THE OPTIMISATION OF COMBUSTION SYSTEMS FOR THE BURNING OF CEREAL STRAW AS A FUEL

by

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SUMMARY

International interest in fluidised bed combustion (F.B.C.) derives from the fact that it involves new technology and it is the only combustion system that can use low grade fuels (including those of high or variable ash content) efficiently.

This thesis presents a study of the combustion of straw in a fully fluidised and systematically interrupted flow test rig.

In the interrupted flow mode, it was found that during the period in the cycle when the bed was slumped, due to the reduction in primary air, a gasification process took place. This resulted in a higher percentage of volatiles being burned in the bed and preliminary results indicated that interruption gave an increase in overall efficiency of approximately 4%. It also led to lower losses being incurred in the flue gases and, to a lesser extent, to a reduction in the losses expected due to incomplete combustion of the fuel. Elutriation and carbon losses were also reduced to 88% and 86% respectively of those recorded during uninterrupted fluidisation.

By increasing the area of the zone directly above the fluidised bed, the products of combustion were decelerated sufficiently to ensure that ash was not found on the walls or roof to the same extent as that previously noticed on commercial straight-sided combustion chambers. Small quantities of ash were found in the ducting leading to the cyclone but they were approximately the same for operation in both the uninterrupted and interrupted fluidisation mode.

GLOSSARY

Fluidised Bed

A layer of particles through which a gas (most commonly air) is passed vertically at a velocity that is sufficiently high to cause motion resembling that of a liquid.

Fluidised Bed Combustion (FBC)

The combustion of fuels in a fluidised bed

Elutriation

The loss of fuel or bed particles carried off the surface of the bed by the upward flow of the products of combustion

Distributor Plate (or Bedplate)

The base on which the bed material is supported.

Standpipes

Short pipes attached to the bedplate through which air is passed into the bed material.

Sparge Pipes

Horizontal pipes with drilled holes through which air is passed into the bed material.

Drain Pipe

A pipe fitted to the bedplate through which bed material can be drained.

Plenum Chamber

The space below the bed plate into which air from a fan passes and from which it is distributed to the bed plate, stand pipes or sparge pipes.

Subsidiary Fuel

Fuel, either gas or oil, used to initially heat up the bed.

Stoichiometric Air

The theoretical quantity of air required to burn the fuel completely.

Excess Air

The air required in practice in addition to the stoichiometric air to achieve complete combustion.

Freeboard

The space above the fluidised bed in the combustion chamber.

Forced Draught

The air supplied by a fan, normally at ambient temperature, to support the combustion process.

Induced Draught

The withdrawal of products of combustion from the furnace or boiler by means of a fan.

Balanced Draught

A combination of Forced and Induced Draught with usually slightly negative pressure in the freeboard.

Cooling Coils

Pipes passing through either the boiler shell or access ports into the combustion chamber, with pumped circulation of water.

Thermic Syphons

Tubes attached to the inner walls of the combustion chamber through which water or water and steam are circulated by gravity.

Cooling Surface

Those parts of the combustion chamber backed by water and of the thermic syphons or cooling coils which are immersed by the fluidised bed when firing. These surfaces are used as one of the means of controlling bed temperature.

Static Bed Depth

The depth of bed materials when at rest between the freeboard surface and the level at which air is introduced into the bed material.

Bed Expansion

The increase in the depth of bed material when fluidised.

Bed Slump

The return of the bed material from the fluidised state to the static condition.

Polycyclic Aromatic Hydrocarbons (P.A.H.)

A member of the class of hydrocarbons producing carcinogenic activity.

NOMENCLATURE

A_b	Cross-sectional area of bed (m^2)
d_p	Particle diameter (m)
$\bar{d}_p$	Mean particle diameter (m)
g	Acceleration due to gravity (m/s^2)
H	Fluidised bed height (m)
H_o	Bed height at incipient fluidisation (m)
k	Constant
m	Mass of particles in bed (kg)
t	Time
T_b	Bed temperature (°C)
T_o	Initial air temperature (°C)
U	Superficial gas velocity (m/s)
U_{mB}	Gas velocity at which bubbling first occurs (m/s)
U_{mf}	Gas velocity at which bed becomes fluidised (m/s)
P	Pressure (bar)
R	Universal gas constant = 1.986
R_e	Reynolds number $d_p U_{pg}/\mu_g$
Re_{mf}	Reynolds number based on U_{mf}
ε	Average voidage fraction in a fixed bed
ε_o	Voidage fraction at incipient fluidisation (0.5)
ε_{mf}	Voidage at minimum fluidisation
μ	Viscosity of gas (kg/m/s)
v	Kinetic viscosity of gas $= \frac{\mu}{p}$ (mm/s)
p_g	Gas density (kg/m^3)
p_a	Ash density = 2360 (kg/m^3)
p_c	Carbon density in char, coke or coal particles (kg/m^3)
p_f	Density of fluid (kg/m^3)

ρ_p	Density of particle (kg/m^3)
μ_g	Viscosity of gas (kg/m-s)
X_a	Excess Air
η_c	Combustion Efficiency
h_{max}	Maximum bed to surface heat transfer coefficient (W/m^2k)
k_g	Gas thermal conductivity (W/mK)

INTRODUCTION

The development of modern agricultural systems is economically and technologically dependent on there being convenient sources of energy available. The efficiency of developed agriculture relies on the use of tractors, artificial fertilisers, fresh water and, finally, on transport for the distribution of produce; all these operations are energy-consuming. However, agriculture also contributes to the world's energy resources through biomass cultivation. Wood, for example, both cultivated and otherwise, probably satisfies something like ten per cent of the world's energy needs. Some one thousand million tonnes of wood are used as a fuel annually (D. COUNIHAN 1981); most of this is consumed in the southern hemisphere, however as much again is produced in the north but is used mainly for construction and paper making.

It is not always necessary to grow crops for the production of biological fuels; vegetable and animal wastes are readily fermented to produce methane and research is under way in the U.S. and elsewhere to produce liquid fuels from biomass.

About three per cent of the energy consumed in the U.S. is used in the production side of agriculture out of the total of fifteen per cent for the entire food system. (B. A. STOUT - 1981). The industry is regarded as so important that in the event of future energy shortages, agriculture and defence have the top priorities for fuel allocation. The United States Department of Agriculture has taken the energy position seriously and apart from more efficient management and conservation, it is developing biomass-based alternatives to petroleum and liquid petroleum gas. Concern, though, is already being expressed about this programme, and it is argued that while the utilisation of agricultural wastes is one thing, the use of land to produce biomass specifically for oil production for fuel is something quite different (Tables 1 & 2). Whilst each American farmer can produce food for at least 73 other people (B. A. STOUT - 1981), the removal of land from direct food production could seriously affect another important problem - that of world hunger.

While it may be practical to raise energy crops in countries where there is underutilised land, i.e. in the tropics where annual dry matter yields can reach 85 t/ha, it is unlikely that such an approach could be widely followed in the U.K. where maximum dry matter yields are of the order of 20 t/ha. To produce the energy equivalent of present U.K. oil consumption, an area of 25mt ha - some 6mt ha more that the whole of the country - would be required (D.H. WHITE 1979).

Suggestions have been made that energy crops in this country could be sandwiched between rows of food crops, or even that marginal land where other crops will not grow satisfactorily should be used for afforestation; while these possibilities should not be ignored, it must be realised that the potential benefits will never be substantial. Of greater benefit in the short term would be a closer look at the utilisation of biomass which is currently dumped or burned as a convenient method of disposal. Straw and forestry waste are the two materials in this category which are most readily available.

Before the introduction of the combine harvester, all cereals were stacked in the sheaf and protected by a straw thatch. Only a small proportion of the straw, however, was used for anything other than feeding to cattle or for bedding them, with the subsequent production of farmyard manure. Today most of the nutrients required for crop production are supplied by manufactured inorganic fertiliser and all grain is stored in bulk, already threshed.

TABLE 1 - SOURCES OF VEGETABLE OILS

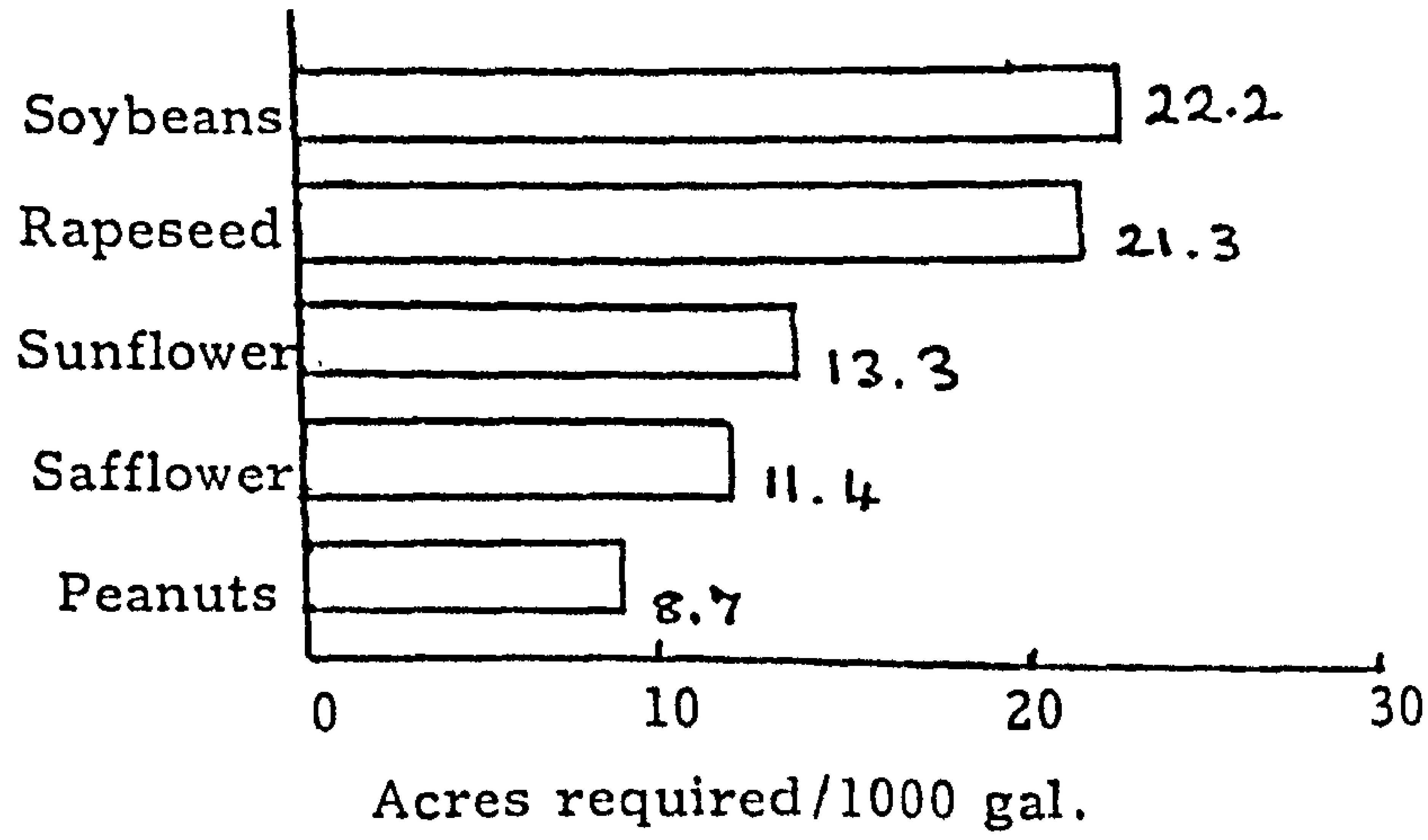
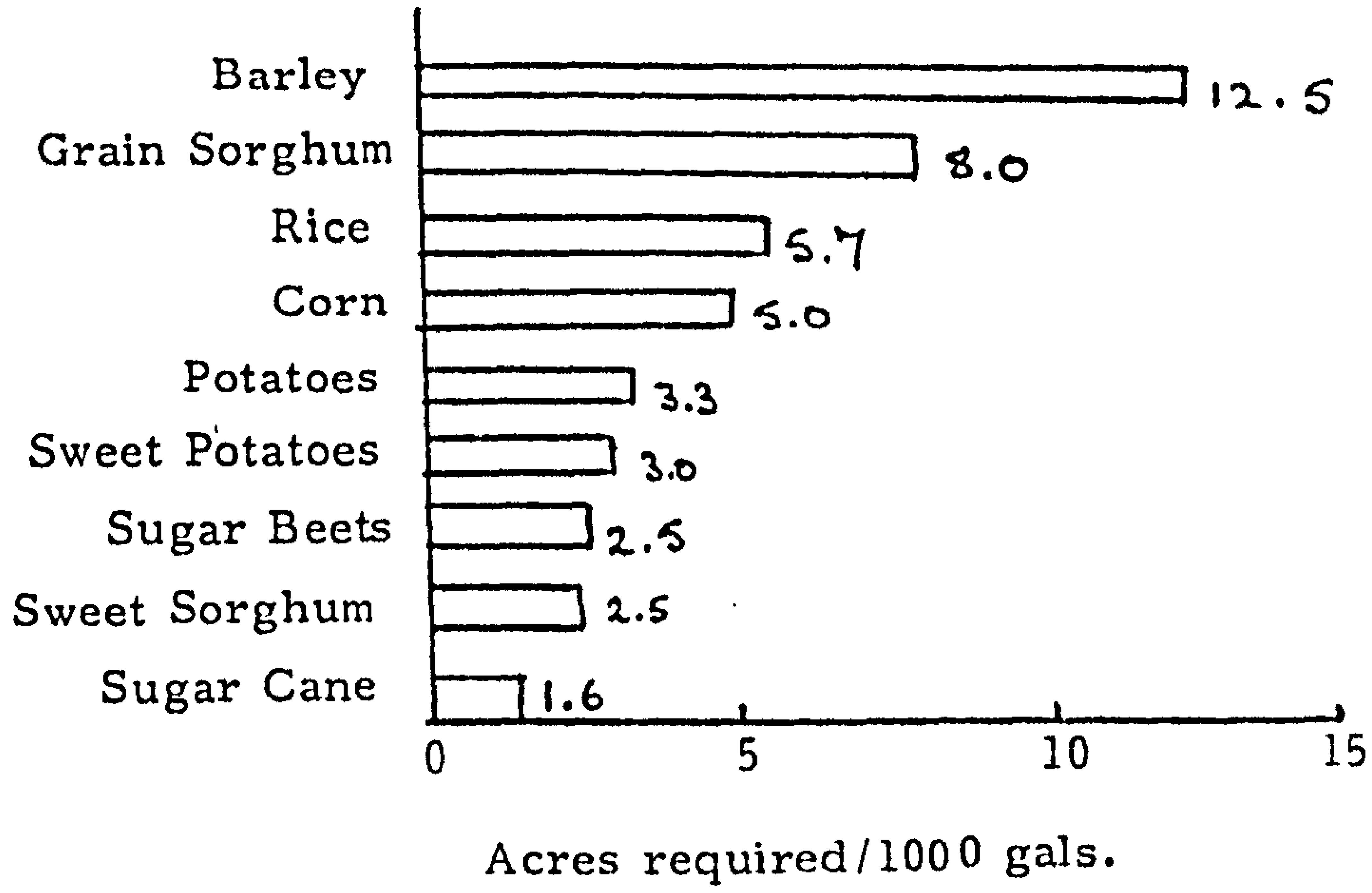


TABLE 2 - SOURCES OF ETHANOL FEEDSTOCK



SOURCE - MILLER 1981

Present intensive methods of livestock production require little straw and a massive surplus has been produced for which there has been no obvious use. Because of this, farmers have resorted to the practice of field burning of straw as a means of clearing the land quickly to aid in the preparation of seedbeds for the following crop. The argument put forward is that substantial benefits are obtained by clearing weeds and leaving the soil in a suitable condition for direct drilling or minimal cultivation. In his book 'Straw for Fuel, Feed and Fertiliser', STANIFORTH (1982) argues the case for and against burning and the conclusion drawn is that any advantage in either direction is very marginal. Nevertheless, burning is a practice that has been widely adopted because it saves time and allows a high proportion of autumn sown crops to be grown.

Although codes of practice have been developed for field burning, considerable public opposition to it has built up, not just because of the effect on the environment but because of the loss of potential energy. It is estimated that in the U.K., the average production of cereal straw is 12.5 m/tonnes (LARKIN 1982) (Table 3) of which at least one-third is surplus to agricultural requirements. This surplus has an energy equivalent of over 2.5 m/tonnes of coal. These figures are confirmed by Staniforth (Tables 4 & 5) (STANIFORTH 1982) and by White (WHITE 1979) who states that straw's gross energy value is equivalent to about four per cent of the U.K. petroleum usage.

It seems unlikely that the amount of surplus straw will decrease, indeed it is predicted that total annual yields will soon average 15 M tonnes and that the surplus will increase to 10 M tonnes/annum. (STANIFORTH 1982). The farming world and associated interested bodies are aware of the situation and a computer-aided system for showing the density of straw production in England and Wales (Fig 1) has been developed at the National College of Agricultural Engineering (CLEGG et al 1984).

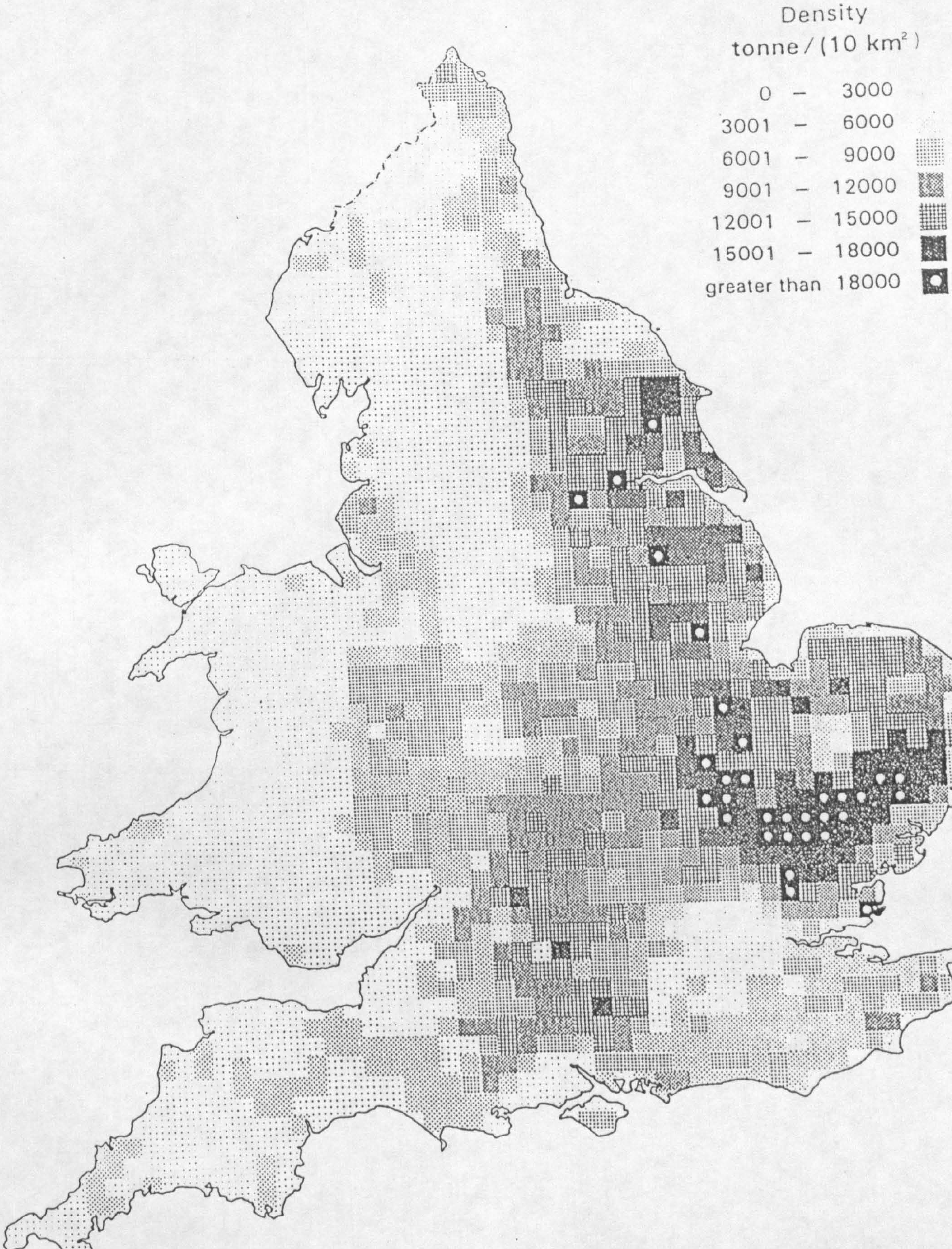


Figure 1. Density of total cereal straw production in England and Wales, tonne of dry matter per 10 x 10 km square grid

TABLE 3 - TOTAL U.K. AGRICULTURAL WASTE AND RESIDUE PRODUCTION

	Fresh Weight	Dry Weight	Gross Energy Content
	kt	kt	pj
Cattle Wastes	46,377	4,637	81.2
Pig Wastes	10,538	894	18.7
Poultry Wastes	2,163	1,542	22.9
Cereal Straw	12,488	10,739	204.1
Sugar Beet Tops	7,225	1,084	16.7
Other Crop Residues and Wastes	4,564	1,206	20.8
TOTAL	83,355	20,189	364.3

SOURCE - LARKIN 1982

TABLE 4 - STRAW PRODUCTION FOR WHEAT, BARLEY AND OATS - ENGLAND AND WALES, 1977-81

Year	Crop	Area '000 ha	Straw Yield '000 Tonnes	Total Cereal Straw Yield '000 Tonnes
1977	Wheat	1049	4196	-
	Barley	1882	5834	-
	Oats	127	483	10,513
1978	Wheat	1230	4982	-
	Barley	1808	5695	-
	Oats	120	462	11,139
1979	Wheat	1340	5494	-
	Barley	1796	5747	-
	Oats	86	335	11,576
1980	Wheat	1413	5864	-
	W. Barley	721	5944	-
	S. Barley	1108		
	Oats	105	415	12,223
1981	Wheat	1458	6124	-
	W. Barley	790	6046	-
	S. Barley	1042		
	Oats	100	400	12,570

SOURCE - STANIFORTH 1982

TABLE 5 - STRAW YIELD, STRAW BALED & SURPLUS STRAW, IN ENGLAND & WALES - 1977-81

Year	Total Straw Production '000 Tonnes	Straw Baled '000 Tonnes	Surplus Straw '000 Tonnes
1977	10,513	5370	5143
1978	11,139	4900	6239
1979	11,576	5365	6211
1980	12,223	5368	6855
1981	12,570	5107	7463

SOURCE - STANIFORTH 1982

Today in the U.K., some fifteen per cent of the straw produced is fed to ruminant livestock. Many attempts have been made to improve its feeding value, the most recent innovations having been treatment with alkalis. When treated with sodium hydroxide, this nutritionally improved straw (N.I.S.) can be added to commercial animal compound feeds at inclusion rates of around fifteen per cent. A few years ago there were fourteen straw processing plants operating commercially in Denmark and ten in the U.K., but production of N.I.S. has now been dramatically reduced in both countries, partly because of the difficulty of obtaining and treating supplies of high quality straw economically. Some on-farm treatment with caustic soda has been carried out but currently, interest in the U.K. is concentrated on the use of ammonia; this is applied either in ovens or to straw encased in plastic film.

What are the alternatives to field burning that could be profitably applied by farmers? They could incorporate more straw in the soil (but it is unlikely that this would increase income); there might be limited scope for them to increase the amount used as feed for cattle and sheep, but this is doubtful, or they could use some of it as a fuel to replace purchased oil, coal or electricity on the farm. Off-farm uses envisaged are again as a fuel, for horticulture or drying purposes, or as a fibre source in paper and board making.

The gross energy value of straw depends upon species and variety, and with an average moisture content of around 15% it is between 17.5 and 18.5 MJ/kg, i.e. approximately the same as that of wood. As far as can be ascertained, all the oat straw produced is used for feeding; some barley straw is also used for feeding while both wheat and barley straw are used for bedding livestock. Both wheat and barley straw burn equally well.

Another material which has become available in the last few years is oil seed rape straw; this has no value for either feeding or bedding, however it has a calorific value similar to the cereal straws and has the added advantage that it burns more uniformly than the others in 'whole bale' furnaces. (Min. Agri. Fisheries & Food 1983)

When straw is pyrolised by heating in a closed retort at temperatures betwen 500 and 1000°C a combustible gas, a liquid containing organic compounds (including tars), together with oils and ash are produced. Work on wastes at Warren Springs Laboratory (DABORN - 1973) suggests that the amount of useful combustible gases depends largely on the temperature of the retort (Table 6).

When straw is burned it produces more ash than many other fuels (Table 7) but on the other hand, a comparison of gross energy costs with the major fuels favours straw. It can be seen from Table 8a that straw bought at £30 per tonnes is approximately equal in gross energy to coal at £60 per tonne and oil at 10p/litre. Unfortunately straw is a difficult material to deal with; it is very bulky and has to be removed quickly from the field in a form in which it can be transported and stored.

A question that must be asked is what are the cost/efficiency relationships of the present methods of utilising straw as a fuel in, on or near farm installations? The source of the straw as well as the cost of packaging affect its cost at the furnace.

If the cost of buildings for storage is ignored, the only ones involved would be for baling and short distance haulage; current costs of these operations would be some £0.83 per 1000 MJ at 60% combustion efficiency, which is less than one quarter of the cost of using oil.

TABLE 6

THERMAL DECOMPOSITION OF STRAW BY PYROLYSIS

Mass Balance

	Products (% W/W)	
Pyrolysis Temperature	500°C	800°C
Gas	16.0	31.9
Liquid *	44.3	42.9
Char	39.4	25.6

* This is liquid produced not moisture (i.e. water is formed in the pyrolysis process.).

