalyst performance is also an important problem in the combustion study for both systems. Heat transfer coefficients may be measured as part of the hot combustion studies. Concepts for controlling the heat transfer should be developed based on the experimental results. The concepts developed during the experimental study should be important in start-up, operating, and shut-down for each system.

Although there is available information on cyclones, it is recommended that an attempt be made to use cyclones to segregate catalyst and useful sorbent from sorbent fines and ash. The catalyst may then be regenerated to remove any inhibitors (heavy hydrocarbons) on the surface. Although this concept is in its early stages, a successful regeneration process may have a significant economic impact on the performance of the proposed system.

APPENDIX A
THE CONSULTANTS REPORT

**CONCEPTUAL DESIGN OF A FLUIDIZED BED COMBUSTOR
FOR VOLUME REDUCTION OF WASTE
CONSULTANTS REPORT AND COMMENTS**

APRIL 14, 1992

PREPARED BY: T.E. STRINGFELLOW

APPENDIX A
CONSULTANTS REPORT

GENERAL

As a result of the meetings held on the campus of West Virginia University, on April 9 - 10, 1992, Stringfellow Energy Consultants (SEC) was requested to provide comments on the materials presented at these meetings. The following comments and suggestions are based on the combined 35 years of experience in designing, operating, and testing fluidized bed combustion systems.

One general comment should be noted on the ability to model the combustion system. During the early stages of the meeting a statement was made that the combustion system could not be modeled, due to the relatively low temperature, the catalyst, and sorbents used. We believe that this system can be modeled to show off-gas constituents, ash, and combustion efficiency. A model similar to this is already in use at our office. We would require a proximate, ultimate, ash mineral, sorbent, and catalyst analyses to perform in the model.

Another point is that all three designs can be provided with a close approximation of material balance. Issues such as solids loading, and solids distribution can be calculated. If the designers wish to pursue this approach, SEC would be interested in assisting with this effort.

PROPOSED MODIFIED SYSTEM

The modified system proposed is an adaptation of the existing combustion system that includes a primary reactor, followed by an afterburner. From appearances, the proposed changes would include a mixing chamber eductor, a new air distributor, and the addition of CO₂ and O₂ as the prime fluidizing/combustion gas. Another major change is the recycling of solids back to the primary reactor and afterburner systems.

The use of a compression screw feeder would be utilized as the primary waste feed device. There was no indication of how the liquid waste would be fed, nor was there an indication of how the bed make up (catalyst and sorbent) would be introduced. We would recommend that these issues be resolved early in the next phase, perhaps using equipment similar to that shown in sketch #1 and sketch #2.

This design can be modeled using heat and material balance calculations found in the boiler and industrial arena. We would recommend that this approach be adopted as soon as possible. We would further recommend that the cyclones and eductor design be optimized, in order to evaluate the this approach in detail.

The air distributor used in the existing design, did not provide good mixing for the material, and was probably the cause of some problems during combustion. We would not recommend the use of an angle bed distributor. A flat plate arrangement, coupled with nozzles that prevent back leakage of material would be preferred. This approach would also provide a positive influence on the fluidization of material, and provide a directional flow for solid waste. A typical nozzle using this approach is shown in sketch #3.

We would not recommend using a mechanical device to remove solids from the bed. Problems with this type of system would be expected due to material erosion of the device, coupled with differential expansion problems between the cylinder rod and the outside casing. A simple loop seal arrangement, seal pot, or bed cooler would provide the efficient removal of material.

PROPOSED HYBRID DESIGN

The proposed Hybrid design offered by the University, is very similar to the attached (sketches #4 & #5) two patents. The use of a bubbling bed, located under a fast bed arrangement offers some advantages to conventional designs. First, this design offers the ability to extend particle retention time, in either part of the combustor. The lower zone is an ideal application for carbon utilization, and the upper zone further adds to the ability to reduce carbon and optimize any sorbent or catalyst usage.