Analysis of Products

	Gas (% V/V)	
Pyrolysis Temperature	500°C	800°C
Hydrogen	3.87	24.89
Carbon dioxide	44.72	23.25
Carbon monoxide	38.71	31.48
Methane	6.31	16.01
Ethane	1.94	1.32
Ethylene	2.71	1.92
Propane	0.41	0.25
Propylene	1.26	0.87

	Liquid (% W/W)	
Pyrolysis Temperature	500°C	800°C
Aqueous	88.9	69.0
Oil	11.0	31.0

	Char (% W/W)	
Pyrolysis Temperature	500°C	800°C
Ash	13.0	14.5
Hydrogen	4.3	2.4
Carbon	73.4	72.5
Oxygen (by difference)	9.3	10.6

Source: Daborn 1973

TABLE 7 - APPROXIMATE CALORIFIC VALUES & ASH CONTENTS OF FUELS

Fuel	Calorific Value GJ/Tonne	Ash % by weight
Propane	50.0	-
Gas Oil	45.5	-
Medium Fuel Oil	42.9	0.03
Heavy Fuel Oil	42.5	0.04
Bituminous Coal	26.0	5.00
Wood	18.0	2.00
Straw	18.0	3.90

Table assembled from various sources

TABLE 8a COMPARISON OF GROSS ENERGY COSTS
FOR MEDIUM FUEL OIL, COAL AND STRAW

Oil							
Purchase Price	Pence per Litre	10	12	14	16	18	20
Gross Energy Cost	£ per GJ	2.3	2.8	3.2	3.7	4.2	4.6
Coal							
Purchase Price	£ per Tonne	45	50	55	60	65	70
Gross Energy Cost	£ per GJ	1.7	2.0	2.1	2.3	2.5	2.7
Straw							
Purchase Price (at 15% MC)	£ per Tonne	10	15	20	25	30	35
Gross Energy Cost	£ per GJ	0.70	1.10	1.40	1.80	2.20	2.50

TABLE 8b - ON FARM PELLETING COSTS

	£/t
Tub Grinder (elec. + wear)	4.00
Hammer Mill (elec. + wear)	4.00
Pelleter (elec. + wear)	2.00
Pellet Cooling & Conveying	2.00
Amortisation (ten year pay back)	3.00
Total	15.00

TABLE 8c - COMPARISON OF EFFECTIVE ENERGY COSTS FOR MEDIUM FUEL OIL, COAL, CHOPPED & PELLETED STRAW

Fuel	Cost per t £	MJ/t	Cost per '000 MJ £	Eff. of Combustion %	Cost per '000 MJ Effective £
Oil	100	40,000	2.50	80	3.13
Coal	60	30,000	2.00	70	2.86
Chopped Straw	20	16,000	1.25	60	2.08
Pelleted Straw	35	16,000	2.19	75	2.92

If it is necessary to compress the bales for storage purposes or to lower transport costs for eventual sale, a further £0.55 per 1000 MJ would have to be added. To date, hydraulic bale compression machines which reduce them to approximately one third of their original length, have only been developed for 46 x 36 x 92 cm straw bales. Such levels of densification can be accomplished with comparatively little force however, if further densification is required, the forces involved rapidly escalate (Fig 2) (Cranfield Institute of Technology 1982). Unfortunately, compacted bales do not lead to greater combustion efficiency but they do enable furnaces to be stoked at larger intervals, thereby effecting a small saving in labour cost.

It would appear that a majority of field baling is now by big balers. In 1978, only the Hesston producing either 1.4 or 1.8 m diameter x 1.5 m cylindrical bales or 1.5 x 1.5 x 2.33 rectangular bales were being used. Today, Claas, Deutz-Fahr, John Deere, Gallignani, Sperry New Holland and Welger have entered the market with cylindrical or rectangular products varying in weight from 350 - 573 kg and densities from 200 - 350 kg/m³.

It is found that even for a large farmhouse, one of these 350 kg bales, fed into a boiler once a day by a tractor fore-end loader, is sufficient to heat the house as it will supply an average of 875 kWh if the overall efficiency of the unit is 60% (STANIFORTH 1982).

The latest range of straw-fired furnaces incorporates methods of automatic stoking to eliminate the chore of frequent manual stoking although the bale magazine will normally be loaded by hand; these boilers are thermostatically controlled and incorporate fan assisted draft features to improve on the combustion efficiencies of the earlier models. Efficiencies in excess of 60% are achieved but this has a cost - an automatic system

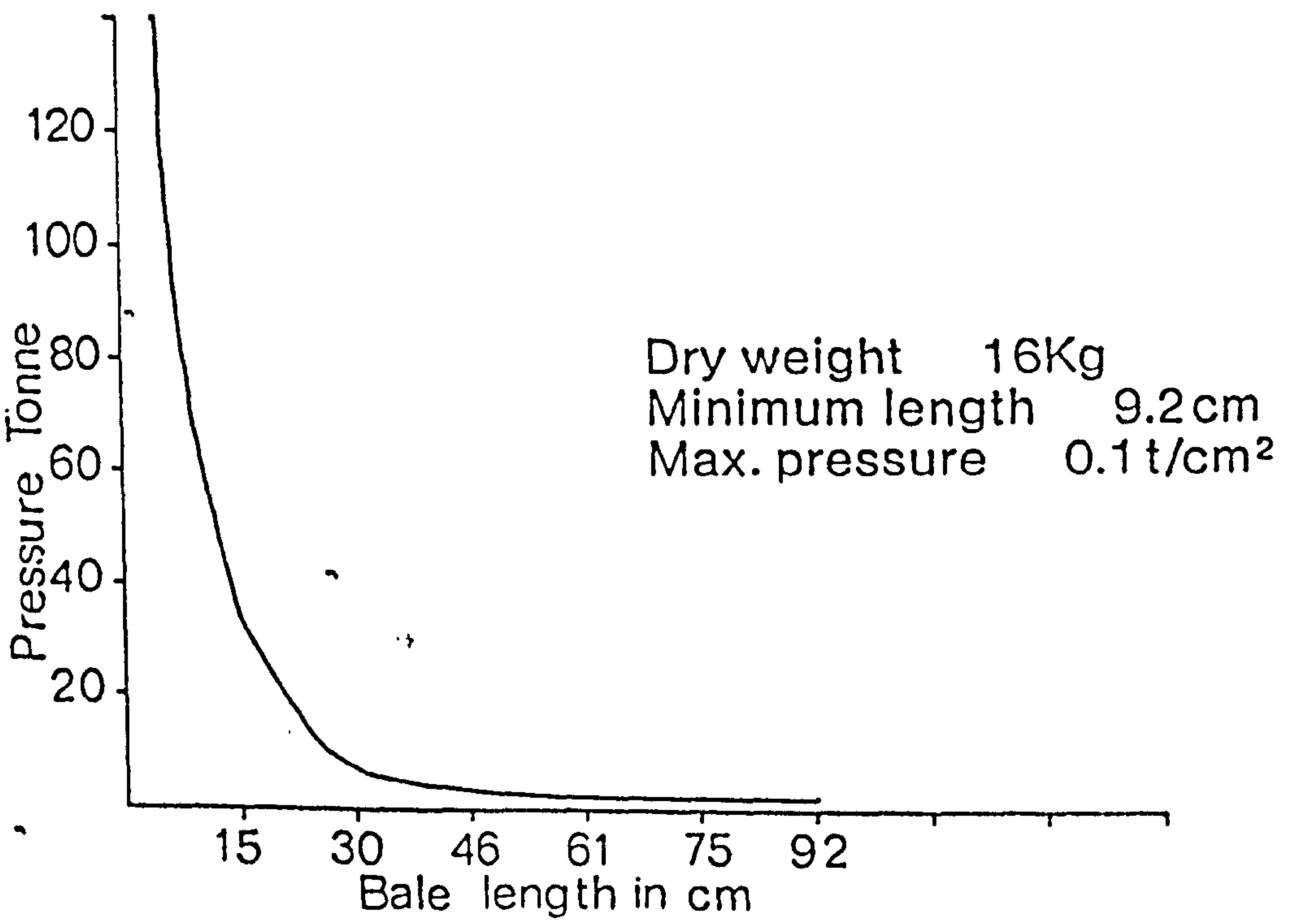


Fig. 2. Compression of straw bale 46 × 36 × 92cm.

SOURCE - STANIFORTH 1982

could add a further £5000 to the cost of the boiler. Writing this off over ten years at 12% interest would add approximately £1 per 1000 MJ to the cost of running a heating system. There is also an additional electricity cost to be borne, amounting to approximately £0.50 per 1000 MJ. This makes a total extra cost of £1.50 per 1000 MJ for an automatic feeding system.

When the work started on this thesis in 1981, oil prices were some 50% higher than they are now which presently lessens the advantage of burning straw as a fuel. It is thought that this situation will not continue indefinitely and it is expected that we will eventually return to something approaching the relative fuel costs of 1981: these were £5.00, £2.66 and £1.25 per 1000 MJ for oil, coal and straw respectively. The fuel prices shown however, refer to their cost in 1986. As the cost per tonne of fuel, its quality and the efficiency of its combustion, together with the increased price of a boiler is changed, straw competitiveness against coal and oil will change. Table 8a shows the cost of the fuels, the oil and coal at generally accepted farm prices and the straw at the price the farmer could sell his bales. This relationship only holds good for air dry straw, if the moisture content of the straw is increased from 15 to 27% (reducing its heating value to 12 MJ/kg) the cost of energy produced from the straw would increase by one third.

There are two other forms of compaction - briquetting and pelletising - which respectively require the straw to be chopped or hammer milled before it is fed into the presses. Both chopping and milling have high specific energy requirements.

BRIQUETTING

A briquetting press comprises a hollow cylinder fitted at one end with a tapered die. A reciprocating piston rams plugs of chopped straw into the die and solid briquettes are extruded. The binding of the material is achieved by the combined effects of heat and pressure, both generated by friction. However, unless certain conditions are maintained, the briquettes will be unstable and will readily break down. If this happens, they become almost impossible to transport or store. It has been found that for stable briquettes, the straw must be within the moisture content range of 8-14%, which means that it normally needs some form of drying before treatment. If the straw is too wet, free water acts as a lubricant limiting the pressures that can be developed in the die, and it also causes briquette fracturing; if it is too dry, the fibres are too elastic and they will not bind together. Briquettes must be retained under compression while they cool in order to prevent the fracturing which results from high thermal stresses. Briquettes absorb moisture easily and therefore dry storage is essential to prevent disintegration.

PELLETING

Ring die presses are normally used to process straw into pellets and again the moisture content of the straw is critical, 12% being the maximum level. Because the straw has had to be finely ground, after extrusion, the pellet does not require any further compaction and it binds together making storage in bins or bulk relatively simple.

As a general rule, a tub grinder reducing bales to lengths of straw of 2-4 cm needs 100 H.P. for a capacity of 4 tonnes per hour. For briquetting, a further 100 H.P. is required for operation at 2 tonnes per hour. Hammer milling at a rate of 2 tonnes per hour through a 3 mm screen prior to pelleting, also needs 100 H.P. then pelleting at the same rate needs 50 H.P.

The combustion efficiency of straw is increased from 60% to 75% by pelleting but the cost to do so is increased from the initial straw price of £20/tonne to that of £35/tonne for pellets (Table 8b). This increase gives a cost per 1000 MJ of (effective) £2.92 which is 93% of the cost of the equivalent heat from gas oil (Table 8c).

Stable briquettes and pellets can be handled and stored in bulk; they have a relatively high density which helps when they have to be transported and they can both be fed into a boiler by an automatic stoker.

COMBUSTION SYSTEMS

Increasing the efficiency of a boiler will reduce the running costs but the additional capital outlay on a small boiler when writing off over ten years at 12% interest, would make it uneconomic. For larger installations, however, automatic stoking systems would be necessary and cost and payback of capital including saving of labour costs would have to be seriously considered. Whilst automatic stoking systems are expensive, and as previously shown cannot be justified economically for boilers of a relatively small output, such systems improve the efficiency of combustion and a furnace previously rated at 60% if increased to 75% would reduce the effective cost per 1000 MJ by 20%. (Repayment of capital and interest charged based on standard amortisation tables from Wye College).

In 1980 in Denmark, the municipality of Svendberg provided hot water to 6000 people by using three sources of energy; household refuse available all through the year provided 30% of the total energy requirement; straw, for a period of nine months which excluded the three summer months, contributed a further 50% while oil, used only during the coldest weather, made up the remaining 20%. The straw was ground to lengths of 8-12 mm and blown into silos from which it was augered out under thermostatic control, and fed by blower to the boiler at a rate of three tonnes per hour. This corresponded to about one tonne of fuel oil which cost approximately £190 per tonne at that time.

With straw varying from £17 per tonne at harvest to £20 per tonne in March/April, and the boiler efficiency calculated by the Danish Association of Boiler Manufacturers at 78%, the project was judged to be sufficiently viable and cost saving for the Danish Government to install a similar unit in Aarhus in 1982. The straw in both cases was in bales, although for ease of handling, straw pellets were preferred; pellets however, would have made the project uneconomic.

In the short term in the U.K., the most obvious outlet for straw as a fuel is for on-farm applications such as heating farmhouses and glasshouses and for crop drying. Silsoe College has reported that the maximum utilisation of straw on farms could be between 1.0 and 2.1 million tonnes (0.5 to 1.1 mt. coal equivalent (c.e.)) by the year 2000. (SILSOE COLLEGE 1984).

Because of the rise in coal prices relative to the price of straw during the last few years and also because of the increased availability of straw brought about by restrictions on straw and stubble burning, its use as a fuel could become increasingly attractive in rural/industrial applications. Silsoe College have studied industries located within straw producing areas and have estimated the potential for replacing coal and oil by straw. Their findings suggest that a maximum of 4 million tonnes of straw (2.2 mt.c.e.) could be used in boilers and 420,000t (0.22 mt.c.e.) in furnaces. (SILSOE COLLEGE 1984).

It is estimated that some 300,000 tonnes (0.16 mt.c.e.) of this potential could be realised by the year 2000 mainly in the malting, sugar beet processing, cement, brick and aggregate industries. Silsoe also suggest that up to 200,000 tonnes (0.11 mt.c.e.) could be used in the institutional/ commercial sector by the same date. (M.A.F.F. 1984).

DEMONSTRATION UNIT

A 6.8 M.W. straw fired cyclone furnace is being installed to provide heat for drying chalk in Suffolk, the plant being situated in a large straw producing area (PLASKETT & NICHOLSON 1984). Such units are not without problems, so means other than cyclone firing need to be found to burn unprocessed straw. One possibility would be to use a fluidised bed combustion unit; these are now used with considerable success for burning high ash, poor quality fuels and it was felt that this system could perhaps be adapted to burn chopped straw.

It was therefore decided to build a fluidised bed combustor with an in-bed feeding system to burn chopped straw.

CHAPTER 1

REVIEW OF EXISTING BIOMASS FUELLED FURNACES AND BOILERS

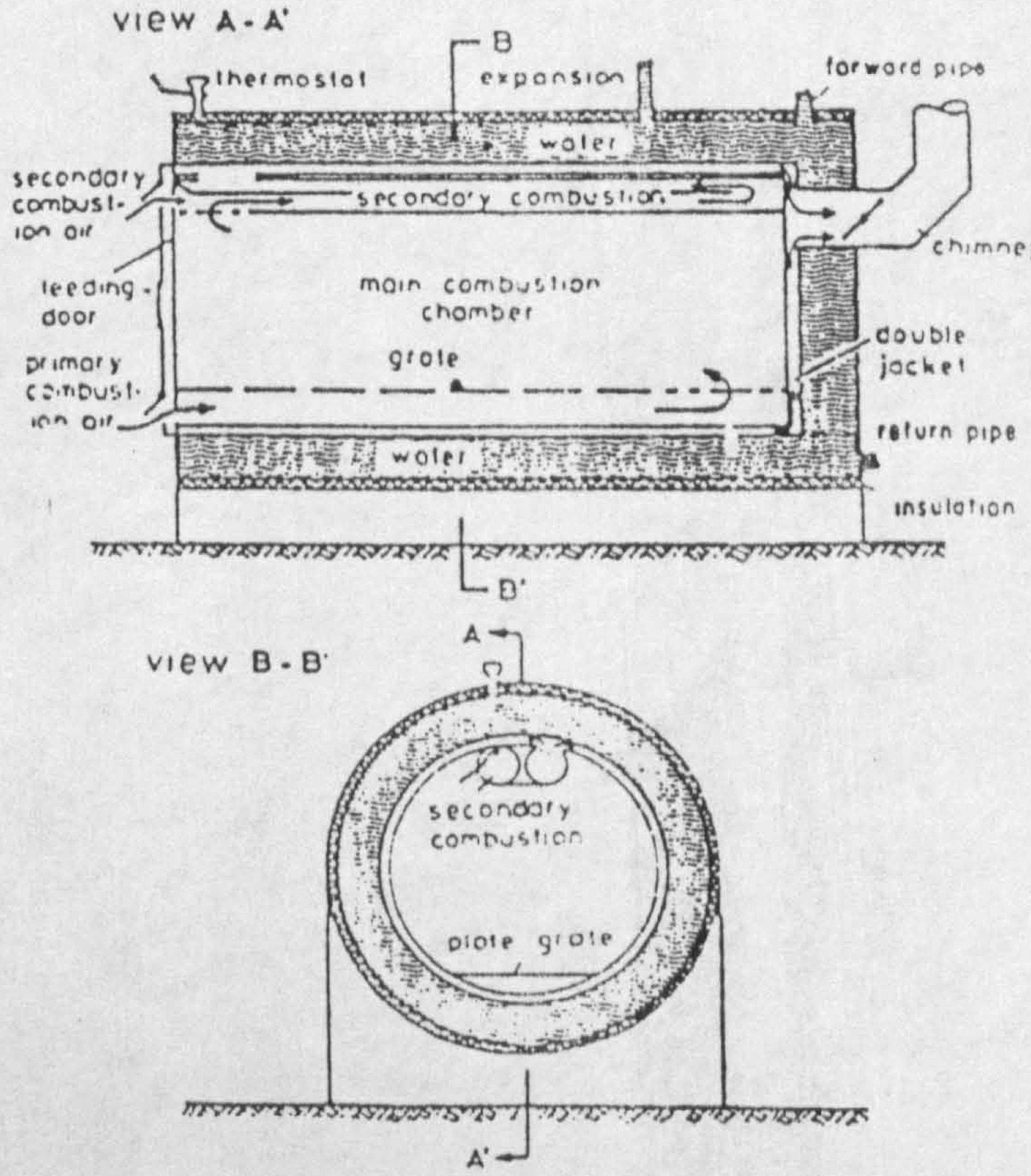
The designs for the majority of the straw burning boilers in the United Kingdom originated in Scandinavia as those of waste wood burners. These units utilised residues from tree felling and they are still being used for that purpose in Scandinavia and in parts of Europe. The boilers consisted of a horizontal cylindrical chamber surrounded by a water jacket, the wood (or bale) sitting on a grate with the combustion air being regulated under it by an adjustable inlet in the bottom of the end fire door.

The original units were manually fired by opening the fire door and replacing by wood or straw bales the burned out charge. Little regard was given to the need for efficient combustion, in some cases the air supply was fixed throughout while on slightly more sophisticated models, the position of the air regulator was controlled by a bi-metallic strip.

Later models, such as those manufactured by LOIBL (fig. 3) incorporated a secondary air inlet at the top of the chamber for better control, a secondary chamber and a water cooled grate.

Over the years, research work has been mainly concerned with the automation of the straw feed to the boiler, and towards improving the efficiency of combustion. The intention has been to produce boiler plant that would be acceptable for commercial purposes as well as to improve units designed for the agricultural market.

LIN-KA Maskinfabrik of Denmark have developed an automatic feeding system in which the fire door is replaced by a sliding door that is operated hydraulically. Small bales are placed on a conveyor at head height then allowed to fall by gravity down a 50° slope until they reach the



Throughburning boiler for high pressure bales with double jacket and secondary combustion

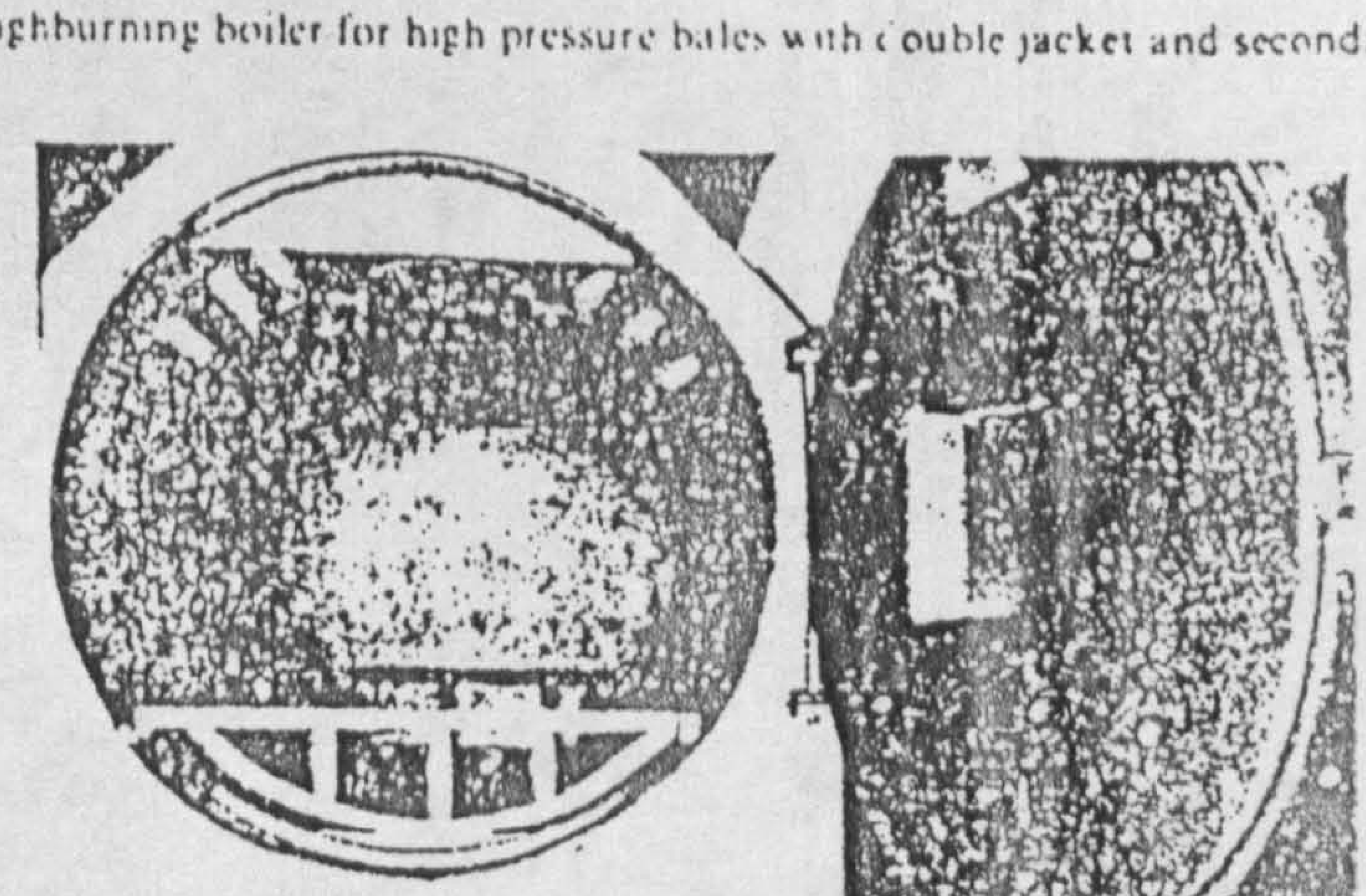


Fig. 3 Combustion chamber of Loibl furnace

sliding door. As the door opens, a horizontal ram automatically extends and pushes the bale into the combustion chamber. (Fig 4). The ram retracts, the door closes and a further bale is moved by the conveyor to the ramp ready for the next insertion. While this arrangement overcomes the chore of frequent manual stoking, it is still essentially a batch stoking system and so has an inherently low efficiency.

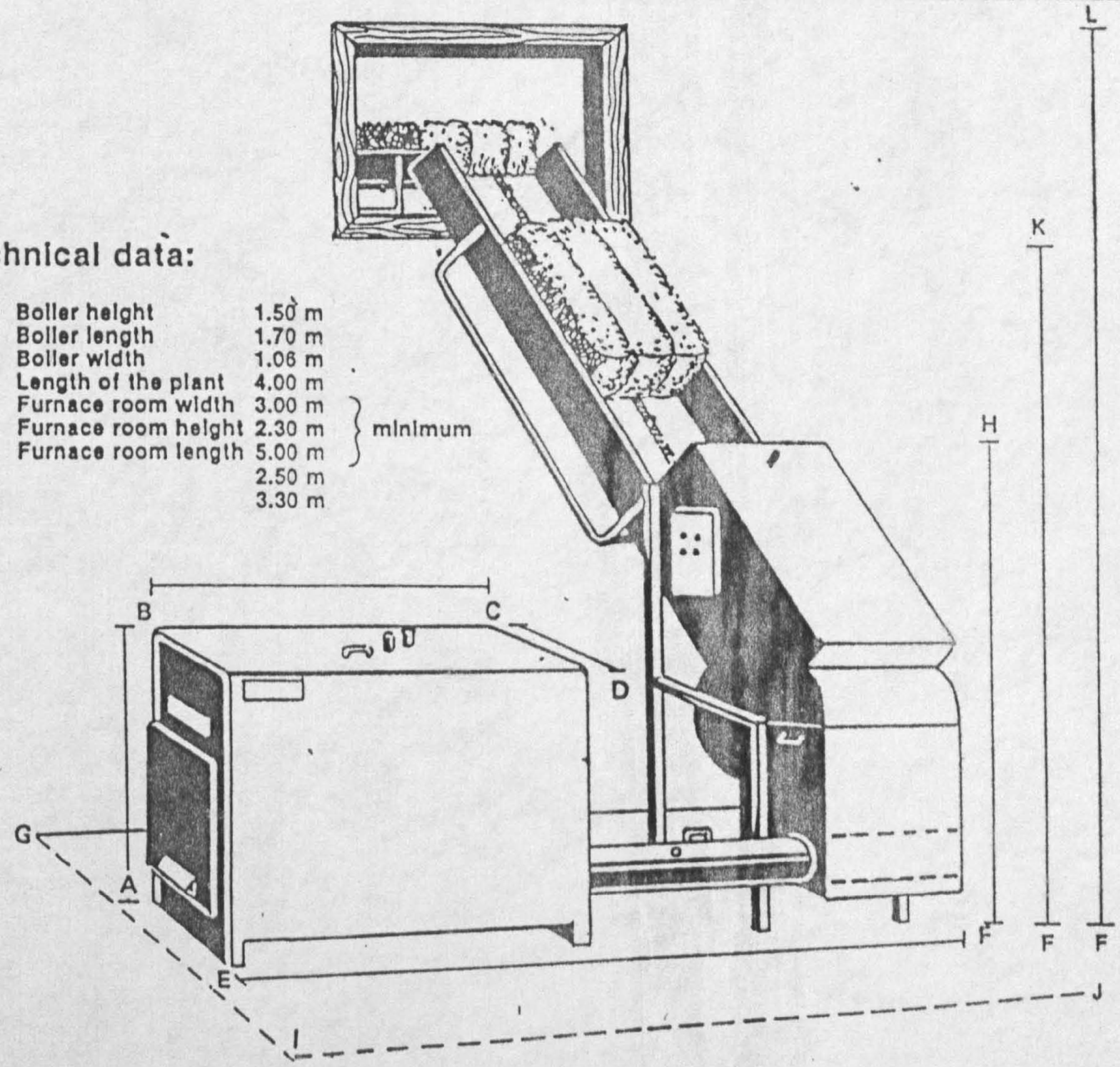
PASSAT of Denmark have a different automatic feeding system in which a conveyor feeds bales into a grinder or chopper. The straw particles produced fall on to a screw conveyor which feeds them into the furnace (Fig 5). Similar types of equipment are also manufactured by RONTGEN (Fig 6) and SCHMIT (Fig 7), and since they are truly continuously stoked units, they have a higher potential efficiency than the simpler batch fed units.

In France the heating company SELF-CLIMAT, and ROULIN, manufacturer of grain driers, have together developed what they claim is a straw burning furnace. Straw, however, accounts for only three-quarters of the energy supply, the remainder being provided by either oil or liquid petroleum gas (L.P.G.) Since it is easier to control the heat output from the liquid fuels they are used to 'top up' the base load provided by the straw. With this system, about 5 - 6 bales are burning at any one time and an adjustable time clock on the feeding conveyor allows a bale to pass through an air-lock into the furnace every 8 - 12 minutes. This feed rate gives an output of around 1300MJ/hr which is sufficient to evaporate 800 - 900 kg of water per hour in the attached grain drying unit (Fig 8).

The PILLARD company have manufactured a cyclone suspension burner for crop residues with a claimed output temperature of up to 1500°C.

Technical data:

A-B	Boiler height	1.50 m	
B-C	Boiler length	1.70 m	
C-D	Boiler width	1.06 m	
E-F	Length of the plant	4.00 m	
G-I	Furnace room width	3.00 m	minimum
H-F	Furnace room height	2.30 m	minimum
I-J	Furnace room length	5.00 m	minimum
K-F		2.50 m	
L-F		3.30 m	



Linka also supplies a fully automatic straw burner designed for caloric consumptions of up to 2,000,000 B.T.U., or more. It uses whole bales and is controlled by the same advanced components as those used for Linka "Little Sister". Ask for our special brochure.

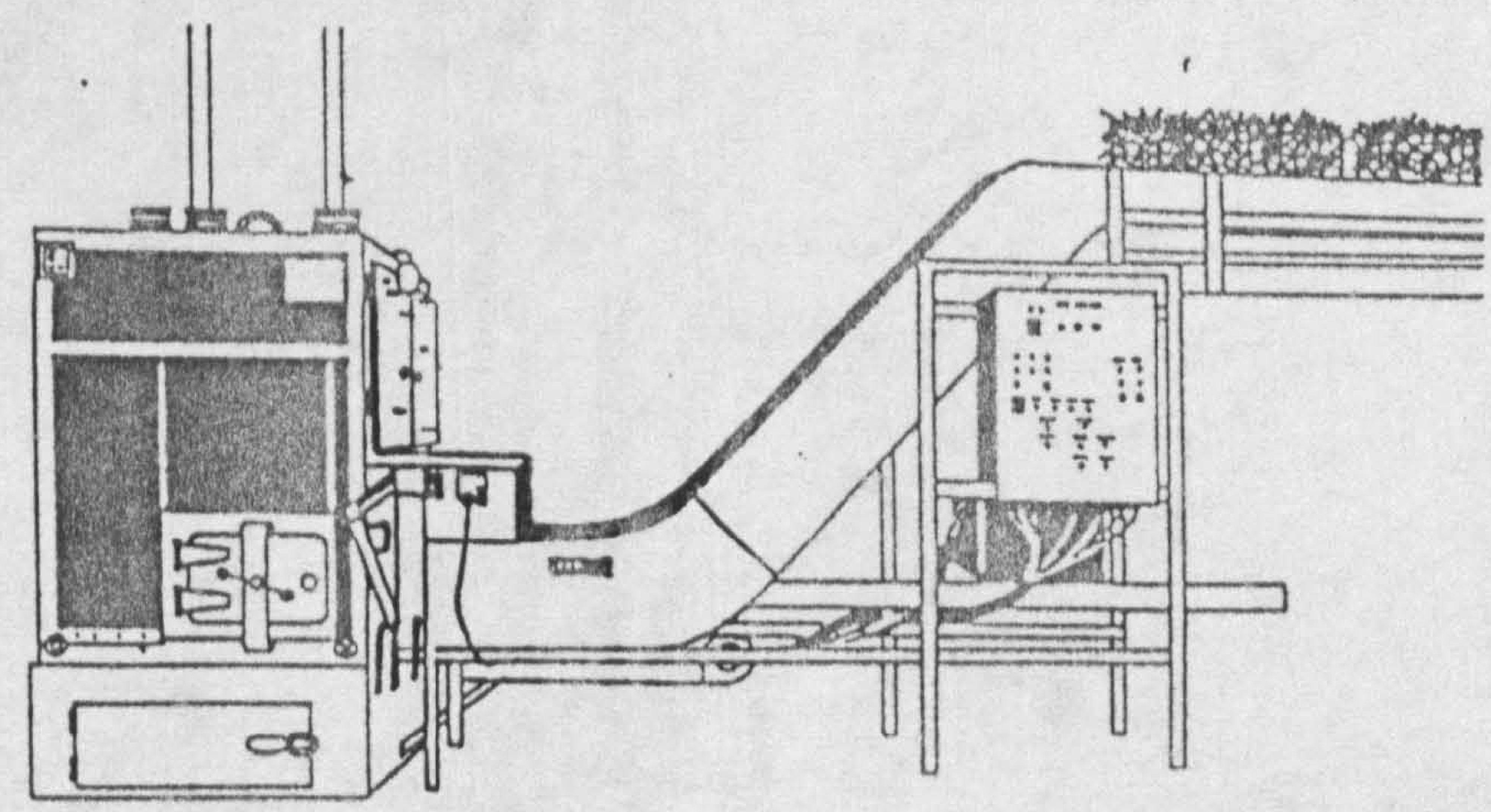


FIGURE 4.

LIN-KA

LIN-KA MASKINFABRIK . DK-6940 LEM . DENMARK . TELEPHONE (07) 34 16 55

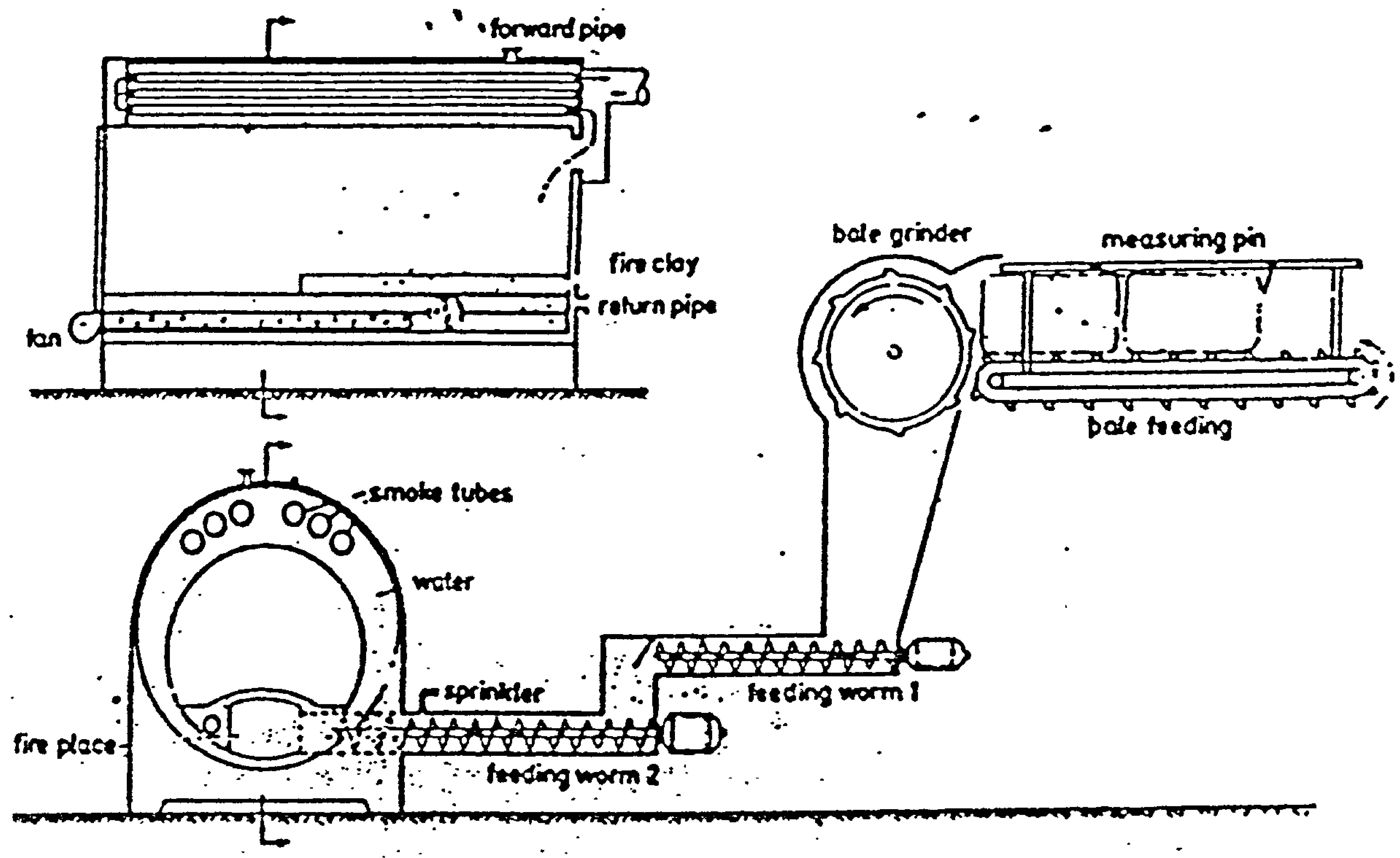


Fig. 5 Straw furnace with automatic feeding system (Fa Passat, Ørum, Denmark).

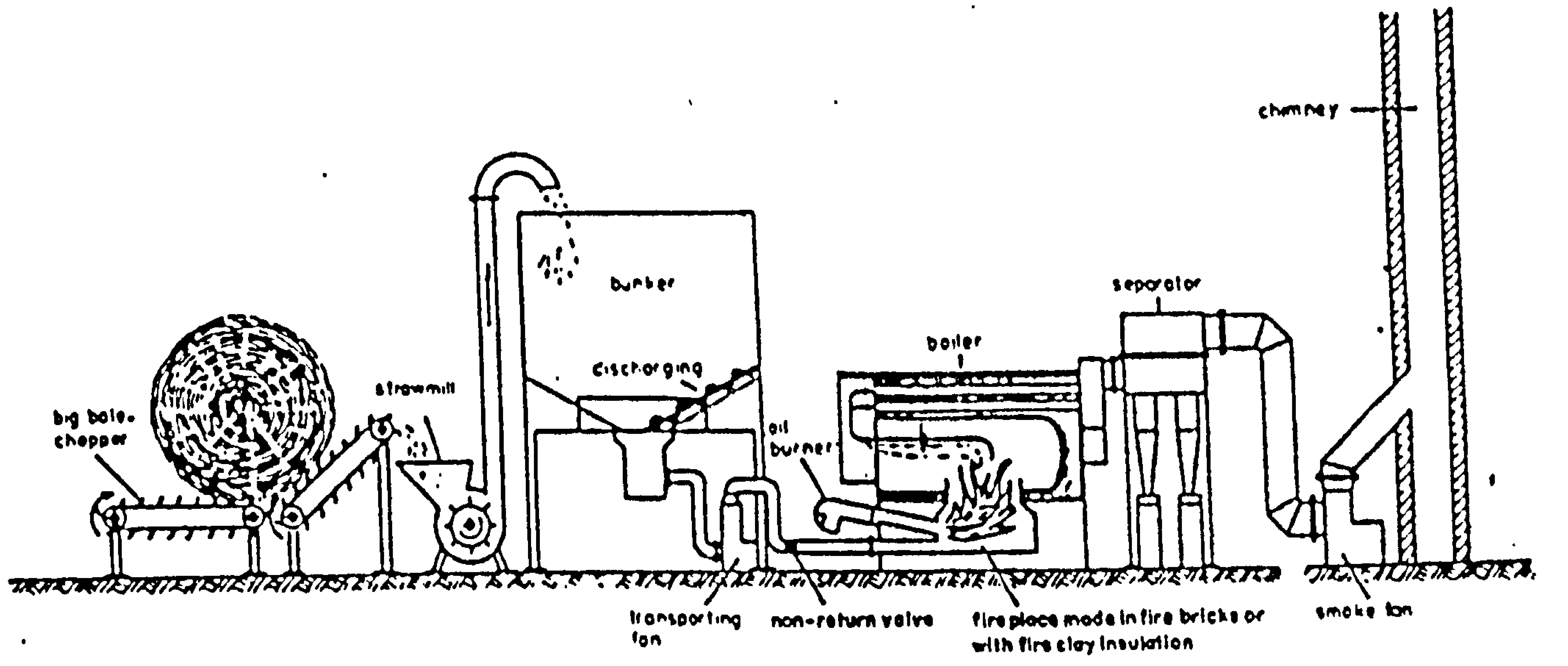


Fig. 6 Automatically stoked furnace with auger and fan feed system and straw chopping plant. (Fa. Rontgen & Co.)

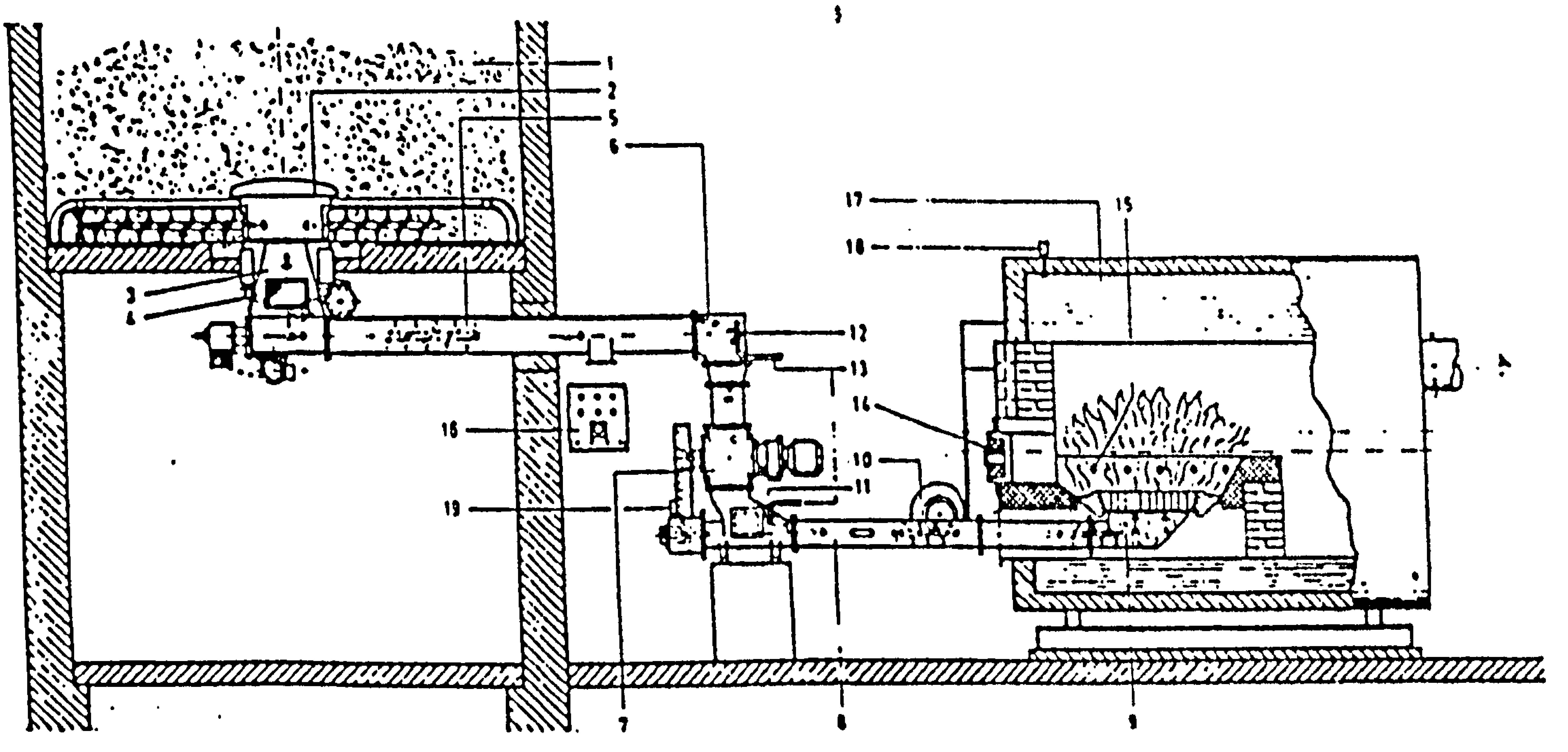


Fig. 7 Automatically stoked boiler with auger feed system for wood chips or chopped straw. (Feeding system: Fa. Schmidt. Boiler: Fa Loos.) 1 = Chip-bunker, 2 = discharging plant, 3 = collecting box, 4 = indicator for rate of emission, 5 = feeding worm with regulatable drive, 6 = double knee with security flap, 7 = rotary cellular sluice, 8 = feeding worm, 9 = fireplace with twyer box, 10 = combustion air fan, 11 = primary fire extinguisher, 12 = secondary fire extinguisher, 13 = water lead, 14 = fire hole, 15 = secondary combustion air blast pipe, 16 = switch board, 17 = boiler, 18 = thermostat, 19 = speed control.

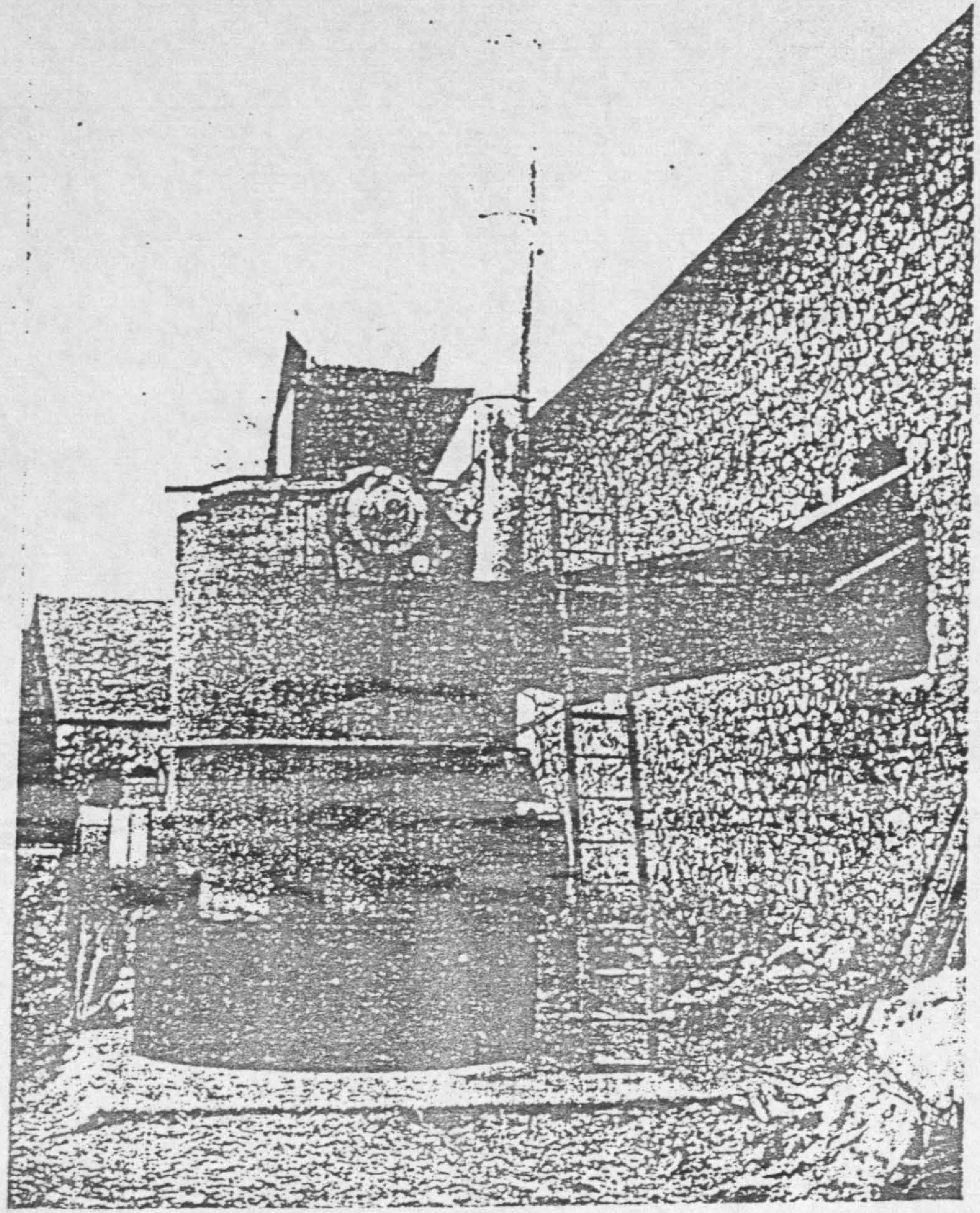


Figure 8 SELF-CLIMAT-ROULIN straw bale furnace connected with a grain dryer (inside the barn)

Although it can be used with an air/air heat exchanger for grain drying, it has been built mainly for lucerne drying plants, of which there are between 100 - 150 in operation. After the lucerne drying season, these plants, mainly owned by cooperatives, dry sugar beet pulp giving about a nine month operating season (Fig 9).

VERNON-MERVER have also produced a complete straw burning system for lucerne drying which includes a regulating feeder for chopped straw. This unit has pneumatic loading and a combustion furnace of the suspension type coupled to a mixing chamber in which furnace gases are mixed with a regulated supply of exhaust recycled gases for temperature control. Several sizes of plant have been built ranging from 16,000 - 50,000 MJ/hr (Fig 10).

In North America, developments have tended to concentrate on gasification of biomass and most of the commercial units are of the relatively cheap fixed bed updraft type (Fig 11). They are coupled to existing boilers which then only require a burner modification or replacement. When used originally as a heat source for commercial drying plants, it was mainly for wood or lime kilns, although current emphasis is towards the agricultural market where in the main, gas oil or natural gas have been used as fuels.

There are two basic gasifier designs - updraft and downdraft - (Fig12). In updraft units, the hot gases flow counter to the feedstock, whereas in the downdraft units, pyrolysis products are broken down as they pass through the reaction zone before combining with the exiting gases. Some downdraft gasifiers have the potential to eliminate tar from the gas and they are probably the best suited for the burning of crop residues as a fuel.