As stated during the meeting, we have had experience with eductors in a high solids concentration atmosphere. The eductors in these cases could not survive the abrasive conditions. We have seen cases where ceramic lined eductors are used to feed coal and limestone, but the reliability is questioned. We would recommend eliminating both eductors (between 5 & 12, and 6 & 13 of figure 4.1a) and adding a larger eductor between 7 & 8. This will provide the system with the required effluent pressure, and reduce the risk of eductor erosion.

The solid refuse lines from the cyclone should be reduced to only one refuse line from cyclone #2. This probably would never be used, but the ability to drain off waste at this point is practical. The solids loop that will probably be utilized is the drain from the bubbling bed. Again, we would recommend that this drain be designed similar to the bed drain in the modified design.

To further enhance the carbon burn out cycle of this system we would recommend that all solids from both cyclones be introduced into the lower chamber. Although the solids will not stay long in this area, it will add heat to the lower zone and reheat the solids as they travel up the shaft of the combustor.

This design in all is basic and simple, it offers a wide range of control options, and can be constructed easily. The remaining system would probably be the same as the modified system. We again conclude that a model can be developed for this system, and that a close proximity of a heat and material balance can be produced.

PROPOSED CFB SYSTEM

The system described in the WVU CFB system is similar to those currently used in the industrial application of Fluidized Bed Boilers. The obvious changes in this design are the use of CO₂, and the gas mixture controller. This design offers expanded capabilities to reduce solid and liquid waste. Further the design offers operating flexibility to control temperature, solids, and liquid flow.

The only concerns we have at this point is the velocity of the lower zone, and the size of the proposed cyclone downcomer. First the fluidizing velocity of the lower zone must be modeled to prevent hydraulic decrepitation of the solids material. Even if 100 micron material is used, this material can severely fracture if the velocity is too low. In shaft type combustors this phenomenon is very common. Essentially we are using a straight shaft combustor in the application.

Second, the cyclone downcomer is too large and gas leakage back up the cyclone leg is a definite possibility. A small downcomer would provide resistance to flow, and can be easily filled with solids to maintain some type of seal. Further we would remove the valves shown in the downcomer lines, and remove the ash discharge point (19). These are not necessary, and only complicate the operation of this simple system.

The gas mixture controller is one of the more unique features of this design. It has the ability to control percentages of CO₂, O₂, and recycle gases in any area of the combustor. To further enhance this control, we would hope that this

controller also is utilized as a flow control device, where the flow to each of the four areas can be controlled based on temperature and CO of a given area.

We do not recommend partial combustion/pyrolysis to be utilized throughout the shaft of this system. We believe that this process should occur in the lower zone of the chamber, and the remaining height be used to attempt complete carbon utilization. This will reduce the load on the cyclones, and improve the overall efficiency of the combustion process. Further using the staged air process better utilization of the catalyst and sorbent would be apparent.

COMMENTS AND RECOMMENDATIONS

All of the proposed systems have their advantages and disadvantages, but the Hybrid design and the CFB designs appear to be simple in design, compared to the Modified version of the existing system. Due to the simplicity of these two designs, we believe that the intended goals, and constraints can be better served by the Hybrid and CFB systems.

Each of the three designs can be physically modeled to provide data, and show each individual viability. The costs associated with modeling is expensive, but in order to achieve a proven technology for this application, models must be used. We can recommend an approach that will reduce the time requirements, and costs associated with testing. We would strongly recommend that the participants in this project consider using a Pyrex (or similar material) type of model that can be utilized in a hot or cold phase of testing.

This approach will be able to show the solids flow aspects of each system, and can provide needed data under hot conditions. Waste material for this application, can be obtained from a number of sources. We can closely construct similar materials for combustion to resemble the analyses of the current wastes at Rocky Flats. The data from these types of tests would be the foundation for presenting the concept to the public, and other concerns.

The cost effective approach mentioned above would provide the investigators with significant data, cut research costs, and provide a compact schedule for eventual implementation. Should the investigators wish to proceed along this concept then modeling would be able to begin, once a final design for each system is completed.