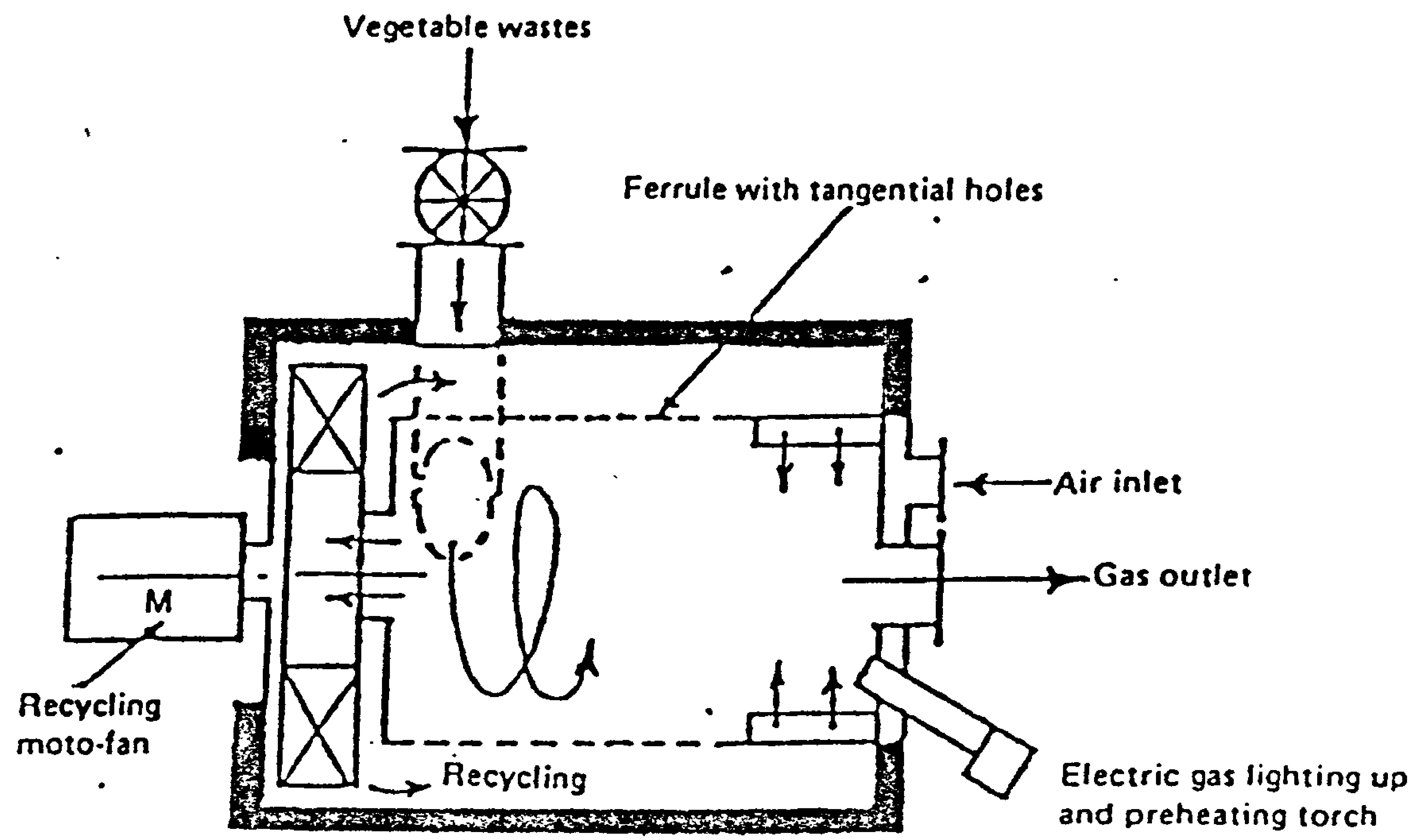


Figure 9 PILLARD "suspension" gas producer for fine granulometry crop residues

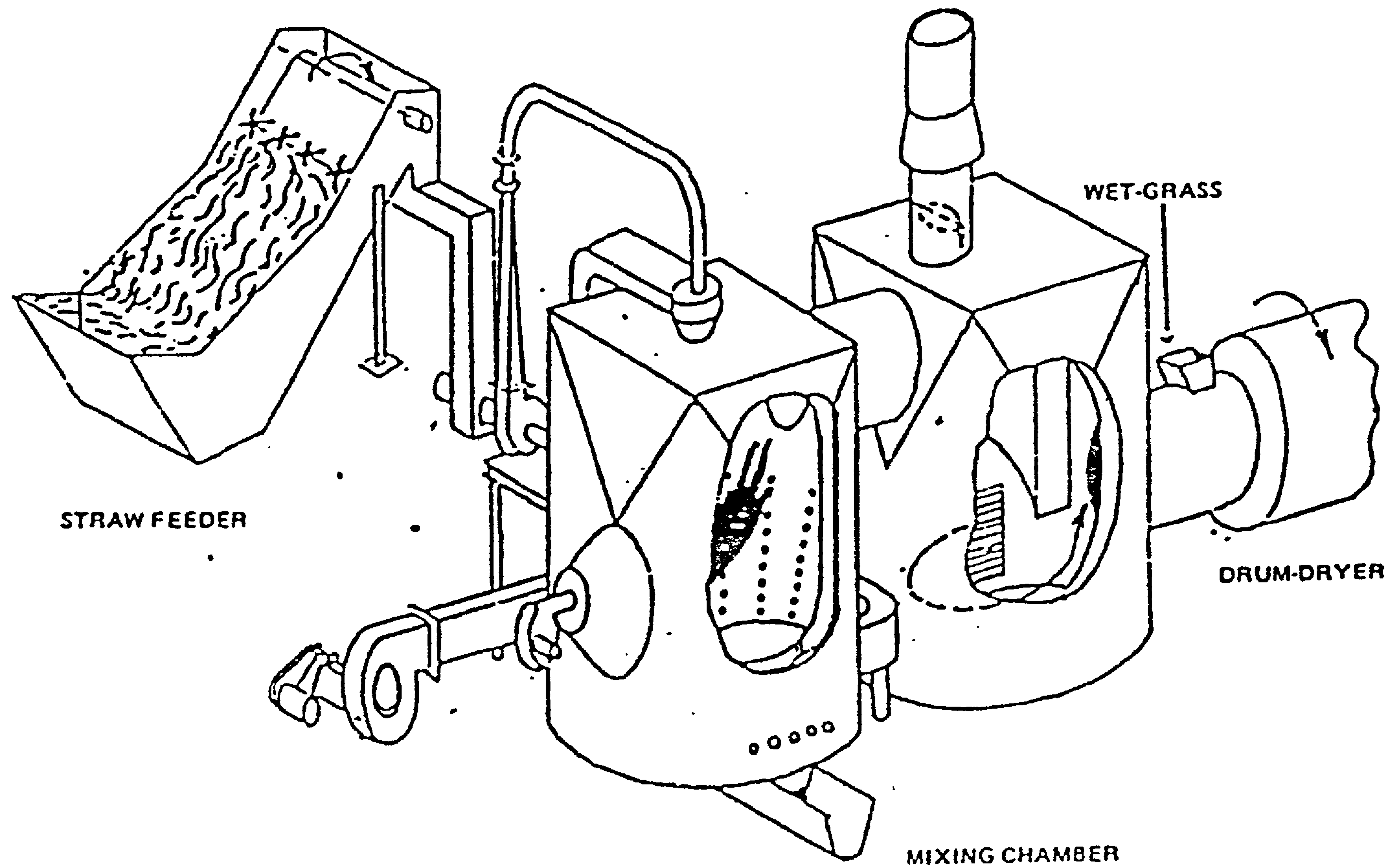


Figure 10 Combustion chamber

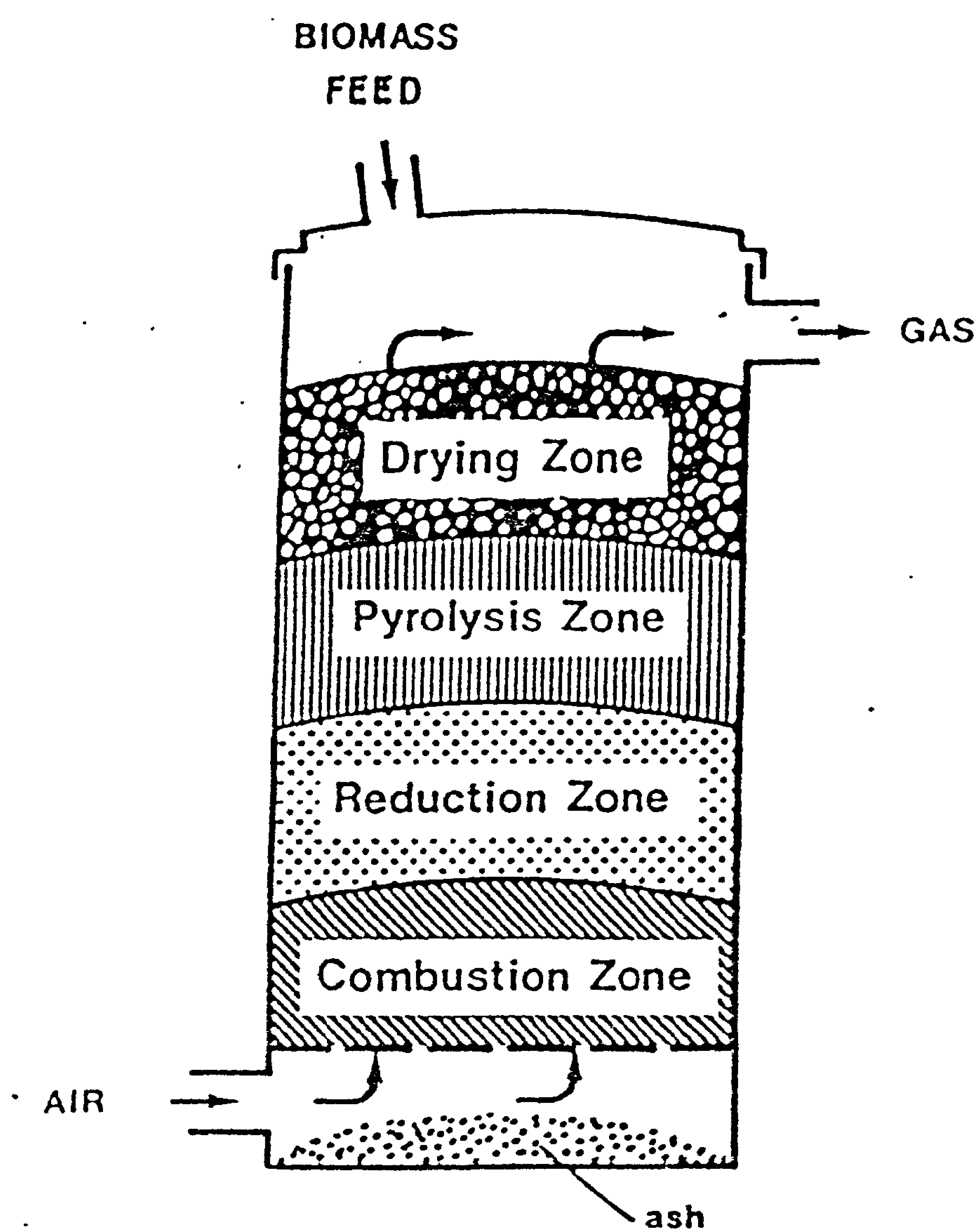
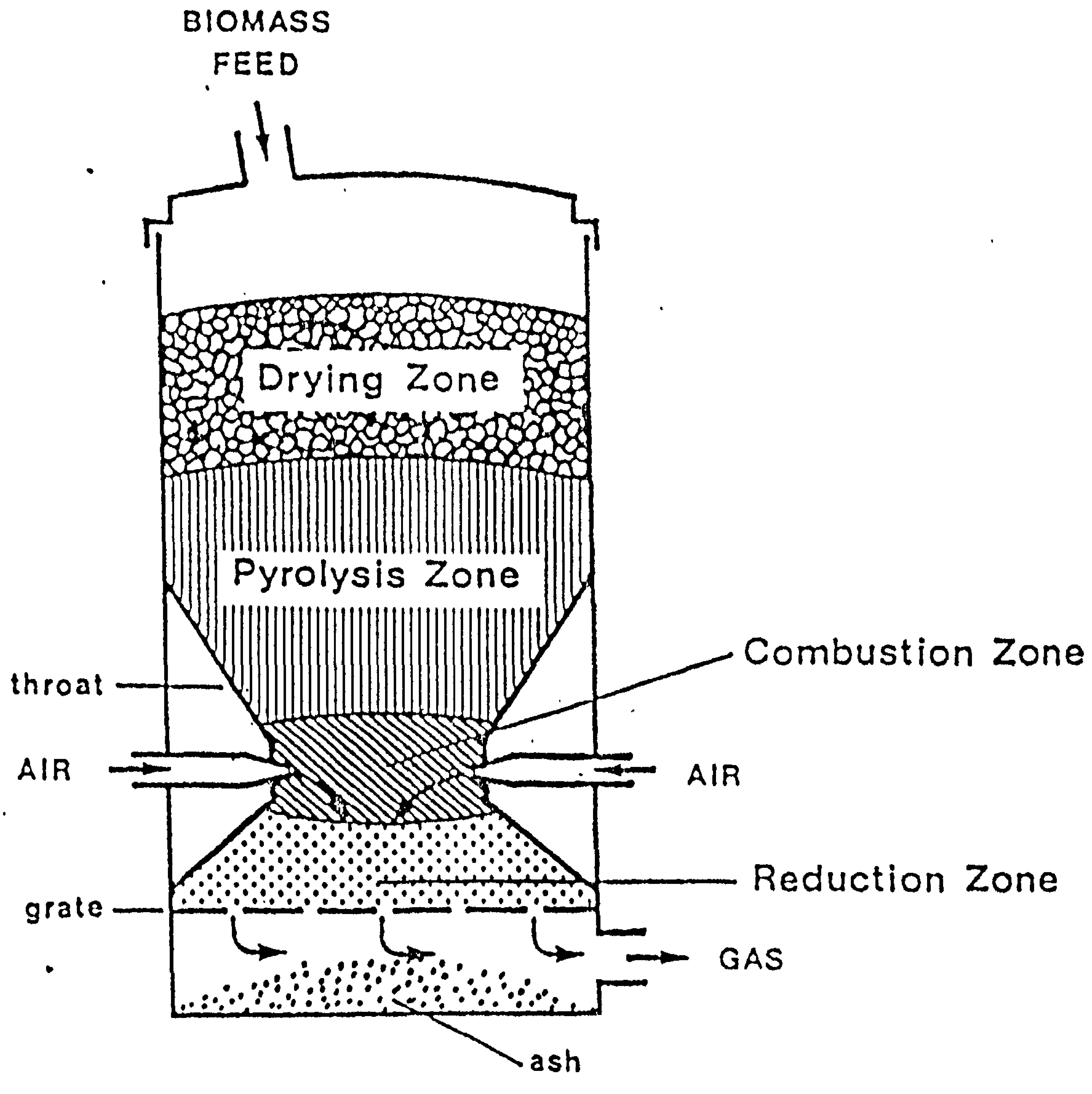


FIG 11 Schematic Diagram of an Updraught Gasifier



SCHEMATIC DIAGRAM OF A DOWNDRAFT GASIFIER

FIGURE 12

University research in the U.S. is under way to develop suitable combustion systems for wet biomass, and also to study optimum fuel particle size for various gasifier feeding systems, particulate control in the products and the effects of different biomass mixtures on performance. At Iowa State University, a direct combination concentric-vortex cell furnace to burn agricultural residues has been developed and connected to a crop dryer. (CLARR et al 1981) (Fig 13). During prolonged tests in the 1981 season, it was discovered that this type of equipment had severe limitations from an environmental point of view. It produced excessive amounts of smoke that has an unacceptable smell and also when it was burning wood chips, it was a definite fire hazard (WAHBY et al 1981).

At Purdue University, a greater degree of success was achieved with a downdraft channel gasifier furnace (RICHEY et al 1981); the installation of a secondary combustion chamber resulted in temperatures above 1350°C with complete combustion and the hot gases were diluted with ambient air for drying corn.

For curing tobacco, workers at Clemson University opted for an updraft gasifier whose flue gases were piped to water tanks; these acted as thermal energy reservoirs and, as required, hot water was pumped to the curing barns. The updraft system was selected because it was considered the simplest to design and construct, although it was realised that a tar-laden gas would be produced. It was argued that in a close-coupled combustion system, with the gas burned immediately after being produced, the presence of tar would be of little or no concern. The system worked well for a limited period whilst the initial fuel charge was burning, but during operation it was found to be impossible to feed from the top while there was a partial load of fuel. This was due

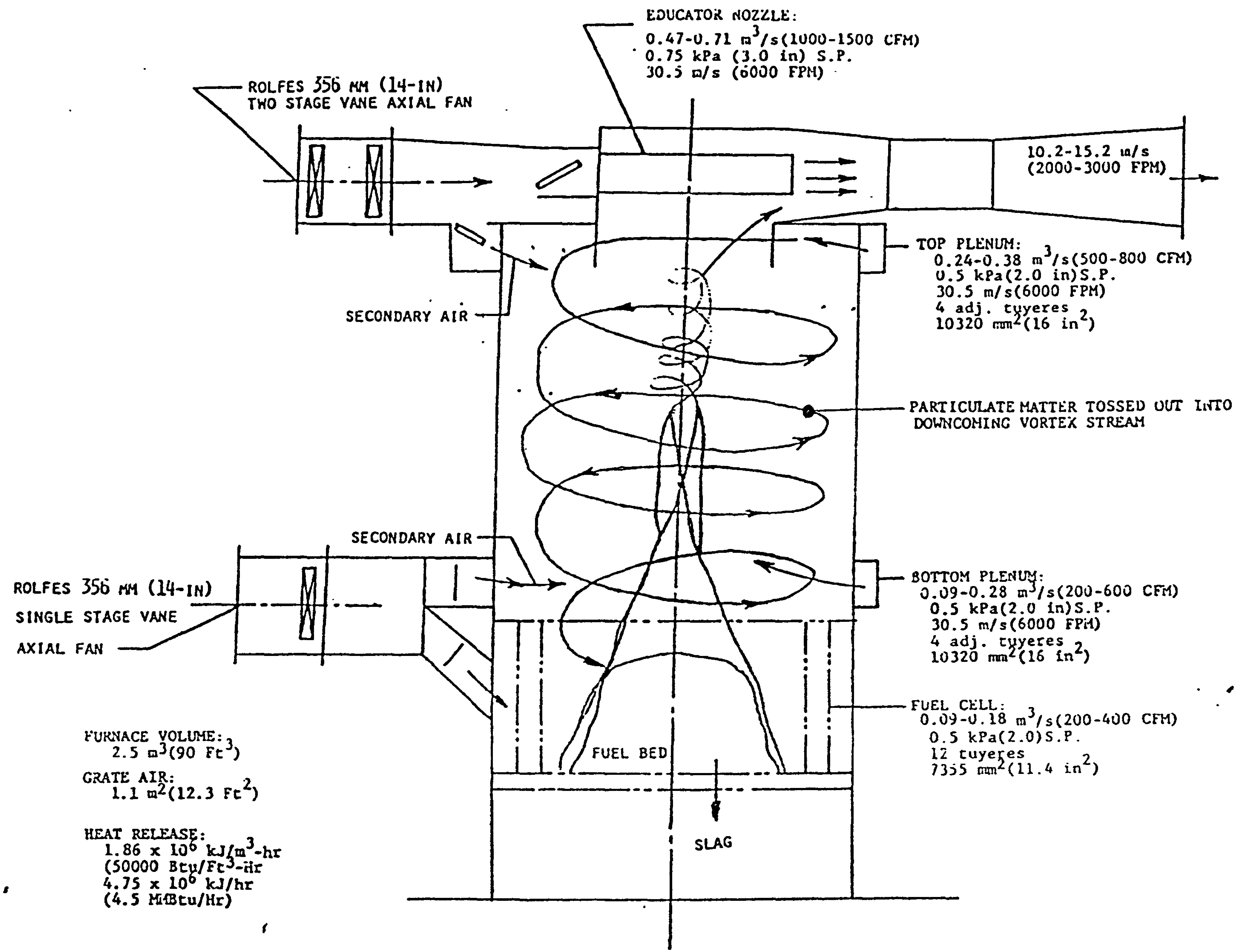


Fig. 13 Schematic Diagram of the Prototype I Concentric-Vortex, Cell Furnace

to the intensity of the radiant heat from the bricks at the top of the chamber (PAYNE et al 1981). A completely new top-feeding design was being sought but no details have yet been made available.

Research work in the U.K. has been mainly at University level where at Nottingham, stepped grate furnaces for burning chopped straw for drying purposes in a whole crop harvesting system have been developed (WILTON 1981).

At Cambridge University, whilst working on fluidised combustion of char and volatiles of coal, some experiments were carried out with straw pellets and wafers. The results are shown in Appendix 3 (TURNBULL 1983).

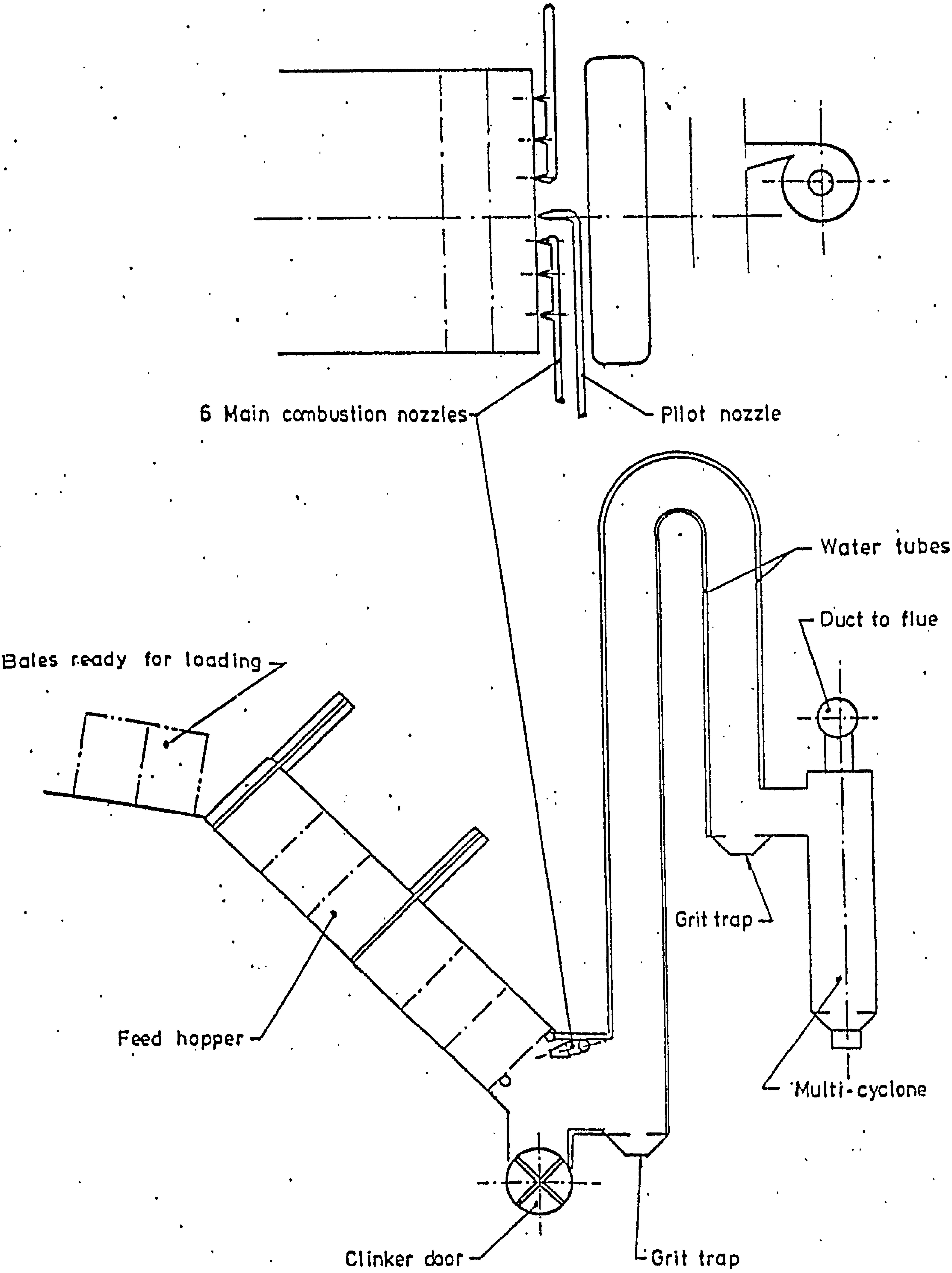
Work on cyclone combustion of poor quality fuels was undertaken at University College, Cardiff and described in a paper by SYRED & SAHATIMEHR (1983). Two types of combustor were reviewed - the multi-inlet Agrest type which was originally developed for burning vegetable waste, and the Compact combustor developed from the standard dust collecting cyclone. Both operate on similar principles although the internal flow patterns of the two are different. Both have relatively long residence times at high temperatures for the fuel particles, especially the multi-inlet type, and unacceptably high NOx emissions are created.

Designs have been put forward, but never built, for a downjet combustor for burning bales (THRING & CROOKES 1980). This is a system in which the whole of the combustion air is blown into a surface of the bale through a nozzle with a variable aperture; this would allow the velocity of the air jet to be varied while the quantity was kept constant. By this method, the depth of penetration of the gases into the fuel bed could be varied, and thus the contact between gases and fuel adjusted to the optimum. Bales would be pushed against a barrier of water-cooled tubes between which the combustion gases would flow, or they could be fed by gravity down a 45° slope (Fig 14). It was claimed that with the downjet system, the bale of straw would burn steadily through (like a candle) and the volatiles would burn at the same time as the residue charcoal. The ash would be removed by a continuous belt grate and dumped into a bunker.

A feasibility study has been carried out by the Energy Technology Support Unit (E.T.S.U.) at Harwell, entitled the 'Potential for Large Scale Projects Featuring Straw as a Fuel' (MARTINDALE 1982). This study has been partially carried out to establish the viability of installing a straw-fired boiler at a maltings located in a grain producing area (SMITH 1982). Two options were considered; suspension firing in a water-tube boiler or a cyclonic firing in a combustor mounted externally to a shell boiler.

The first option involves blowing finely chopped straw into the furnace above a fixed grate; some 60 - 70% burns in suspension, the remainder falling onto the grate where combustion is completed on a coal bed. An efficiency of 71% is claimed with complete volatile burn-off, and with low gas velocities there is negligible straw or char carryover. However, it was estimated that in order to maintain output and keep an incandescent bed of coal, up to 30% of the fuel would need to be coal - a factor which reduces the potential for saving fuel costs.

The second option, a cyclonic combustor, consists of a cylindrical refractory furnace into which air and chopped straw are fed tangentially at high velocity to create a downward vortex around the perimeter of the vessel. Ignition is carried out by natural gas burners and the gases follow a normal cyclonic action, turning at the bottom of the vessel and passing back up through the centre of the vortex to the exit. Again, very high efficiency is claimed - around 67% - but a second fuel, natural gas, has to be used.



DOWNJET COMBUSTOR

FIGURE 14

In Lincolnshire, vegetables are being grown in polythene tunnels heated by hot water supplied by a straw burning boiler. In one case, an 800,000 BTU/W boiler (which normally would consume 27 - 32 litres of oil per hour) was installed. This holds two large cylindrical bales (equivalent to 60 small bales) at one stoking. Although a large capacity water jacket holding 55m^3 was built in to accomodate fluctuations in heat output and demand, even this proved unable to maintain a continuous supply of hot water. The consequent loss of 25% of the crop was attributed to the failure of the unit to burn bales evenly (R.C.BALLS 1982).

When bales are burned by simply placing them into a combustion chamber, the loose outer layers burn rapidly, leaving a dense inner mass masked by a layer of ash. This makes it difficult to obtain efficient combustion.

Near Orleans in France, a small boiler was developed to provide heat to a greenhouse of 2000 m^2 at a small farm. It was fired by straw pellets quite successfully, but after taking into account the total costs of producing the pellets, it was cheaper to heat with conventional fuel oil (HURAND 1978).

Although the French Government estimate that by the construction of decentralised straw pellet production plants, which will utilise material from a maximum radius of 20 kilometres, between 3 - 5 million tonnes of straw will be used as a fuel by 1990, (REQUILLART 1982), much more could be utilised if the additional processing cost could be eliminated.