COMPRESSION SCREW/LIQUID FEEDER

SKETCH #1

SKETCH #2

DIRECTIONAL FLUIDIZING
NOZZLE

SKETCH #3

United States Patent [19]
Stringfellow et al.

[11] Patent Number: 4,598,653
[45] Date of Patent: Jul. 8, 1986

[54] COMBUSTION SYSTEM FOR BURNING
FUEL HAVING VARIOUS PARTICLE SIZES

[75] Inventors: Thomas E. Stringfellow, Aurora
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[73] Assignee: Stearns Catalytic World Corporation,
Denver, Colo.
[21] Appl. No: 755,981
[22] Filed: Jul. 17, 1985

Related U.S. Application Data

[63] Continuation-in-part of Ser. No. 649,558, Sep. 12, 1984,
Pat. No. 4,365,119.
[51] Int. Cl.: F23D 1-00
[52] U.S. Cl.: 110/347; 110/345;
110/263;
[58] Field of Search: 110/263; 245; 347;
122; 412

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Primary Examiner—Edward G. Favory
Attorney Agent or Firm—Mark L. Rodgers, E. Eugene
Innis, James C. Simmons

[57] ABSTRACT

The present invention is a process and associated apparatus for obtaining energy from a fuel source having particles varying in size from fines to coarse. The fuel source is introduced into a passageway formed between an upper combustor and a lower combustor. A gas stream is moved upwards through the passageway such that the fine particles are entrained in the gaseous stream and carried into the upper combustor.

16 Claims, 1 Drawing Figure

Sage et al.

(iii) Patent Number: 4,565,139

[45] **Date of Patent:** Jan. 21, 1986

[54] METHOD AND APPARATUS FOR OBTAINING ENERGY

[75] Inventors **Warrnie L. Sage, Littleton, Thomas F. Stringfellow, Aurora, both of Colo**

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(21) Appl No 649,558

[22] Filed Sep. 12, 1984

[51] Int. Cl.⁴ F23D 1/00

[52] U.S. Ct. 110/347; 110/344.

[58] **Field of Search** 110/245, 110/263, 110/274, 110/347, 263, 244, 245

122/4 D

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Primary Examiner—Edward G. Favors

Attorney, Agent, or Firm—Mark L. Rodgers, E. Eugene Innis, James C. Simmons

[57] ABSTRACT

A low BTU, low volatile content, low carbon and high ash fuel is combusted in a combined shaft kiln combustor and fluidized bed combustor to generate heat which is converted to useful energy. The fluidized bed combustor is mounted vertically above the shaft kiln combustor. A venturi is provided in a passageway connecting the shaft kiln combustor and the fluidized bed combustor to produce a gaseous stream moving at a velocity sufficient to separate finer particles from the low BTU, low volatile content, low carbon and high ash fuel so that the coarser particles are burned in the shaft kiln combustor and the finer particles are burned in the fluidized bed combustor.

19 (7 aims, 1 Drawing Figure)

SKETCH #5

APPENDIX B

EJECTOR ANALYSIS

APPENDIX B

EJECTOR ANALYSIS

Ejectors had been utilized in developing the designs of the proposed incineration systems. Thus it was appropriate to briefly describe the performance of these devices.

The purpose of an ejector is to move fluid (secondary flow) into a region of higher pressure. A basic ejector design is shown in Figure B.1. A high pressure (primary) gas is accelerated so that when it interacts with the low pressure (secondary) gas, the low pressure gas will be entrained into a region of high pressure. This device is rather simple, with no moving parts, and its efficiency is acceptable considering the simplicity and reliability of the ejector.