CHAPTER 2

HISTORICAL REVIEW OF FLUIDISED BED PROCESSES

The earliest application of fluidisation was unearthed by BROTZ (1952) and refers to the purification of ores in the sixteenth century. After this process was described by AGRICOLA (1556),there was little further documentation until the nineteenth century when U.S. patents for mineral separation by fluidisation were granted to ROBINSON (1879) and CHARD and DANE (1884). A similar Polish patent was granted to MISCHKE et al (1938).

The first patent that mentions fluidisation in connection with a gas was issued to PHILLIPS and BULTEEL (1910); they used a finely divided catalyst which was swept by a gas in the dilute fluidised phase into a reaction chamber. The products then carried the catalyst into a recovery vessel from which it was returned to the feed point by way of a standpipe.

Probably the first fluidised unit to operate commercially was the air-blown gasifier patented by WINKLER (1923) which had the disadvantage of producing large amounts of fines that were blown out of the unit. To improve utilisation of the fines, a modification allowed air or an oxygen/steam mixture to be blown above the fluidised bed, raising the temperature at this point and the carbon gasification efficiency from 80 to 85%.

The first large scale fluidisation application in the United States of America dates from about 1940 and refers to catalytic cracking of oil vapours - (MURPHREE et al 1943). A large scale plant was built at the Massachusetts Institute of Technology to test the principles of FBC in the 1950's and was described by OTHMER (1956) in the publication 'Fluidisation'. From the late 1950's, the fluidised bed was also utilised as a technique for metallurgical heat treatment; machined metal parts were treated in a fluidised bed of sand heated by natural gas. From about the same time, fluidised bed ore roasters and incinerators were also being built.

During this period, a number of combustion processes incorporating the principle of fluidisation were being developed in Europe, the U.S. and Japan, although mainly as a means of burning low grade coal fines. The main advantage was that the high cost of grinding these fuels so that they could be used in pulverised fuel burners was avoided. These processes have been described in full by TEAGUE and WRIGHT (1966).

Small scale work on fluidised bed combustion has been carried out by the U.S. Bureau of Mines since 1965. The work has been done at Morgantown and Pittsburgh and has been designed to establish operational conditions and heat transfer coefficients for eventual firing with closed cycle gas turbines. The equipment at Pittsburgh has been described in a paper by ORNING and McCANN (1968); further work on control of SO_2 emissions is continuing. Commercial companies such as Standard Oil of Delaware (1952), Badische Anilin-u-Soda Fabrik (1957), Combustion Engineering (1955) (1957) (1958), Union Carbide (1963) (1964) (1967) and Lurgi (1962) have also tried to incorporate the high heat transfer coefficients obtainable in fluidised beds into chemical processes in various ways.

The U.S. Office of Coal Research together with the National Air Pollution Control Administration has sponsered the early work of Pope Evans and Robbins on small packaged steam raising plants using fluidised bed combustion. Their experiments included studies of the effects of different methods of coal feeding and the effects of varying feed particle size. They measured heat transfer coefficients, sulphur and nitrogen oxide emissions and they endeavoured to minimise the amount of carbon carryover. A report of this work has been presented by BISHOP (1966) (1968).

The best known FBC system which does not employ direct extraction of heat from the bed is the French 'Ignifluid' boiler, developed by Societé Anonyme Activit in 1950 and described by GODEL (1963). The atmosphere in the bed is a reducing one, i.e. insufficient air is passed through to burn the fuel completely but secondary air is introduced to complete the combustion above the bed. The hot gases are passed through a conventional water-tube boiler and carryover is recycled to the bed.

Similar processes to the Ignifluid have been developed in Romania (PANOIU and CAZACU 1962), Czechoslovakia (NOVOTNY 1963) and Belgium (FASSOTTE 1961) whilst in Japan, (OKANIWA and SUZUKI 1959) a system was developed which was a cross between pulverised fuel firing and dilute phase fluidised combustion. In the latter, the crushed coal was P.F. fired above a fluidised bed into which the majority of it fell. About 80% of the combustion air was blown through the bed, the rest being used as primary and secondary air in the P.F. burners. Temperatures of around 1200°C were recorded above the bed.

In the U.K., the British Coal Utilisation Research Association (B.C.U.R.A.) in 1964 constructed and operated two experimental FBC units solely to obtain design and operating data for a prototype FB industrial boiler; the units and their instrumentation are described by WRIGHT and KEATING (1966). As would be expected, the National Coal Board through its Coal Research Establishment (C.R.E.) has also been actively engaged in FBC research and it initially produced both a 150 mm cylindrical and a 300 mm square combustor. These were built to examine methods of fuel feeding, ash removal and fines return, and also to study the effects of factors such as base design and fluidising velocity. The work has been described in several reports from C.R.E. by BAILEY et al (1968), COOKE et al (1968)[1] and COOKE et al (1969)[2].

Before building the combustors, preliminary experiments were carried out using a range of cold models to determine data required for the design of future fluidised beds units. These experiments have been published by C.R.E. in a series of reports (SMALE 1968), (SMALL 1968), (JURY 1967), (HIGHLEY et al 1969).

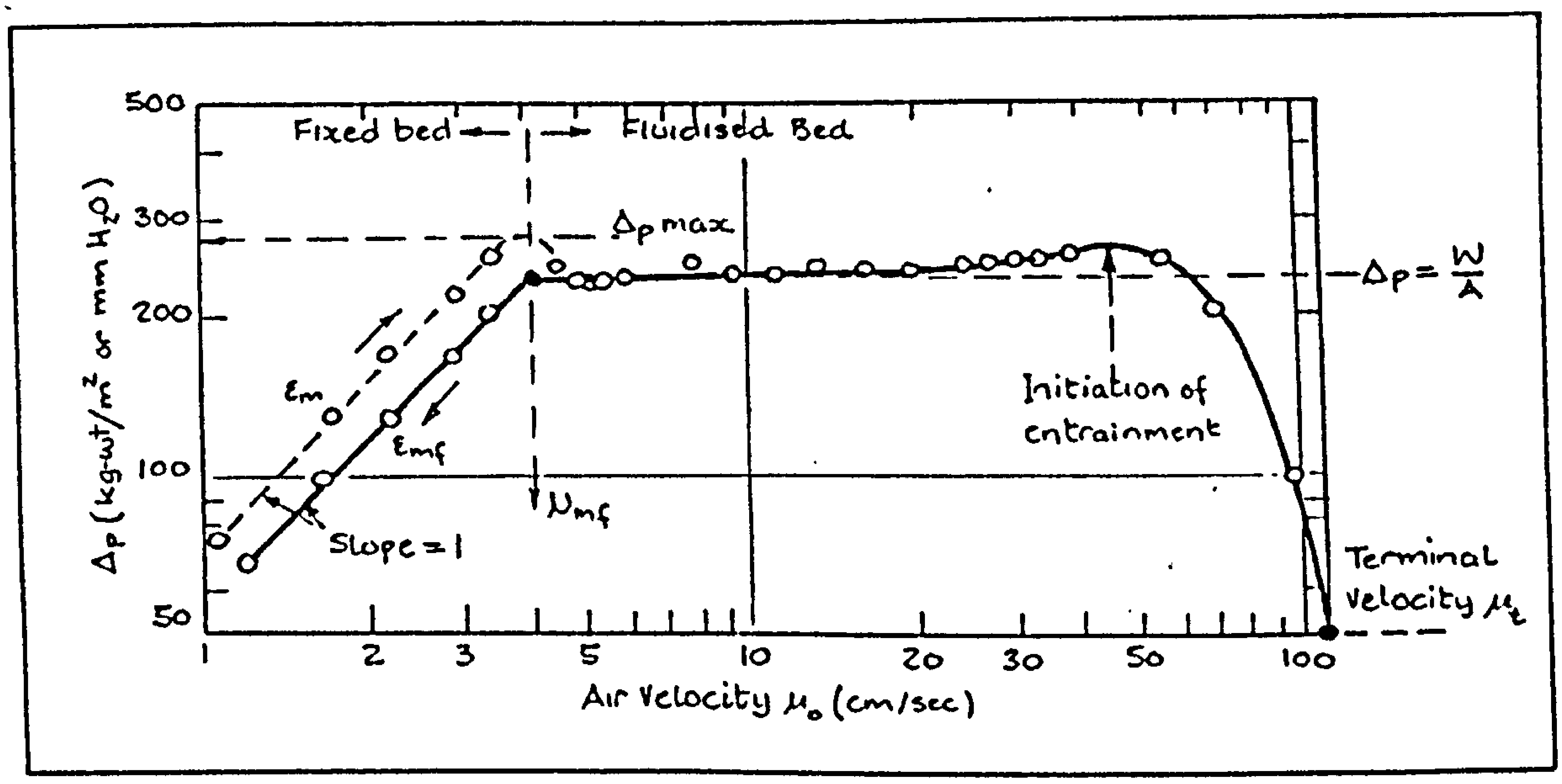
CHAPTER 3

FLUIDISED COMBUSTION TECHNOLOGY

When an upward moving gas is passed at a steadily increasing rate through a bed of particles - sand for example - a point is eventually reached when the pressure drop across the bed becomes equal to the weight per unit area of the bed. The bed is then on the point of 'fluidising' and the point of incipient fluidisation is reached (Fig 15). The gas velocity to achieve this state is called the minimum fluidising velocity (U_{mf}). Then, as the gas velocity increases, there is no corresponding increase in pressure drop across the bed. However, the bed expands to accomodate the additional flow which passes through it in the form of bubbles. When the gas velocity is approximately three times the minimum fluidising velocity, the bed takes on the appearance of a violently boiling liquid and at this stage, it consists of two phases - the bubble phase and the dense phase which has a density very close to that of the bed at the minimum fluidising velocity.

The boiling action creates a high degree of particle mixing in the bed and because the particles have a large surface area ($1m^3$ of 100μ diameter particles has a surface area greater that $30{,}000m^2$ [BOTTERILL 1983]), thermal equilibrium between gas and particles is quickly reached. This is the reason why fluidised bed furnaces are capable of responding quickly to changes in demand for heat. The sand bed is continuously either giving up or absorbing heat and, as would be expected, because of the surface area, small particles will give higher heat transfer coefficients than beds consisting of larger particles (BOTTERILL and BUTT 1968).

From Zabrodsky's correlation (ZABRODSKY et al 1976, den DULK & KORTLEVEN 1986) the maximum heat transfer is approximately 75% of the thermal conductivity of the gas, (BOTTERILL 1985) and will be greater the shorter the residence



Pressure drop versus gas velocity for a bed of uniformly - sized sand particles (measured by SHIRAI)
(Achnowledgements : KUNii and LEVENSPIEL, 'Fluidization Engineering'. John Wiley & Sons Inc. 1969)

FIG 15

time of the particles at the heat transfer surface (SKINNER 1969). To ensure the residence time is at a minimum, the amount of mixing within the bed must be maximised and it is therefore important that the bed is evenly fluidised over its whole area. To obtain rapid solids mixing and good rates of heat transfer, it is necessary to operate fluidised beds at velocities in excess of the minimum fluidising velocity. This causes a large fraction of the gas to pass through as bubbles and its is important to know how effective is the contact between the gas and the solid particles.

Work at the Atomic Energy Research Establishment at Harwell (LEWIS and PARTRIDGE 1967) has shown that the gas flow pattern associated with the bubble is dependent on the ratio of the bubble velocity to the minimum fluidisation velocity. The gas which forms the bubbles is continously changing, rising bubbles are continuously splitting up and coalescing and the average bubble size increases with height in the bed.

When fuel is fed into a fluidised bed in batches, for example by a rotary valve, the bubbles which form around the fuel are large and the rate of combustion of volatiles is limited by the rate of interphase mass transfer. On the other hand, where fuel is added evenly, small bubbles are formed and the rate of combustion is mainly limited by the rate of diffusion of fuel and oxygen to the interface between the fuel-rich and oxygen-rich regions of the bed.

Work carried out at Cambridge indicated that up to 80% of heat produced by combustion in the freeboard returns to the bed via solids in the splash zone (TURNBULL 1983). Therefore freeboard temperature rises are not as large as might be expected.

It was found that as bed temperature decreased between 1073 K and 1023 K, the amount of overbed burning increased dramatically because reactions within the bed were inhibited. For bed temperatures above 1123 K, the rate of combustion of volatiles appears to be controlled by the rate of fuel and oxygen supply (TURNBULL 1983)

Experimental data shows that:

1. The amount of volatile matter that burns in the bed increases with bed depth and excess air.
2. The likelihood of over bed burning increases with volatile matter content of fuel - and -
3. The amount of over bed burning can be reduced by controlling the flowrate of air for conveying the fuel.

There are three ways in which the rate of heat transfer can be influenced in a fluidised bed:

1. By changing the average size of particles comprising the bed.
2. By altering the average residence time of the particles at the heat transfer surface - and -
3. By varying the bed temperature.

As the average particle size decreases, the heat transfer rate increases (KHARCHENKO and MAKHORIN 1964) but work carried out at C.R.E. indicated that in practice, particle size should not be reduced below 60 mesh B.S.S. (0.250 mm) or increased above 10 mesh B.S.S. (1.70 mm) (WILLIAMS et al 1967).

The average residence time of particles depends upon several factors but mainly their concentration in the bed and the rate of agitation. Evenness of flow through the bed and the presence of obstructions within it such as cooling tubes, also have an effect but these are relatively minor.

The bed temperature controls the amount of heat transferred by direct radiation, which for a fluidised bed of particle size 3.2 mm - 0.50 mm contributes up to 30% of the total heat transfer. For a bed of larger particle size, the proportion contributed by radiation would be higher than 30% (WRIGHT and KEATING 1966).

For successful operation of a fluidised bed combustion system, the fluidising velocity of the gas through the bed depends on the one hand on the minimum velocity required to maintain satisfactory fluidisation and, on the other on the maximum velocity which can be used before elutriation from the bed becomes excessive.

Plants must work within these limits; at the lower end a fine bed material can be fluidised at approximately 0.5 m/s and this is satisfactory for finely crushed fuels, but high output units have to be large and hence expensive. To cut down on plant size for a given output, coarser bed materials and larger particle sized fuels can be used at gas velocities of up to 3 m/s without unacceptable levels of elutriation taking place.

In most cases, it is usual to install grit arrester equipment to collect elutriated material; this will either be returned to the bed when low ash fuels are being burned, or it may be sieved and then only bed sand and unburned carbon particles will be returned to the bed.

At the C.R.E. it was found with their small combustor that the loss of carbon elutriated in fuel particles was as much as 0.6% of the input but that this was reduced to 0.2% by recycling sieved material back to the bed (BAILEY et al 1968).

In an FBC unit, a significant part of the total running cost is incurred by the need to use a high pressure fan for the fluidising air. It is therefore desirable that the pressure drop across the distributor should be as low as possible, whilst ensuring an even distribution of air across the bed.

Various methods and materials have been used to ensure efficient distribution - porous tiles, flat metal plates with drilled holes, flat metal plates with capped stand pipes and sparge pipes fitted horizontally across the bed (Fig 15).

Whilst it is generally recognised that the rate of vertical mixing in a fluidised bed is very high, the rate of horizontal mixing is low (HIGHLEY ROGERS & DRYBURGH 1969). This is because as bubbles rise, pushing particles upwards and sideways out of their way, movement of the particles takes place as they follow the bubble up through the bed. The wake of particles following a bubble is estimated as one quarter of the bubble volume and this is continuously being lost and replaced during the upward movement. (HIGHLEY et al 1969). The only real horizontal movement takes place when the bubble reaches the surface of the bed but this contributes little to the overall mixing of the particles.

As stated in Chapter 2, most of the work carried out on fluidised beds has been with coal of various qualities, and to obtain efficient combustion, the distillation products as well as the residual carbon must be completely burnt.

Volatiles, which compose up to 30% of a good quality coal, are released within two or three seconds of entering the bed, and fine particles will burn above the bed. It would therefore seem sensible to inject the fuel as near to the bottom of the bed as possible to allow for a greater burning time. However, because of the practical difficulties of siting and cooling under bed feeders, commercial operators are prepared to lose some efficiency by the use of above bed feeders.

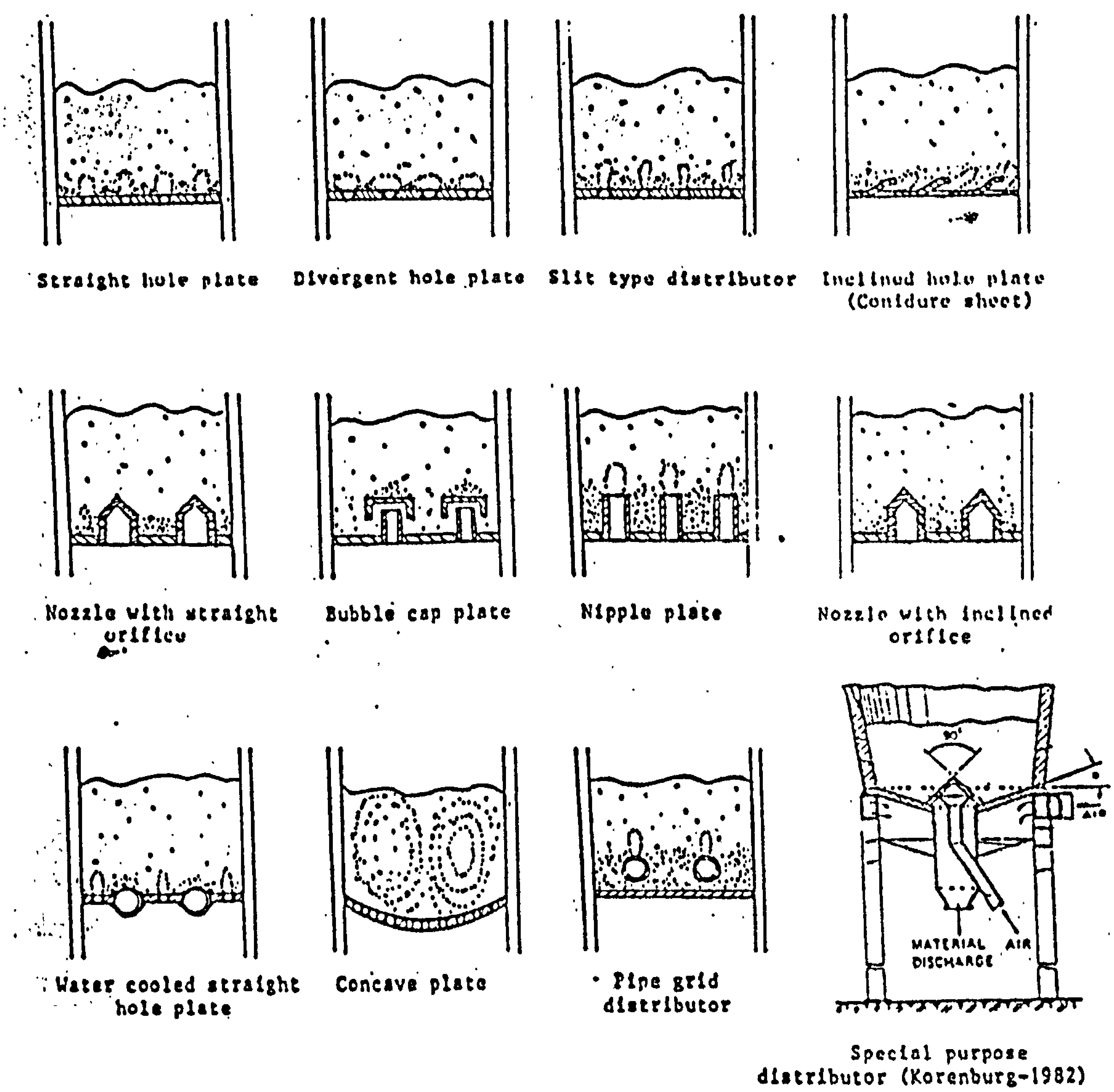


Fig. 16 Different types of distributors for Fluidized Bed Boilers

Whilst work has been carried out on solid fuels injected into the bed in many countries, very little has been reported on light fractions, however at the Department of Fuel and Energy at Leeds University, experiments were carried out with a 0.3m square FBC pneumatically feeding peat into the base of the bed (HAMPARTSOUMIAN & GIBBS 1983). This work formed part of an investigation into the combustion efficiency and suitability of a range of solid fuels under various operating conditions. The fuels chosen - peat, anthracite, coke, low and high ash coals - were selected to encompass a wide range of properties, e.g. heating value, ractivity, density, volatile content and ash content.

The experiments showed that the best combustion efficiencies were achieved with coal and peat, and it was found that the response to changes in fuel feed rate when the unit was fired on peat was almost instantaneous. The CO emissions from the various fuels were dependent on both the level of volatiles in each fuel and feed size. NO emission, however, was dependent on the bed temperature except when peat was the fuel; emissions increased with temperature at a rate approximately equivalent to 1.5 ppm NO/°K.

CHAPTER 4

FLUIDISED COMBUSTION - ITS POTENTIAL ADVANTAGES

Fluidised bed combustion will become a commonly used technique with units adapted to burn coal of all grades, lignite, peat and waste derived fuels, and there are significant advantages when using a fluidised bed boiler:

1. Because of the improved mixing between fuel and air resulting in better combustion and heat transfer; in a unit of standard output FBC would require less steam-tubing and therefore capital costs might be reduced.
2. The increase in efficiency allows the FBC to operate at a lower combustion temperature for a given steam condition with the result that less volatilisation of alkali compounds occur and the corrosion and fouling of tubes is greatly reduced.
3. The operating temperature of the FBC is below the fuel N_2 NOx formation point and by adding limestone or dolomite to it, the SO_2 released during combustion can be absorbed, thereby reducing pollution of the atmosphere.
4. Coal preparation and coal crushing costs would be reduced and pulverising to a fine powder eliminiated.

It is likely that the fluidised bed combustors will be in three forms:

1. Atmospheric (A.F.B.C.)
2. Pressurised (P.F.B.C.)
3. Rotating (R.F.B.C.)

Atmospheric fluidised bed combustion is seen as a likely energy source for industrial and agricultural furnaces and boilers in the range 2MW to 40MW. The more complex pressurised fluidised combustor is likely to be used for power generation up to about 1000MW. All the advantages of A.F.B.C. apply to pressurised beds but in addition, the compactness of the latter is enhanced because the area needed per unit of power output is reduced. Rotating fluidised beds are in their infancy but would appear to have certain advantages over both A.F.B.C. and P.F.B.C.

An important limitation of both A.F.B.C.'s and P.F.B.C.'s in determining the combustion intensity and efficiency, is the fluidising velocity at which the combustion losses by elutriation become excessively high. A further problem is that they only have a limited turndown ratio of some 3 : 1, compared with that of a gas turbine turndown of 10 : 1. One way in which this limitation can be partially overcome is to fit the fluidised bed with a rotating cylindrical distributor; the terminal velocity limit can then be increased because of the centrifugal forces, and as the turndown ratio depends upon the minimum fluidising velocity, this can be varied by simply changing the speed of the devices rotation. (SUBZWARI et al 1980).

On any fluidised combustor it is impossible to completely overcome the problem of elutrition of unburned carbon, which results in a loss of combustion efficiency. Despite this, an FBC will normally operate at efficiencies greater than 85%, the actual value depending on initial particle size, fluidising velocity, the amount of excess air and the amount of unburned carbon particles recycled from grit arrestors.

CHAPTER 5

GASIFICATION OF FUELS

In addition to the complete combustion process of a fuel being carried out in an FBC, fluidised bed units can be designed to generate combustible gases. Such units are now being used where a large quantity of a clean burnable gas is required for a particular process and where neither town gas nor natural gas is available. Practically all gasification of fuels has been carried out with coal as the feedstock.

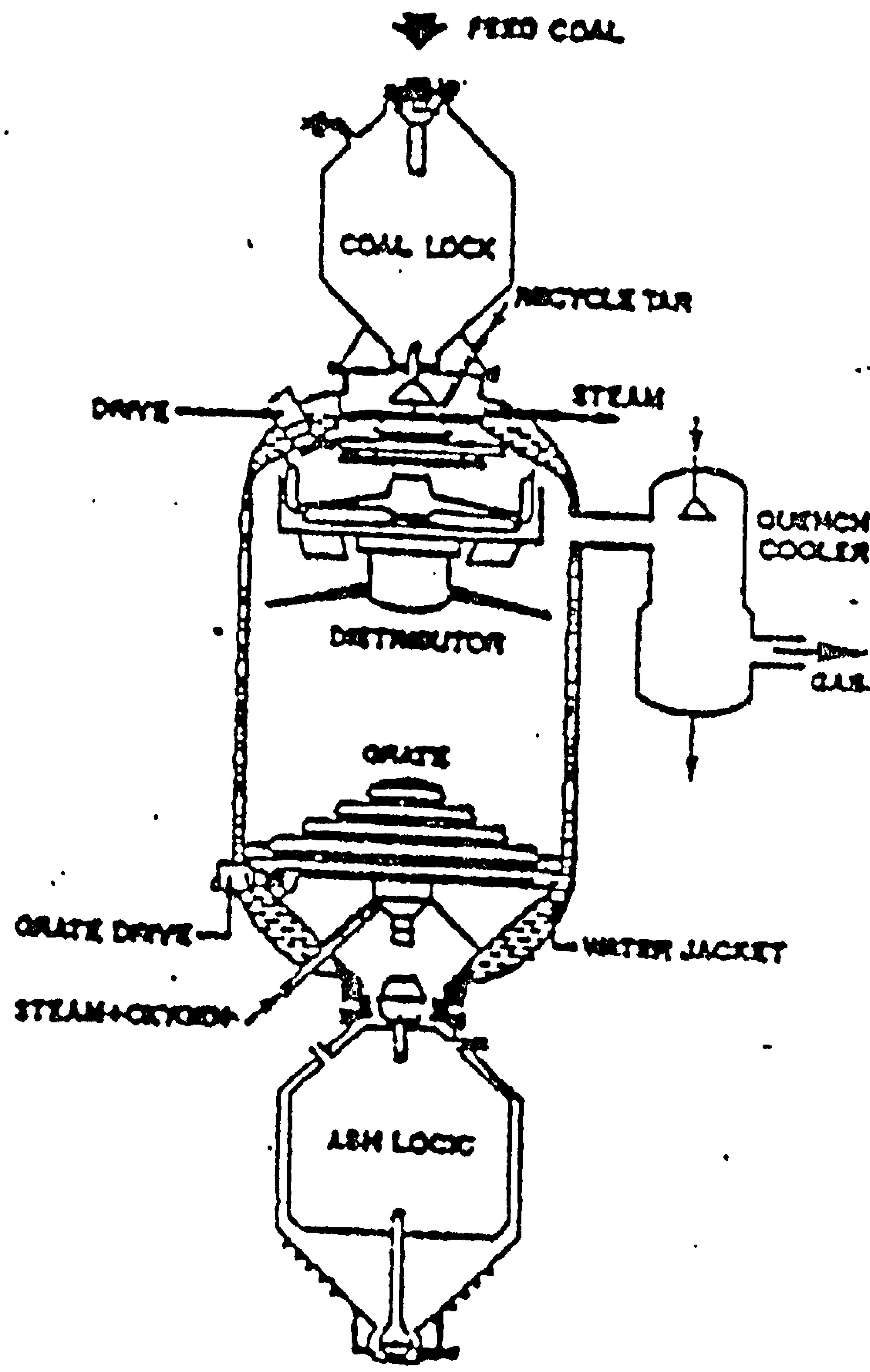
The original town gas system was based on the carbonisation of coal and the gas produced had a calorific value of about 20 MJ/m^3. To increase the yield, processes were developed in which air and steam were blown through static beds of red hot coal or coke. Used industrially, the gas produced had a calorific value of about 6MJ/m^3 and was known as a low calorific value (L.C.V.) gas.

In an attempt to upgrade the gas, oxygen was used instead of air, thereby avoiding the diluent nitrogen. This concept was commercially adopted by companies like LURGI of Germany in its high pressure (20 to 30 atmospheres) single stage fixed bed gasifier. (Fig 16)

The product gas was discharged at about 600°C and contained tars, ores and fine coal particles which were all recycled to the gasifier.

At a later stage, two-stage fixed bed gasifiers were developed in which part of the hot product gas was tapped off directly from the gasification zone, thus reducing the temperature of the distillation zone from which the product gas leaves at 150°C, taking only the fluid tar component with it.

This principle is also used in the Wellman-Incandescent process, which like all other gasifiers of the type suffers from the severe



Lurgi fixed bed gasifier

FIGURE 17

limitation that it can only work satisfactorily when fed with carefully sized coals with a minimum fines content.

An alternative process developed by WINKLER in Germany and using the atmospheric pressure fluidised bed principle (Fig. 18) used mainly fine brown coal ($<$ 6mm) and again used oxygen and steam as the gasifying agents. An important feature of this design is the large freeboard zone which is necessary to crack tars and contain elutriated carbon.

Whilst this process is being upgraded by several other companies to operate at elevated pressures, most of the companies developing fluidised bed processes are aiming to improve the performance of their FBC units to cope with bituminous coals, waste derived fuels (W.D.F.) and coking coals. This research, mainly the work of the National Coal Board, has led to the creation of rapid in bed circulation by a spouted gas inlet and central draught tube. (Fig.19).

Fuel gasification signifies the reaction of the fuel with air, oxygen, steam, carbon dioxide or a mixture of these materials to form a gaseous product whose principle components are formed by a combination of the following overall reactions:

1. Oxidation of carbon

   $C + O_2 \text{ ---> } CO_2$

2. Bernoullis reaction

   $CO_2 + C \text{ ---> } 2CO$

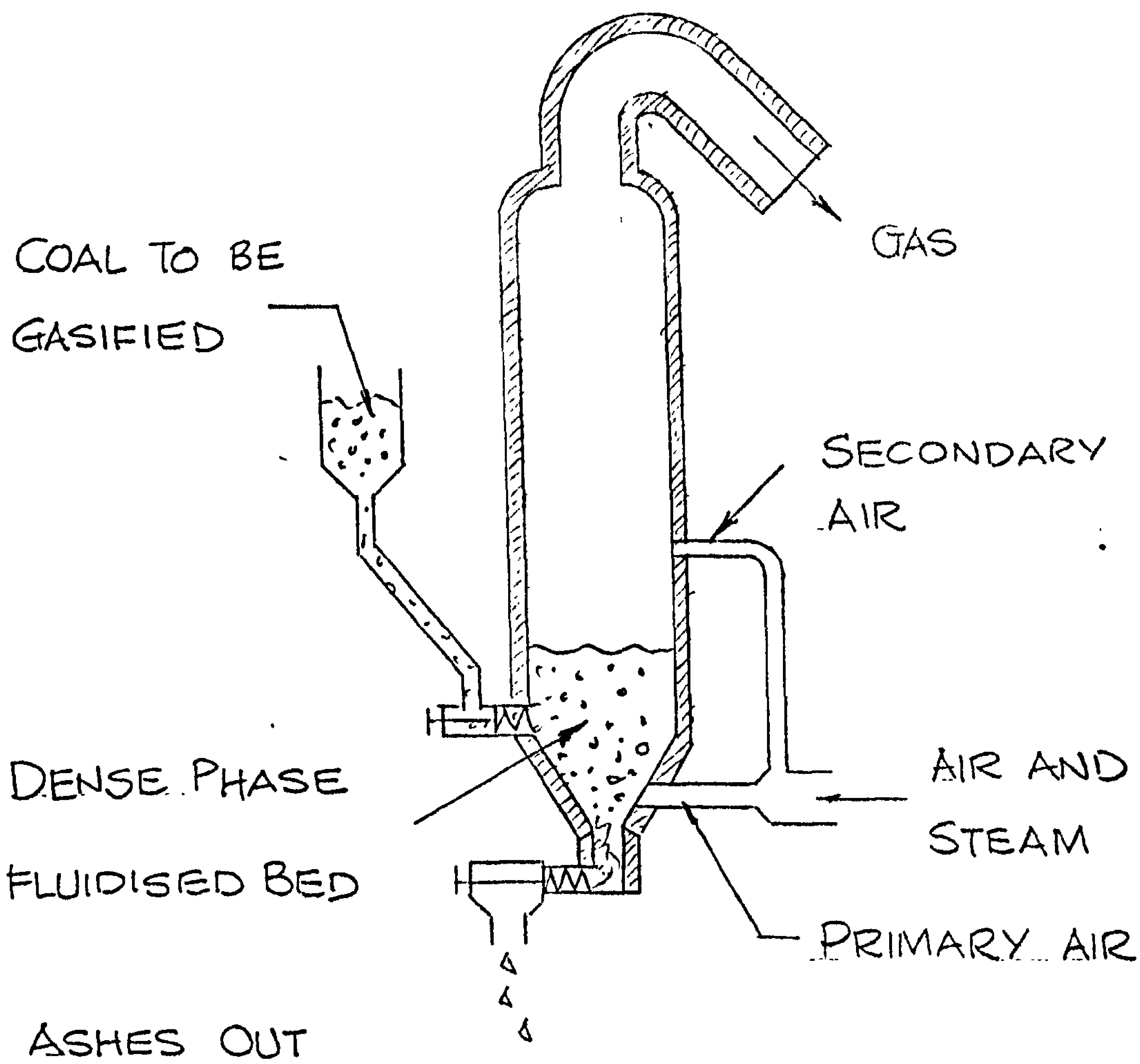
3. Water gas reaction

   $C + H_2O \text{ ---> } CO + H_2$

4. Water gas shift reaction

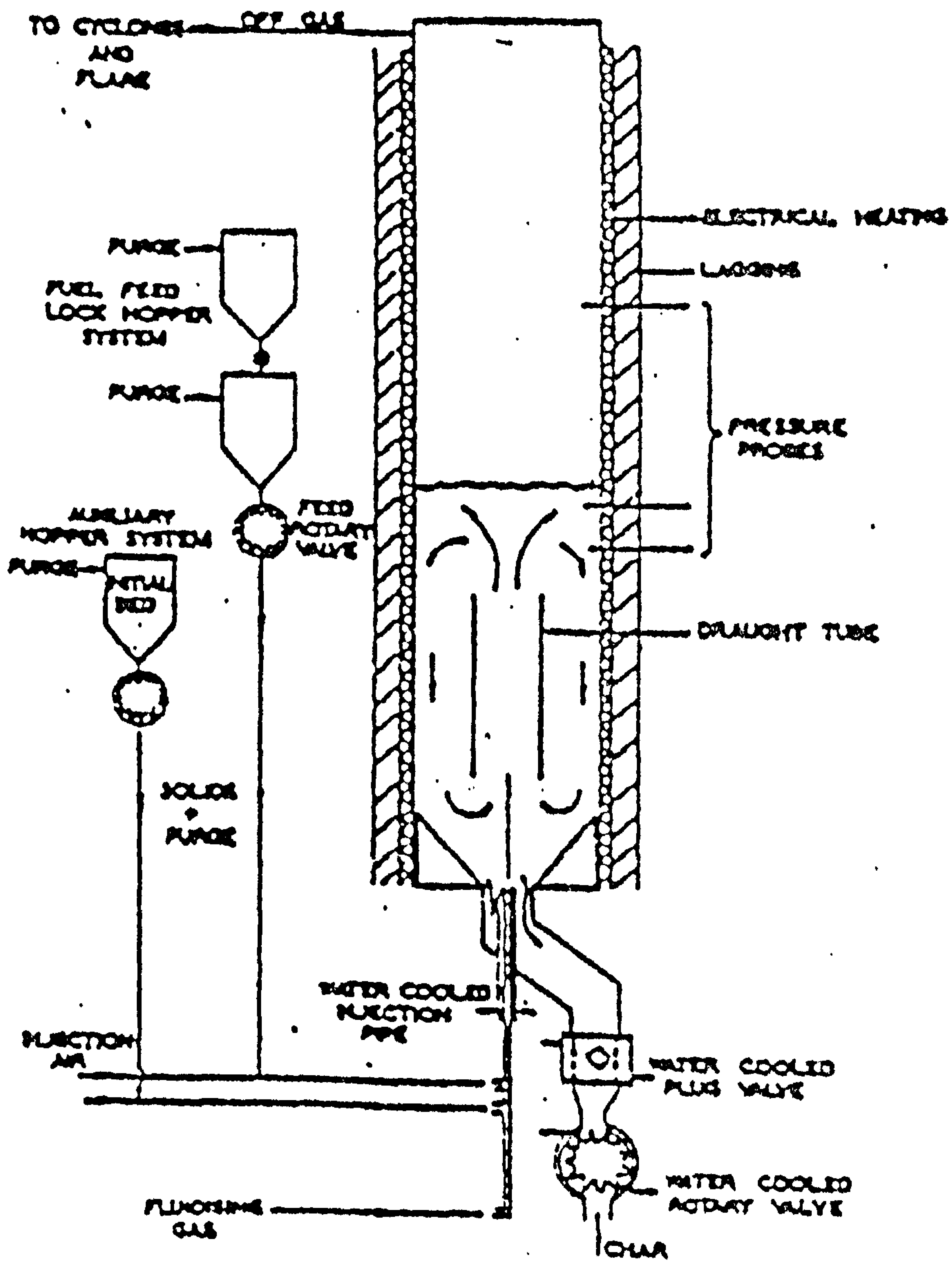
   $CO + H_2O \text{ ---> } CO_2 + H_2$

Decomposition of the volatile matter also yields CO, H_2, CH_4 and higher hydrocarbons.



WINKLER GASIFIER

FIG 18



NCB pilot scale gasifier

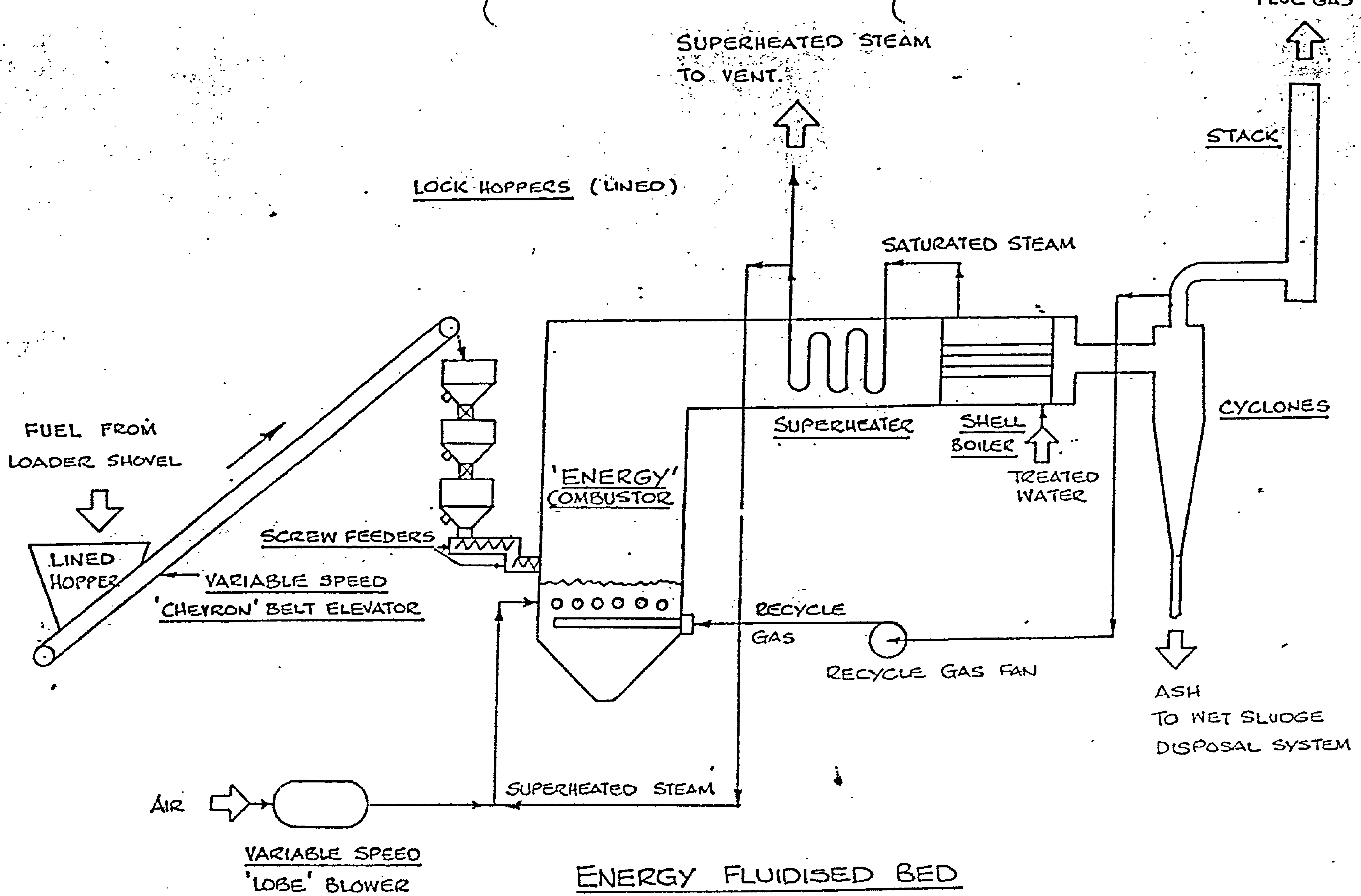
FIGURE 19

The gasification reactions of 2 and 3 are highly endothermic and require heat to be supplied at relatively high temperature - this is normally provided by the combustion reaction 1. The main problem is that combustion is so rapid that it proceeds to completion while gasification reactions, which are about 10^3 times slower, never reach equilibrium at exit conditions (GREEN and CRICHTON 1981). The Product gas therefore contains undissociated carbon dioxide and steam.

With air-fed gasification processes, dilution of the gas by nitrogen limits the calorific value to between 4 - 7 MJ/m³, but when air is replaced by oxygen, the resulting medium calorific value (M.C.V.) gas has a heating value of around 11 MJ/m³.

The Energy Equipment Company Limited (E.E.C.L.) through their own patented process, have evolved a different method of L.C.V. and M.C.V. producer gas production. Adapting their two phase combustion, hot gas generator by replacing the recycled gas (previously used for restricting the oxygen content of the bed) with a mixture of air, highly superheated steam and a proportion of the made gas, a combustible gas was produced. This L.C.V. gas has a calorific value of only 5.0 MJ/m³ with virtually no vapourised tar content; it is hot (about 950°C) but can be cooled to any temperature suitable for its end use.

In the layout of the plant (Fig. 20) gas passes to a waste heat boiler which incorporates a steam superheater, providing steam capacity and a gas cooling effect down to 450°C. The steam, which is produced at a pressure of 17.2 bar is superheated to more than 400°C and is used for bed attemperation. Because the made gas temperature is produced at only 450°C, expensive high temperature materials are not required for duct fabrication, and there is a lower requirement for insulation.



ENERGY FLUIDISED BED
L.C.V. GASIFICATION PROCESS

FIGURE 20.

Where the process being served by the producer gas unit requires a high standard of particulate removal from the gas, for example in a brick kiln, this can be achieved with a suitable grit arrestor - one with a separation efficiency of 80 - 85%.

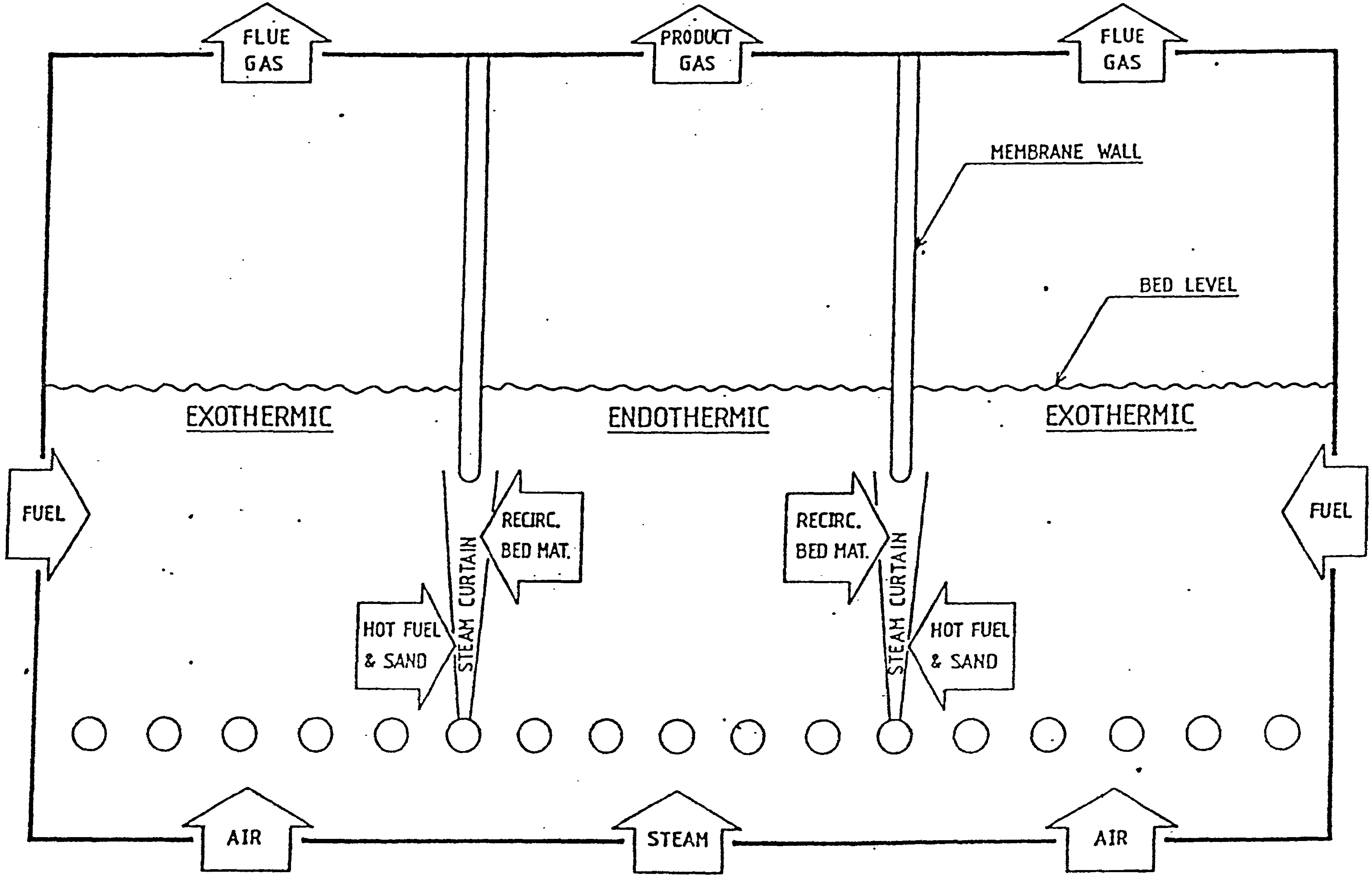
The Energy Equipment M.C.V. units generate a nitrogen-free synthesis gas from solid hydrocarbons using air rather than oxygen, mainly because of the cost of running an oxygen fed unit. There are other gasifiers which operate without oxygen but they use some mechanical form of pumping system to circulate the hot bed material between the heat producing part of the plant and the gasifier, which absorbs heat.

The Energy Equipment system has a common fluidised bed dynamic base for both sections and heat is carried by the transverse progression of the fluidised particles (transmigration) through the bed (Fig. 21).

A full description of the process, together with diagrams and photographs, is included in Appendix 3.

Energy from biomass has always been realised as an important addition to exisiting energy sources by the European Economic Community, and their Research and Development programme has given financial assistance to a number of different types of experimental plant, many of them being designed for biomass gasification (CHARTIER & PALZ 1981) (PALZ & PIRRWITZ 1983).

One of these plants, designed by Foster Wheeler Power Products Limited of London, is a 40 kg/hr pressurised biomass gasification test facility with high temperature and pressure operation (up to 1200°C and 30 bar). A partial oxidation reactor has also been developed by Foster Wheeler and this can be used commercially to retrofit to any gasifier system where the aim is to either modify the output or enhance operational flexibility.



SIMPLIFIED SCHEMATIC OF 'ENERGY' DIVIDED BED M.C.V. GASIFIER – ELEVATION

FIGURE 21

At Creusot-Loire Enterprises (C.L.E.) in France, a simple fixed bed oxygen gasifier for methanol synthesis, using biomass as a feedstock, has been built (Fig 21). This consumes biomass at the rate of 5 tonnes/hr at atmospheric pressure and has a recycling system which allows heavy hydrocarbon compounds to be burnt, thus avoiding contamination of the exhaust gas with these undesirable products. The same company has also constructed a pressurised oxygen blown fluidised bed gasifier, designed for use with wood as the feedstock. The initial results have shown that the composition of the syngas was affected mainly by the moisture content of the wood, and that to obtain a gas of significant heating value, it was necessary to use a pre-dried material with a maximum moisture content of 20%. With a bed temperature of 850°C, 30% M.C. wood gave a gas heating value of 6490 kJ / m^3 whilst 10% M.C. wood at the same temperature produced gas of 9422 kJ/ m^3.

After prolonged testing, a secondary reactor for partial oxidation of the gas was added to this plant. The temperature in the first reactor was able to be increased to 700 - 800°C by the addition of oxygen - this is below the ash sintering limit. Then a higher temperature (1300 - 1400°C) was possible in the secondary reactor which it is claimed ensures that the level of unburned carbon is low and the levels of ores and tars negligible.

At Wellman Mechanical Engineering Limited, an oxygen donor gasifier has been built with continuous recycling of solids between two fluidising beds, one being the gasifier compartment and the other the oxidiser compartment. Oxygen and heat for the endothermic gasification reaction are donated in the gasifier compartment by the bed material which is comprised of calcium oxide and calcium sulphate. This is fluidised by recycled product gas, whilst in the oxidiser compartment, the bed material is re-oxidised by fluidisation with air.

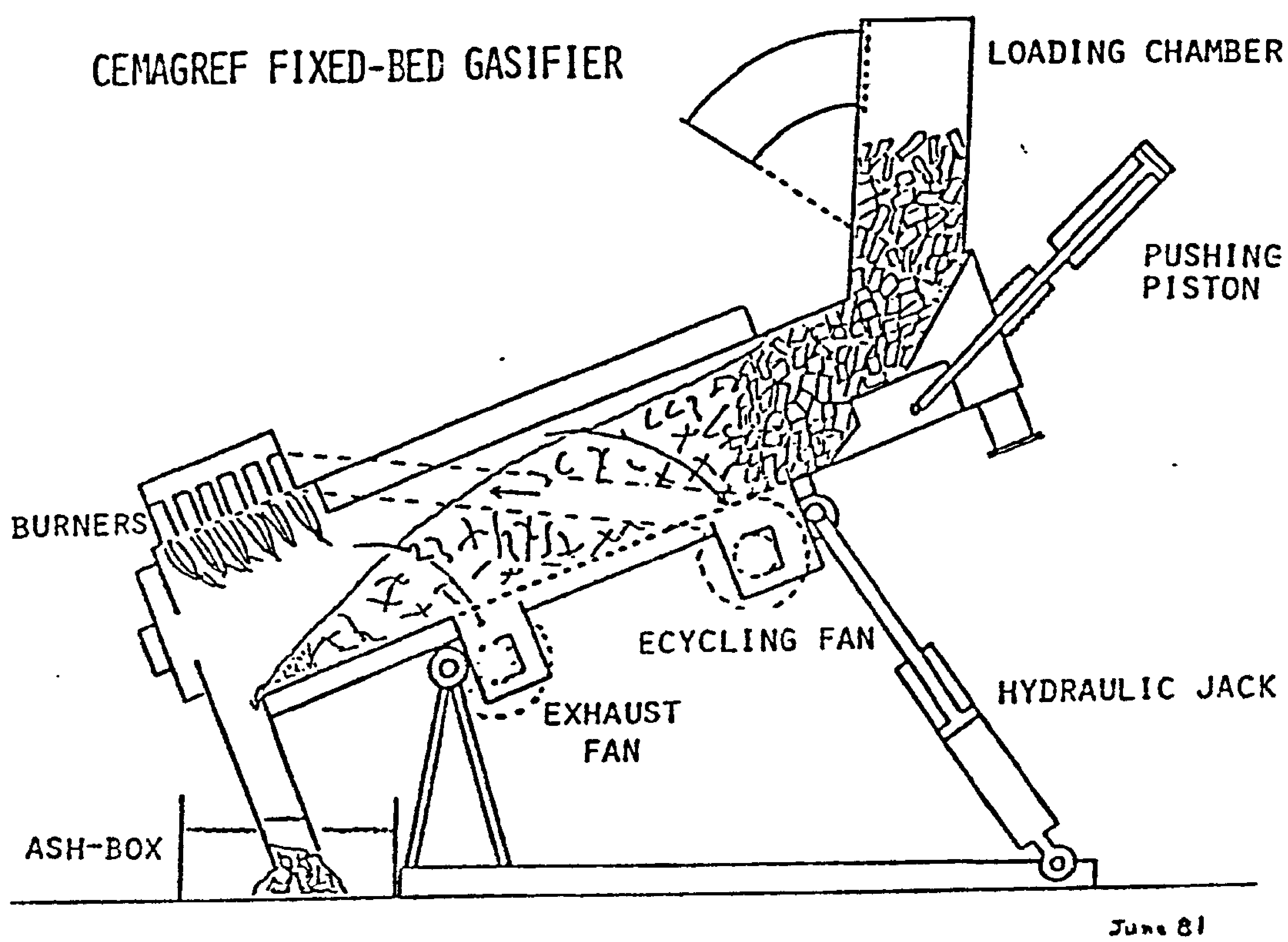

CEMAGREF FIXED-BED GASIFIER
LOADING CHAMBER
PUSHING
PISTON
BURNERS
ECYCLING FAN
HYDRAULIC JACK
EXHAUST
FAN
ASH-BOX
June 81

FIGURE 22

The rate of recirculation of bed material between the two compartments is critical, as sufficient material must pass into the gasification compartment to supply both the required amount of oxygen and to maintain the heat balance between the two compartments.