In order to quantify the efficiency of an injector, [B.1], its performance is compared to that of an ideal compression system. The secondary flow is isentropically compressed in a compressor driven by an ideal turbine. The ideal turbine is represented by an isentropic expansion of the primary flow. The exit streams from the compressor and turbine are at the ejector exit pressure. For incompressible flow, the compressor power (see Figure B.1 for terminology),

$$\dot{W}_{C,I} = \dot{m}_s [P_e - P_s] \frac{1}{\rho} \quad (\text{B.1})$$

and for the turbine

$$\dot{W}_{P,I} = \dot{m}_p [P_p - P_e] \frac{1}{\rho}. \quad (\text{B.2})$$

These equations assume that the kinetic energy is negligible where the pressures are measured. The isentropic efficiency for the ejector may be written as:

$$\begin{aligned} \eta_I &= \frac{\dot{W}_{C,I}}{\dot{W}_{T,I}} \\ &= \left(\frac{\dot{M}_s}{\dot{M}_p} \right) \frac{[P_e - P_s]}{[P_p - P_e]} \end{aligned} \quad (\text{B.3})$$

Introducing the dimensionless quantities

$$r = \frac{P_e}{P_s} \quad (B.4)$$

$$R = \frac{P_P}{P_s} \quad (B.5)$$

$$m_i = \frac{\dot{m}_e}{\dot{m}_s} \quad (B.6),$$

then one may write

$$\eta_I = \frac{1}{m_i - 1} \frac{(r - 1)}{(R - r)}. \quad (B.7)$$

The mass flow ratio, m_i , can also be expressed as a function of the area ratio α ($= A_1/A^*$), R , and r . This allows for an equation of the form

$$\eta_I = f(\alpha, R, r) \quad (B.8)$$

The ejector efficiency is also dependent on all the geometric details; however, the diffuser area ratio β ($= A_m/A_e$) is a significant parameter.

Typical ejector efficiencies [B.2] are shown in Figures B.2 and B.3. The efficiencies are shown as a function of $(r - 1)/(R - 1)$ and α . From the known pressures and geometry, one may determine efficiency which in turn leads to the mass flow ratio. The diffuser geometry also plays an important role in the efficiency. Figure B.2 shows the efficiencies for $\beta = 1$ and Figure B.2 shows the efficiencies for $\beta = 0.25$. Ejectors are shown to have a reasonably high efficiency, if designed for a specific condition. The smaller the β parameter, the higher the possible ejector efficiencies.

Empirical formulas have been developed for the performance of ejectors with compressible flow. The quantity R is either known or may be estimated. The quantity r is based on the pressure drops in the system. The dimensionless flow ratio m_i may be calculated from

$$\frac{1}{m_i} = \{1 - \ln(r)/c_1 [\ln(R)]^{c_2}\}^{2.6}, \quad (B.9)$$

if the flow is isoenergetic.

The constants C_1 and C_2 depend on the ejector geometry. For the optimum value of α , use:

for $\beta = 1$, $C_1 = 0.830$ and $C_2 = 0.753$
 $\beta = 0.25$, $C_1 = 0.912$ and $C_2 = 0.808$.

If the flow is non-isoenergetic, then the actual mass flow ratio is

$$m = 1 + (m_i - 1) \left(\frac{a_s}{a_p} \right)$$

where

$$a = \left[\frac{\alpha RT}{M} \right]^{1/2}.$$

This brief description of ejector performance is only intended to serve as an introduction to the subject. Additional performance data may be found in [B.2 - B.5] and in the literature of those who market ejectors.

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$$\alpha = \frac{A^*}{A_1}, \beta = \frac{A_m}{A_e}, A^* \text{ Nozzle Throat Area}$$

Figure B.1 Basic Design and Terminology for an Ejector

$$\frac{(P_{oe} - P_{os})}{(P_{op} - P_{os})} = \frac{r - 1}{R - 1}$$

Figure B.2 Ejector Efficiency for a Constant Area Flow, $\beta = 1$

Figure B.3 Ejector Efficiency for an Ejector
With a Diffuser, $\beta = 0.25$

APPENDIX C

RESPONSE TO THE TECHNICAL COMMENTS OF THE REVIEWERS

APPENDIX C

RESPONSE TO THE TECHNICAL COMMENTS OF THE REVIEWERS

It is unfortunate that each concept group used a different solid feed rate in their analysis. The values selected were based on either the heat generation condition, the feed rate of combustible material, or the total solid (and liquid) feed rate. Each concept group selected that value which was deemed appropriate for their analysis. The tabulated flow rates may be used to reduce the data to a specific flow rate; that is, flow per unit mass of feed.