An interesting feature of this unit is the use of a metallic container as the gasifier compartment; this is mounted centrally to allow for thermal expansion and can be easily removed for inspection and repair (Fig. 23)

In Italy, Agip Nucleare of Milan are gasifying biomass for the production of synthesis gas with the idea of producing synthetic fuels in a later process. The concept of the plant is based on the gasification of biomass in the absence of air, the necessary thermal energy being produced by the combustion of part of the made gas in a separate combustion chamber (Fig. 24)

To transfer the generated heat into the gasifier the biomass in the gasifier and the inert material in the heating area are fluidised; this is achieved by partially recycling the made gas as a flushing gas while the remainder, after scrubbing, is used as a syngas.

As previously reported, in the U.S. there is considerable interest in the gasification of corn cobs and wood waste to produce a fuel gas for drying agricultural products. There has been no interest in the gasification of straw in the existing gasifiers because of the high cost of packaging it into a form which would make it acceptable for burning and handling. However, the disposal of various agricultural, industrial and municipal solid wastes in an environmentally acceptable manner are such that the Energy Resources Company Incorporated of Cambridge, Massachusetts has developed a fluidised bed gasification system for them. This converts these waste materials into low C. V. gas, pyrolytic oil and a char/ash mixture. (CHOROSZY et al 1981)

WELLMAN MECHANICAL ENGINEERING.

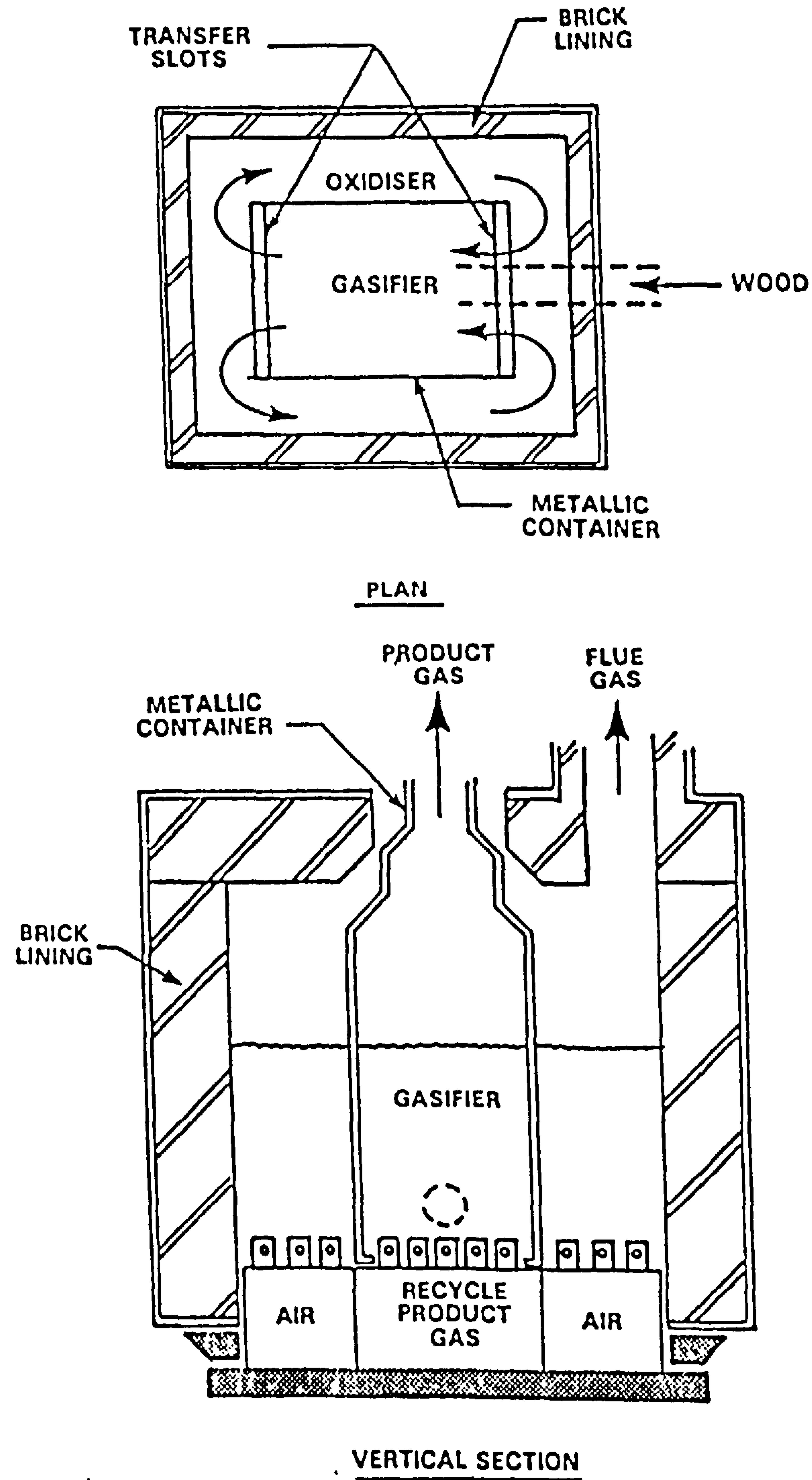


FIGURE 23: DESIGN FEATURES OF THE "OXYGEN DONOR GASIFIER"

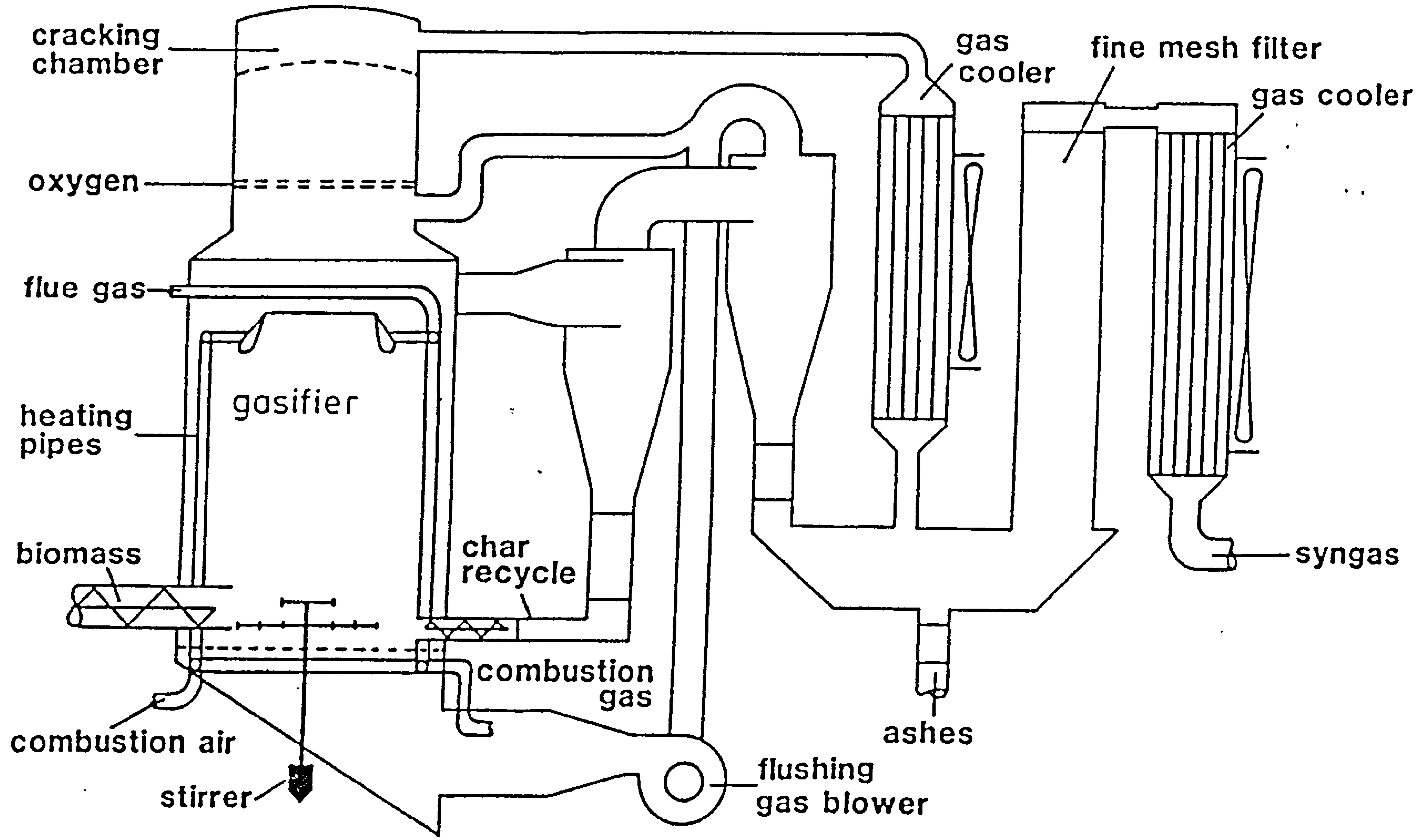


Fig. 24 AGIP NUCLEARE

It is claimed that up to 80% of the energy value of most materials is recovered in the form of these fuels.

The relative quantities of gas, oil and char produced are depend nt on the temperature of the reaction whilst the quality of the gas produced is a function of the fluidising medium. It has been found in practice that when air is used as the oxidant, the product gas is in the range 2.0 - 6.7 kJ/m³, but when oxygen or steam is used, a medium calorific gas is produced in the range 6.7 - 10.7 kJ/m³.

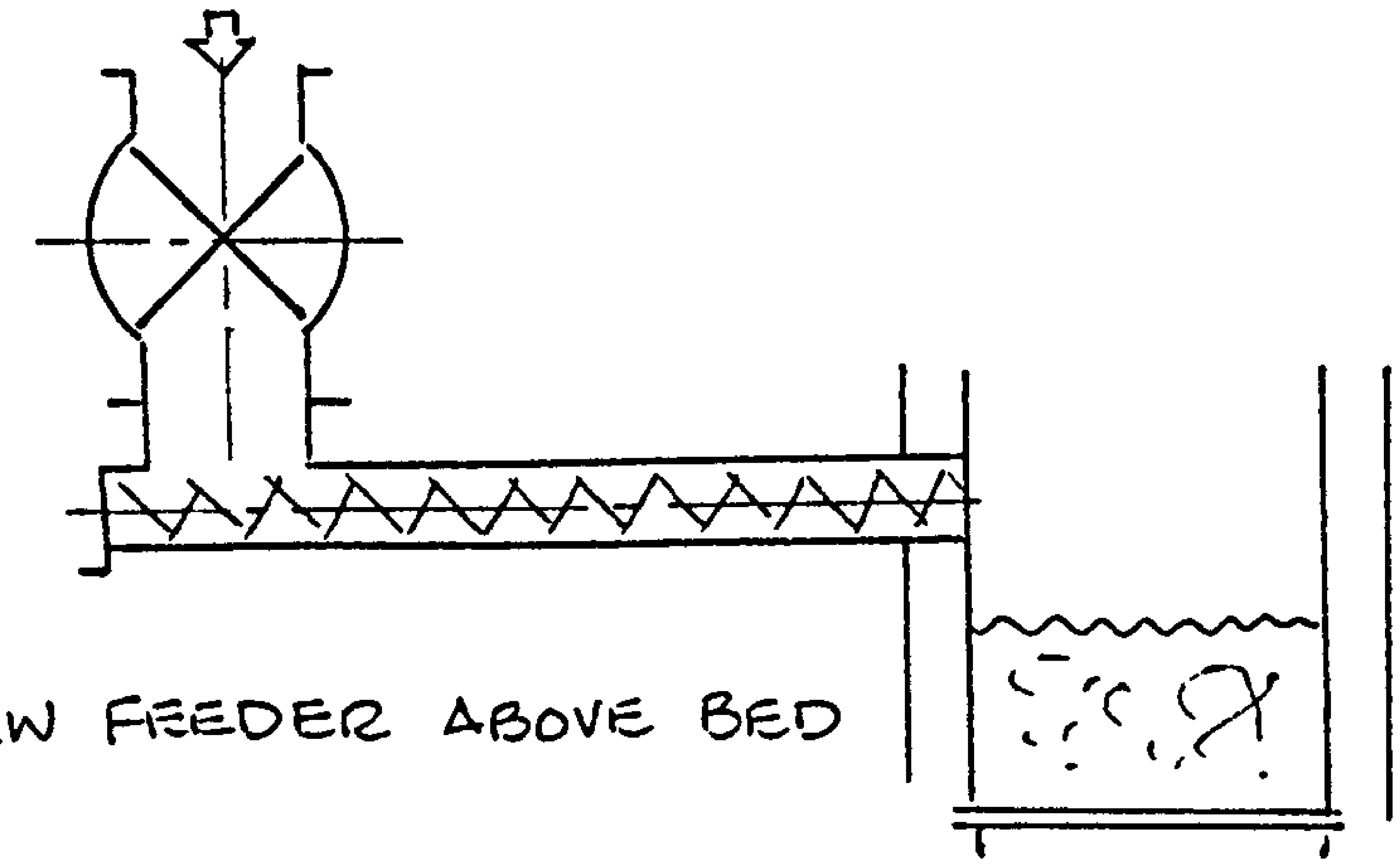
A feature of the system is that there are three different types of feeder (Fig.25) which can be used according to the physical characteristics of the feedstock. The first is the screw feeder of which there are two, one positioned above the bed in the splash zone and the other in the fluidised bed above the distributor plate. A ram feeder is the second type; it is sited above the bed and is the method favoured by the company because the plug of fuel produced acts as a seal which prevents the escape of gas through the feed mechanism. Unlike the screw feeder system the ram feeder does not require a high temperature rotary seal.

The third type of feeder is a pneumatic transport system that feeds directly into the bottom of the bed. It is claimed that this feeder gives the best fuel distribution in the zone near the distributor plate.

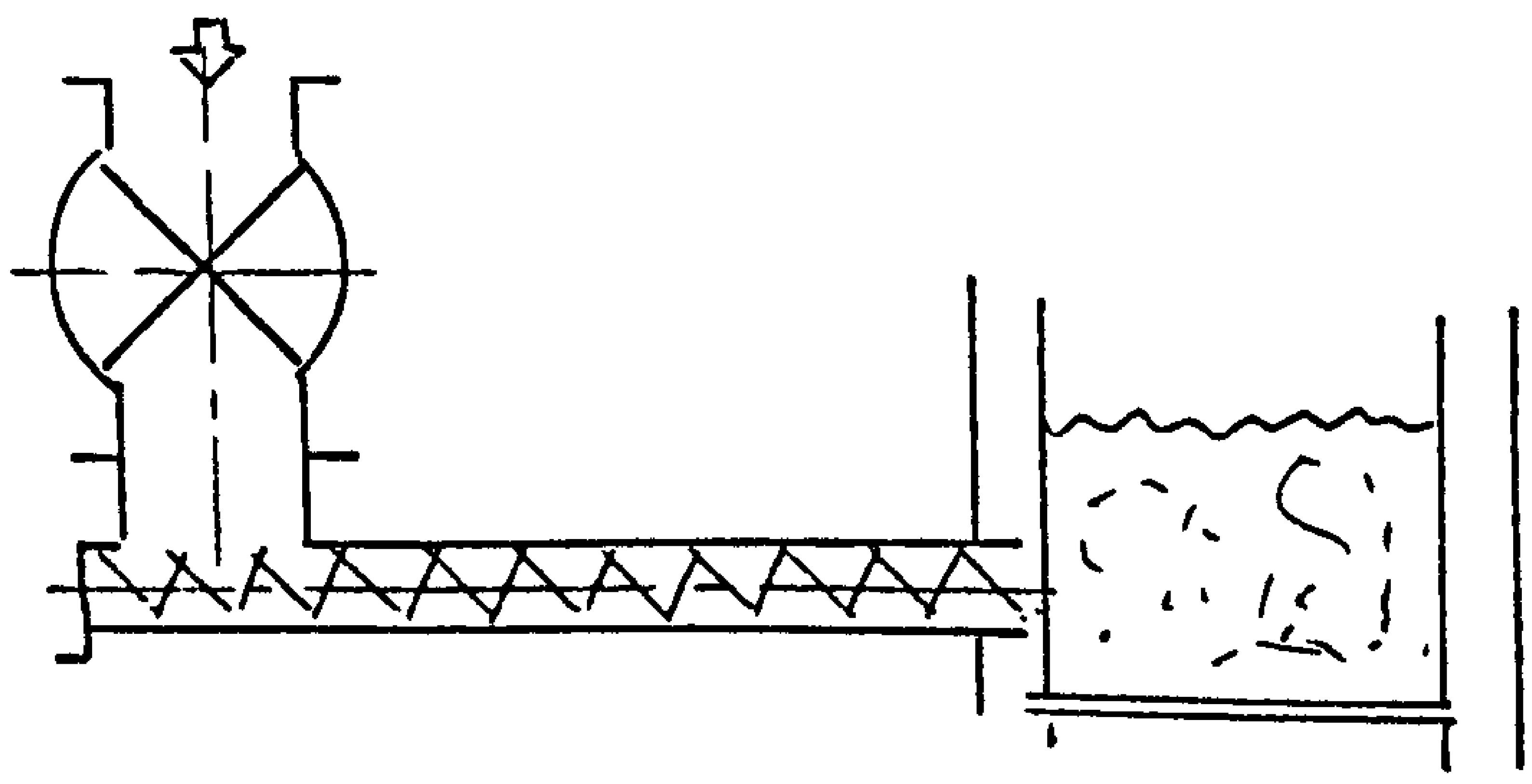
The fluidised bed in this unit is designed to operate between 370 - 1100°C with fluidising velocities ranging from 0.2 - 3.7 m/s and when operating in the combustion mode, it has a unique system of heat transfer from the bed. In order to accomodate the various types and positions of the feeders and also the variability of heat due to the different fuels involved, the arrangement of incoloy tubes used as the heat transfer surfaces can be varied vertically

FIG. 24

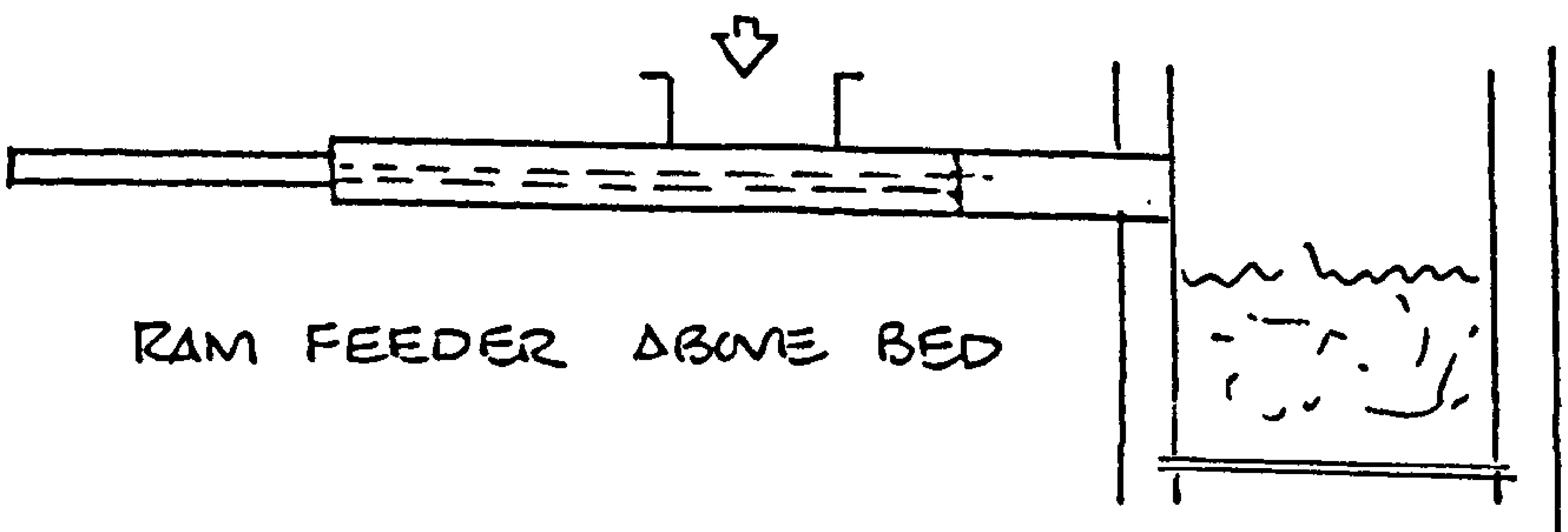
TYPES OF FEEDER



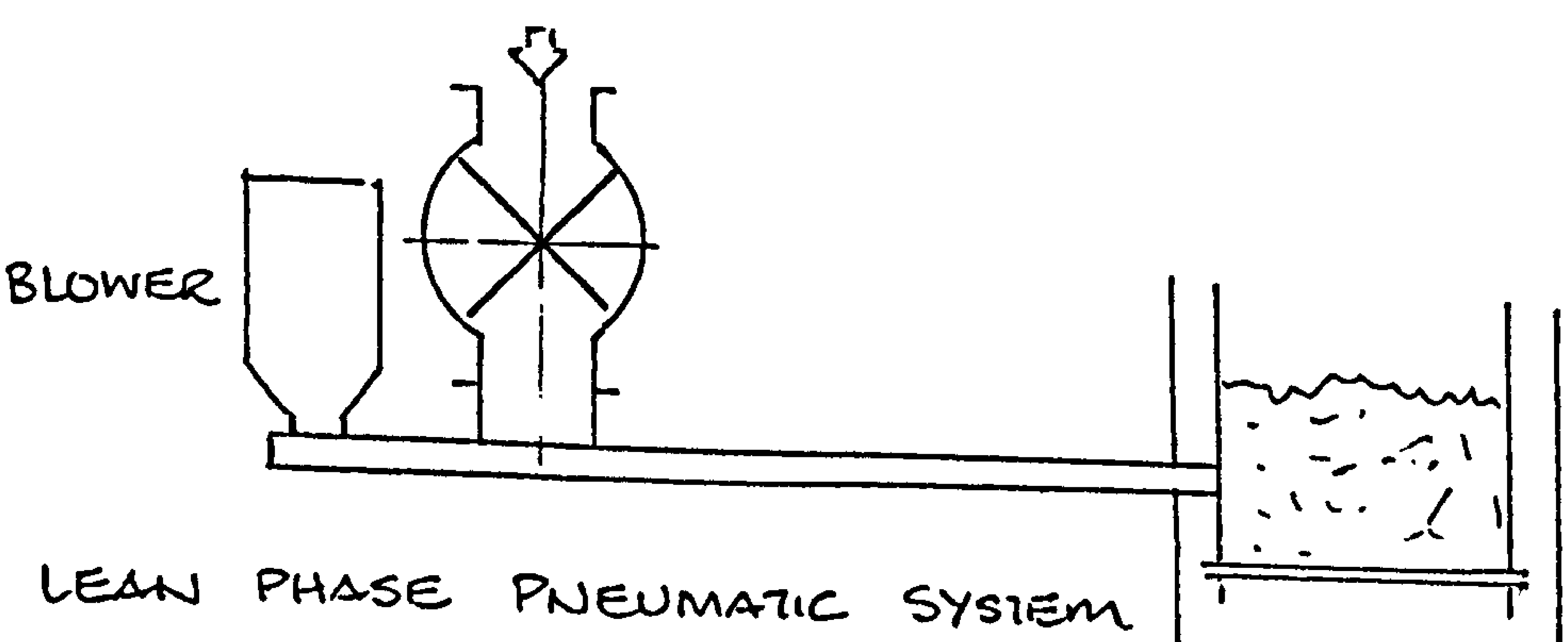
SCREW FEEDER ABOVE BED



SCREW FEEDER INTO BED



RAM FEEDER ABOVE BED



LEAN PHASE PNEUMATIC SYSTEM INTO BED.

to give any required distance between 0.38 and 1.3 m above the distributor plate. They can also be moved horizontally.

Within the total system, a shredder has been included, to reduce the fuels to an acceptable size and a dryer has also been added to ensure that the fuel is introduced into the bed at a suitable moisture content.

When the plant was first announced, it was stated that both rice straw and wheat straw were among the many wastes to be converted but no confirmation that this has been carried out can be found in the literature.

At the Esso Research Centre at Abingdon (MOSS 1983) an oxygen donor gasification pilot plant was built to gasify residual fuel oil. The object was to demonstrate the feasibility of gasifying without introducing oxygen, air or large quantities of steam. A limited number of tests were carried out successfully on liquid and solid fuels but the design of the gasifier was not suitable for continuous working.