The analysis in Chapter 4 employed the Turton, et al. approach in calculating the terminal velocity only to obtain reference values for the two particles of interest (the catalyst and the sorbent). The specification of bubbling and turbulent fluidization velocities were based on Figures 1.5 and 1.6. The analysis in Chapter 5 only involved selecting velocities about or greater than those employed in existing test or transport fluidization systems in the petroleum industry.

The assumption was made in Chapter 4 that all the particles entering the second "fast" bed will remain entrained and will be expelled from the top of the fast section and sent to the cyclone. This assumption was made in order to estimate mass and energy balances in the system analysis. Unfortunately, no significant body of information could be found which would lead to a better assumption. Indeed, the several models available to predict entrainment from a bubbling bed are regarded as approximate. The solids flow through the upper bed was found using a correlation for elutriation above the transport disengaging height (TDH) in the bed. Narrowing the bed above the TDH should not decrease the net elutriation. Some of the rationale behind this assumption was that a significant amount of smaller particles ($<500\text{ }\mu\text{m}$) would be present in the system due to attrition. Thus most of the solids reaching the inlet to the upper bed will be significantly smaller than the $500\text{ }\mu\text{m}$ size. These particles entering the fast section will form a fast bed (significant slip velocities) condition. The larger particles caught in the upward movement into the fast section, it is hoped, will remain in the fast section for a longer time period. The other smaller particles will leave through the exit section cyclone. One may argue that particles predicted to leave the bubbling bed permanently (using an elutriation model) are unlikely to return to the bed once subjected to the higher superficial velocity associated with the upper bed. Clusters forming in the upper bed may fall against the rising gas and are the only apparent means by which solids may return to the upper bed. Fluidization theory is still inchoate in this area. Because this argument (or any other argument) may be supported by experimental observations, a rather thorough research problem has

been recommended to resolve these uncertainties. Attrition studies may have a profound effect of the analysis by determining a more reliable particle size distribution in the bed.

The problem of avoiding "hot spots" in the system has been the subject of many discussions. The oxygen injected into a "fast" section must mix with the reactive mixture such that the heat generated occurs over a large volume, tempered by the flow of solids through that volume. Crossflow jet injection into a fast moving gas-solid mixture is an unexplored problem. This problem is known to be complex. The two-phase flow will probably hinder the penetration of the oxygen into the core of the flow. In addition the strong fluctuating downflow of particles along the wall may cause oxygen to flow in the wrong direction, encountering perhaps another oxygen containing region, which will then certainly create hot spots (although perhaps intermittently). The solution to this problem will involve a unique-creative oxygen injection system which will be solved experimentally. This problem represents a significant effect in the research proposed for the further development of the hybrid and fast circulating bed concepts.

The gas and solid flow rates in the fast circulating fluidized bed require a more elaborate (including experimental) study before acceptable rates may be established. The desire for a fast bed, with its large backmixing of solids (and gases), will lead to a rather uniform temperature distribution in the system. However, this has been established for a fast fluidized combustor with an initial fuel-air mixture and one significant secondary air flow injection. However, the greater the solid suspension, the more difficult it may be to inject oxygen at the various locations. For lack of information, the solids flow rate in Chapter 5 now reflects the condition for "good" fast fluidization as described in reference 5.3. Unfortunately, the restriction of a rather small diameter required due to criticality was lifted after the analysis was completed. A larger riser diameter would allow for smaller superficial gas velocities, which would bring this study more into the realm of established experimental information. Also, since the completion of this study, a special issue devoted to circulating fluidized bed has appeared in the June 1992 issue of Powder Technology. Although a significant amount of information on circulating beds is presented for the operating conditions presented in this report, it appears that the desired research described in Chapter 6 is still necessary before a complete evaluation of the circulating system can be made.

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