CHAPTER 6

ALTERNATIVE USES FOR STRAW

The production of straw briquettes aimed at the domestic market in the U.K. is now taking place as nine presses have been installed to date for that purpose. (Straw Disposal and Utilisation - M.A.F.F. 1984). The output of each unit is relatively low (0.2 t/h) and the economics would seem to indicate that only large production units could compete against other fuels. Another problem is that briquettes do not burn satisfactorily on open fires so that their use is confined to enclosed stoves, where their bulk limits their value. The projected market is not expected to exceed 100,000 tonnes (0.05 Mtce) by the year 2000. So what are the alternative uses?

6.1 Straw Pulp in Papermaking

Modern chemical pulping of straw is carried out in Europe where, when fully bleached, it is added to other pulp in the manufacture of high quality paper. In an unbleached condition it is added to cardboard and corrugated paper. During the 1939 - 45 war, straw pulp was used in place of wood but the manufacturing plants were afterwards closed down because they were environmentally undesirable and produced an inferior quality product. It is assumed that modern equipment could produce a straw pulp that could be used in place of, or in addition to, wood pulp for many grades of household and packaging papers. This could utilise 1.2 million tonnes of straw. (M.A.F.F. 1984).

6.2 Slab or Board Manufacture

There are in existance in the U.K. two companies manufacturing board, one using the stramit and the other the compak process.

The Stramit factory has, for forty years, produced compressed straw slabs by the continuous heated extrusion method without the addition of resin binders. The present capacity is 25,000 tonnes per annum with a proposed expansion to 40,000 tonnes within the next few years.

The Compak process substitutes straw for wood wool in the manufacture of resin impregnated boards. Chopped straw is mixed with a resin binder and pressed under heated plattens to form boards which are used for crates, pallets, furniture, etc. A pilot plant has operated for one year and utilised 1,000 tonnes of straw; several further plants are under construction and should be operating shortly.

6.3 Chemicals from Straw

Straws vary considerably in their chemical composition according to variety, age, leaf and seed case content, but all contain cellulose, hemicellulose, lignin and a small proportion of organic compounds, salts and insoluble ash. Acid hydrolysis, or the use of enzymes can convert the cellulose into sugars, mainly glucose, from which surfactants, textile chemicals, surface coatings, pharmaceuticals, pesticides and synthetic rubber are manufactured. By further fermentation alcohol, antibiotics, edible and fibrous protein, glycerol and citric acid can be produced. Glucose can also be converted to fructose and other sugars.

The I.C.I. Company has built a pilot plant to use glucose in the manufacture of ethanol, acetic acid, butanol or butanediol, and hemicelluclosic sugars as fermentation feedstock for ethanol, amino acids and furfurol derivatives. If straw were to be become widely used for such purposes, the by-product lignim could be used in adhesives and as fillers in paints, tarmac and concrete, but the problem to date has been the low cost at which the glucose can be produced from other readily available materials.

6.4 Straw for Reinforcing Building Materials

Because of its great tensile strength, straw fibre is an excellent reinforcing agent and is used in the manufacture of concrete building blocks where it also improves the insulation properties.

Work has begun on the substitution of straw fibre for asbestos fibre in asbestos cement products, and on the utilisation of it in the manufacture of moulded concrete structures. If these processes prove economically successful, then between 100,000 - 200,000 tonnes of straw per annum would be used (M.A.F.F. 1984).

6.5 Insulation and Packaging

A limited quantity of straw rope for packaging has been manufactured in this country for a number of years and there would appear to be no increase likely in the near future.

In contrast, an area in which there could be a rapid expansion is for a new product which is used for protective covering and consists of a sandwich of straw between two plastic sheets. At the moment, it is imported into this country but there is the possibility of home manufacture in the near future. (M.A.F.F. 1984).

6.6 Poultry Litter

Chopped straw has been successfully substituted for wood shavings as a poultry litter, but the potential market is limited to a maximum of about 100,000 tonnes per annum (M.A.F.F. 1984).

6.7 Hydromulch

This is a material used in considerable quantities in the U.S. and consists of ground straw suspended in water, and mixed with fertiliser and plant seeds. It is sprayed over large areas of bare land such as the sides of new motorways, to establish vegetation. In view of the current concern in agriculture about the harmful effects on germination of incorporating straw into the soil, it seems doubtful if this practice is likely to spread to any great extent. (M.A.F.F. 1984).

6.8 Chemicals

The University of Manchester Institute of Science and Technology has developed a method of making oil from straw. About 40 kg of 'MANOIL' and 45 kg of carbon dioxide is produced from 100 kg of straw.

Salford University have evaluated the process and say it will only be viable when oil costs £150/t and straw £15/t.

To summarise, unless any significant development occurs, there is unlikely to be any change in the present low demand in the alternatives.

CHAPTER 7

THE FLUIDISED BED TEST RIG

Figure 26 shows the layout of the proposed experimental combustor and anciliaries, and Figure 27 the general arrangement of the combustor unit. It is essentially a hot gas generator utilising the underbed method of feeding by screw conveyor, and built to determine the combustion capability of short chopped straw, either on its own or in conjunction with other fuels.

It was appreciated that the high volatile content and very rapid burning characteristics of straw could mean that a second fuel might be needed to stabilise the bed temperature and maintain combustion. It was also realised that the bed depth might be of great importance in the process of burning off the volatiles; therefore all trials were to be carried out with a range of bed levels.

When the design parameters were laid down, it was decided that the unit should have a target heat output of 100 kW so that it would be suitable for heating a farmhouse and for providing sufficient hot water for domestic and dairy purposes. It was to be made as simply as possible because of the limited funding available and therefore there would be no in-bed tubes for heat transfer. The plan was to control bed temperature by varying the feed rate, and the start-up would be by propane gas burner.

An important aspect of the design was to be concerned with the retention of sufficient heat in the bed to maintain combustion and give as high a standard of efficiency as possible. It was decided to tackle this problem of heat retention in two ways. Firstly, the elutriation of the hot sand and unburned carbon particles would be reduced as far as possible by combustion chamber design,and for this reason the cross-sectional

PROPOSED ARRANGEMENT OF COMBUSTOR AND ANCILIARIES

FIGURE 26.

(NOT TO SCALE)

GENERAL ARRANGEMENT OF EXPERIMENTAL FLUIDISED BED COMBUSTOR - FIG 27 area of the freeboard section was to be rapidly enlarged immediately above the expanded bed,taking in part of the splash zone. Secondly, the facility to run in both intermittent and continuous fluidisation states (by means of a cam-operated interruptor in the air supply) was also to be incorporated.

In order to minimise the cost, a porous tile was to be used as the distributor plate. It was hoped that this would both give a uniform air flow which would lead to good fluidisation, and act as a support for the bed when in the slumped condition. The P.4. grade of ceramic tile chosen was manufactured by Doulton Industrial Products Ltd. of Stone, Staffordshire. This has 0.02 mm perforations - small enough to prevent back flow of bed materials but numerous enough to give the air distribution over the operating range without the need for an excessively high pressure supply of air. Failure to mix fuel and air efficiently can seriously effect combustion and cause bed sintering, or spouting, creating excessive elutriation of bed particles.

The unit is 3.35 m high and comprises a lower section housing the combustion equipment, plenum chamber, distributor plate and feed mechanism, and an upper section which includes the expansion chamber,secondary air supply and cyclone separator. The fluidised bed itself is 365 mm square with the freeboard expanding to 1200 mm square above the splash zone to give an expansion chamber volume of 1.73 m^3. (Photograph No. 1).

The outer casings of the bed containment and the expansion chamber were made from 6 mm steel plate and were lined with a mixture of high abrasion resistant, 1200°C quality castable refractory (Steetley 1450), and firebricks of similar specifications. Between the refractory and the steel in the upper section was positioned a high compressive strength calcium silicate board insulation 50 mm thick. The plenum was also built in castable refractory and the zone between it and the steel casing was filled with a weak sand and cement mixture.

Fluidising air was supplied by a 5.5 kW, Type 6110, centrifugal fan manufactured by Fans and Blowers Ltd. of Highbridge, Somerset. The air was introduced into the plenum through a curved duct of square cross section (100 cm^2); it then passed through the ceramic tile and into the bed, which consisted of inert silica sand with a particle size of $\simeq$ 0.9 mm. (Photograph No. 2).

Straw was fed from a variable output hopper via an inclined rubber belt conveyor to the rotary seal mounted above the unit's intake screw feeder. The latter was manufactured from a 75 mm diameter cast iron screw taken from a Hodgkinson Bennis underfed stoker unit; it ran in an 80 mm internal diameter Incoloy 800H tube which had a wall thickness of 3 mm. (Photograph No. 3)

The tube projected into the bed by 178 mm and to reduce the temperature rise along it caused by heat transfer from the bed, a water jacket was fitted around it which extended from the plenum chamber to the intake hopper. In order to enable the volatiles to burn whilst the char remained in the feeder, ten 8 mm dia. holes (five spaced along each midway line) were drilled in the tube where it was immersed in the sand.

The screw stopped some 260 mm short of the end of the tube so that during combustion, the mass of char and ash would compress to form a plug. The idea was that it would create a seal and reduce the chance of gases escaping from the bed and into the feeder system.

Under normal operating conditions it was hoped that combustion losses in the form of carbon monoxide and unburned volatiles would be negligible. One major problem with F.B.C. is that of elutriated unburned carbon; this can be particularly serious with fuels of high fines content, where a long residence time in the bed is essential. To ensure the maximum residence and therefore mixing of fuel and air, the straw screw feeder was positioned only 25 mm above the distributor plate.

As a further measure, in case the volatiles were not completely burned, secondary air to complete combustion could be introduced above the splash zone. This was designed so that it could be supplied either by the fluidising air fan or by a separately mounted Sturtevant type 00 fan. The air was introduced through four nozzles positioned tangentially, one on each of the four sides of the expansion hopper and 300 mm above the bed. This arrangement was chosen to create turbulence, mixing of combustion gases and volatiles and, it was hoped, to make the chamber act like a cyclone to drop particulate matter back in the bed.

From the expansion chamber, the exhaust gases passed into a Van Tangeron powerclone separator to remove elutriated sand, ash and unburned carbon particles; these were to be collected from the bottom of the cyclone in a quick release sample jar. (Photograph No. 3).

Thermocouples manufactured by the Industrial Pyrometer Co. Ltd. with nickel chrome/nickel aluminium elements and nickel chrome 36/17 sheaths, suitable for temperatures up to 1200°C were fitted in the plenum, bed, splash zone and freeboard areas. These were connected to a Comark Temperature Recorder which continuously monitored the temperatures. (Photograph Nos 4 & 5).

The temperature of the cooling water return from the screw feeder cooling jacket was indicated by a 65 mm dia. surface contact British Rototherm model BL 301 thermometer with a 10° - 120°C range. The flow of cooling water through the jacket was set at 200 litres an hour.

In order to inspect the plenum for flamefailure and adjustment, and to observe the fluidised bed during operation of the unit, sight glasses were fitted in the appropriate places, one in the plenum and one above the bed.

To start-up the unit, an Aeromatic two-stage gas injector fitted with a F.R.1 burner nozzle (manufactured by Aeromatic Industrial Ltd., Gerrards Cross) was fitted. This was built into a pre-combustion unit which was bolted on to the side of the plenum chamber hopper in a way which allowed it to be easily removed.

Propane gas from a regulated cylinder was connected to the inlet of the injector and air introduced into the mixing tube from a connection on the primary air duct. The gas/air mixture then passed into a second venturi-shaped mixing tube which created a low pressure area at the throat of the venturi, drawing in air through a second series of air parts. When the gas pressure was low, the air/gas ratio was automatically reduced; as the pressure increased, the percentage aeration also increased, resulting in a very stable flame.

The pre-heater was designed to be lit externally and fitted when correct flame shape had been obtained. This method was chosen to avoid the possibility of an accumulation of gas in the unit should any malfunction occur during an in-situ lighting-up procedure.

The cam-operated fluidising air interruptor mechanism consisted of a butterfly valve mounted in the curved air duct just before the plenum chamber. The butterfly spindle carried a lever to the outer end of which was fastened a ball race which ran on the exterior cam. The camshaft was connected by a multi-speed vee-belt drive to a fractional horsepower geared electric motor. The shaft speed could be varied by changing pulleys and the motor was supported from the main frame by an adjustable plate which allowed the belt to be tightened. (Photograph No. 6).

The whole unit (combustor, fans, feeder, seal and electrical starter panel) was supported on a steel frame made from 100 x 50 mm channel, the base area being 3 x 1.25 m and the total weight approximately 2.5 tonnes.

Calculations were made to confirm the design assumptions (Appendix 1) and were based on the straw analysis produced by B.C.R.A. Scientific and Technical Services 1985.

CHAPTER 8

INITIAL TRIALS WITH THE F.B.C. TEST RIG

8.1 Cold Trials

Before the expansion chamber was fitted in position, all motors were tested and trials were carried out at various conveyor speeds to determine the feeding rates of chopped straw. No distribution plate had been fixed at this stage so straw could be collected and weighed. (Table 9).

The first problem to emerge was that as straw left the end of the screw, it became excessively compacted within the tube. The auger was unable to move this and it eventually stalled. The problem was overcome by cutting two 8 mm wide slots along the sides of the tube where the holes had been. The slots extended 165 mm from the end of tube and the tube was also heated and 'belled-out' by some 3mm at the end to reduce the restricting pressure. This proved successful on all future trials.

The expansion chamber was then bolted in position and the cyclone and chimney fitted. Thermocouples were inserted in the plenum, bed splash zone and expansion chamber, and connected to the temperature recorder. The distributor plate was fitted and bedded in with asbestos webbing packing.

The main fan was started up and pressure readings were taken both with and without the air flow interruptor gear operating (Table 10).

The bed sand was graded by B.S. mesh sieves to determine the mean particle diameter (Table 11) from which initial fluidising velocities were to be calculated (Table 12). A weighed amount of sand was poured through the open sight glass in the expansion chamber to settle in the bed at the first trial test depth of 178 mm. This level was confirmed by depth gauge painted on the inside of the sand containment section.

TABLE 9 - FEEDING SCREW TRIALS

Quantity of Chopped Straw to give Theoretical Output of $\simeq$0.1 MW (360000 kJ/h) at Gross Calorific Value of 16189 kJ/kg			22.24 kg/hr
Feeding Screw:	Diameter of Ribbon	70mm	
	Diameter of Shaft	20mm	
	Conveying Area		3.53^{-3} m^2
First Run:	Motor 125 rpm, 19T Sprocket		
	Screw 63 rpm, 38T Sprocket		
	Weight of Straw conveyed		88.2 kg/hr
Second Run:	Motor 125 rpm, 15T Sprocket		
	Screw 49 rpm, 38T Sprocket		
	Weight of Straw conveyed		68.4 kg/hr
Third Run:	Motor 125 rpm, 15T Sprocket		
	Screw 26 rpm, 72T Sprocket		
	Weight of Straw conveyed		36.0 kg/hr

For mechanical reasons the 125/26 rpm ratio was used and the input to the screw restricted to 22.24 kg/W rate

TABLE 10

MAIN FAN TEST (1) - 5.7.85

Ambient Temperature - 24°C

Barometric Pressure (985.5 mbar) 739 mm Hg   Static Pressure 28" WG

Test carried out without sand in bed

Plenum cut off slide in position 1/4, 1/2 and fully open

VELOCITY PRESSURES

Butterfly valve position	degree	Slide 1/4 open P_v	Slide 1/2 open P_v	Slide fully open P_v
Closed	0	0.095	0.115	0.125
Open	15	0.105	0.170	0.185
Open	30	0.135	0.220	0.265
Open	45	0.160	0.290	0.375
Open	60	0.170	0.345	0.445
Open	75	0.180	0.390	0.550
Fully Open	90	0.185	0.445	0.680

AIR VELOCITIES were computed

Using the equation $V = \sqrt{\frac{760}{B} \times \frac{T}{293} \times \frac{10350}{10350+P_s} \times P_v}$

where
- V = Velocity m-s
- B = Barometric pressure mm Hg
- T = absolute temperature °K (=t°C + 273 where t is airstream temperature)
- P_s = Static pressure mm H_2O
- P_v = Velocity pressure mm H_2O

cont....

TABLE 10 - Cont.

Butterfly valve position	Degree	Slide 1/4 open m-s	Slide 1/2 open m-s	Slide fully open m-s
Closed	0	1.275	1.402	1.462
Open	15	1.340	1.705	1.779
Open	30	1.520	1.940	2.129
Open	45	1.654	2.227	2.532
Open	60	1.705	2.429	2.758
Open	75	1.754	2.582	3.067
Fully Open	90	1.779	2.758	3.410

TABLE 11 - SAND SPECIFICATION

Graded Quartz Sand	Grain Shape	- Rounded
	Colour	- Silver Grey
	Moh's hardness	- 7
	Specific Gravity	- 2.65

Typical Chemical Analysis:

	%
SiO_2	99.50
$AlZO_3$	0.22
$FeZO_3$ & TiO_2	0.05
CaO & MgO	0.07
$K_2O(NaZO)$	0.04
L.O.I.	0.10

Typical Grading Analysis:	Size range	Undersize (Max)	Oversize (Max)
Grade 443	0.7-1.25	5%	5%

Sieve Screen Analysis:

B.S	Metric	% on Sieve	Mean Particle Size
+16	1000	91.58	
+18	850	7.79	0.984 mm
+22	710	0.50	
+25	600	0.13	

$$\text{Mean Particle Size} = \frac{1}{\sum \frac{x}{Dp}}$$

Dp	x	$\frac{x}{Dp}$
1.00	0.9158	0.9158
0.850	0.0779	0.0916
0.710	0.0050	0.0070
0.600	0.0013	0.0022
		1.0166

$$d_p = \frac{1}{1.0166}$$

$$= 0.984 \text{ mm}$$

TABLE 12 - MINIMUM FLUIDISING VELOCITY (Umf)

Calculations for fine sand at 0.984 mm mean particle size using KUNII and LEVENSPIEL formula for fine sand

$$Umf = \frac{d_p{}^2 (p_s - p_f) g}{1650 \mu_f} \text{ m/s} \qquad \text{where } R_{ep} < 20$$

At 850°C

d_p	= mean size of inert particles	$= 0.984^{-3}$ m
p_s	= density of particles	= 2650 kg/m³
p_f	= density of fluidising gas	= 0.314 kg/m³
g	= acceleration due to gravity	= 9.81 m s
μ_f	= viscosity of fluidising gas	$= 4.45^{-5}$ kg/m s
R_{ep}	Reynolds number of inert particles	

$$= \frac{U_{mf}\, d_p\, p_f}{\mu_f}$$

$$U_{mf} = \frac{(0.984^{-3})^2 \; (2650-0.314) \times 9.81}{1650 \times 4.45^{-5}}$$

$$= \frac{9.683^{-7} \times 2649.886 \times 9.81}{7.3425^{-2}}$$

$$= \underline{0.343 \text{ m s}}$$

$$R_{ep} = \frac{U_{mf}\, d_p\, p_f}{\mu_f}$$

$$= \frac{0.343 \times 0.984^{-3} \times 0.314}{4.45^{-5}}$$

$= \underline{2.38}$ which is <20

8.2 Refractory Curing

The bed sand was emptied via a drain tube, the plenum lowered and the distributor plate removed. After the plenum had been refitted, the unit was ready for the curing of the refractory. The burner unit was connected to the pressure regulator of a 47 kg propane cylinder and lit externally, the air/fuel ratio being adjusted manually to give a green/blue flame which indicated approximately the required air/fuel ratio of 15 : 1. When the flame was established, the burner was introduced into the plenum chamber and secured with bolts.

Curing was carried out according to the recommendations of Steetley Refractories Ltd. (Fig. 13) and took 24 hours to complete.

8.3 Bed Expansion Tests at Ambient Temperature

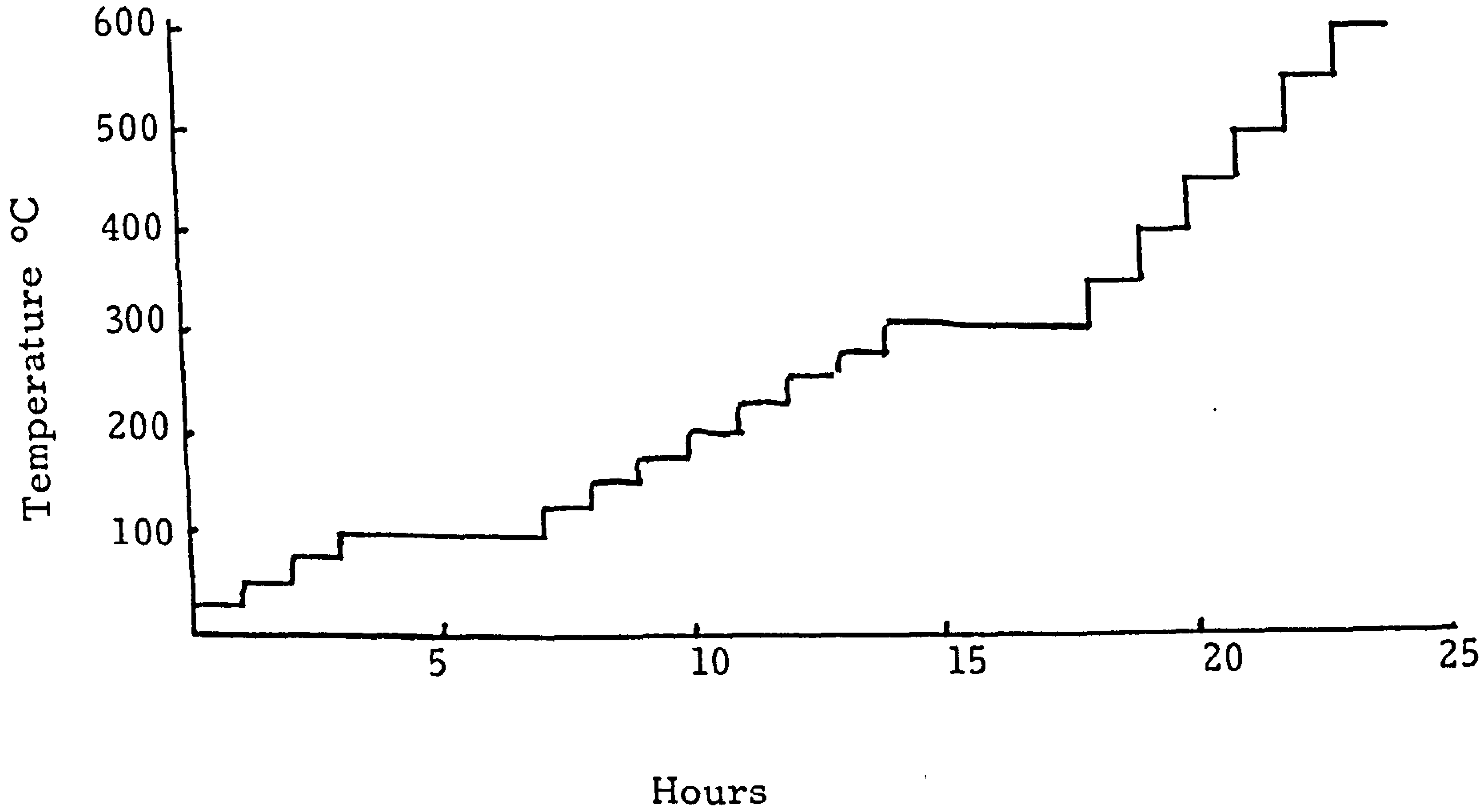
After curing, the distributor plate was replaced and the bed sand was introduced. Before the hot trials started, a series of tests was carried out to establish bed expansion from different depths of slumped bed starting at 178 mm, with and without intermittant fluidisation. Static pressures were recorded (Table 14) and at the sand depth of 178 mm, the effect of the position of the butterfly valve on the velocity of air through the bed was established. A pitot tube, coupled to an Airflow Developments Ltd. inclined manometer was used for air flow measurement and readings at 15° increments of the position of the cam follower arm throughout its entire range were made.

With a particle specific gravity of 2.65, the theoretical weight of the solid bed at 178 mm deep would be 46.72 kg (0.018 m^3). The actual weighed amount of sand for this area was 32 kg equating to a bed voidage (ε) of 0.32 at 20°C.

TABLE 13

REFRACTORY CURE

Number of hours	at	°C
1st		25
2nd		50
3rd		75
4th-7th		100
8th		125
9th		150
10th		175
11th		200
12th		225
13th		250
14th		275
15th-18th		300
19th		350
20th		400
21st		450
22nd		500
23rd		550
24th		600



From the velocity pressure readings, the actual air velocities were calculated (Table 14). Although at the higher velocities fluidisation occurred it was not particularly vigorous and it appeared that the capacity of the fan was not being reached, it was decided that the design and position of the butterfly could be improved. Modifications that involved both the shape and position of the cam were made. The velocity /butterfly position determinations were then repeated with the plenum slide fully open (Table 15).

At the same time, the burner air duct diameter was increased from 25mm to 38 mm.

TABLE 14 - MAIN FAN TEST (2) - 8.8.85

Ambient Temperature - 19°C   Fan Air Temperature - 20°C

Barometric Pressure (1005 mbar) 754 mm Hg

Tests carried out with sand 0.984 mm in bed.

Bed Expansion

Bed depth Slumped mm	Bed depth Expanded mm	Static Pressure (fan outlet duct) mmH_2O
152	178	292
165	187	300
178	197	305
191	203	308
203	216	311

Intermittant Fluidisation

Bed depth		Static Pressure (fan outlet duct)	
Slumped mm	Expanded mm	Minimum mmH_2O	Maximum mmH_2O
152	178	121	400
165	187	124	403
178	197	127	406
191	203	130	409
203	216	133	412

Velocity Pressure

Sand Bed 178 mm deep - Plenum cut off slide 1/4, 1/2 and fully open

Butterfly valve position		Slide 1/4 open	Slide 1/2 open	Slide fully open
	Degree	Vel m/s	Vel m/s	Vel m/s
Closed	0	1.72	2.43	2.53
Open	15	1.74	2.53	2.81
Open	30	1.75	2.62	2.93
Open	45	1.77	2.72	3.03
Open	60	1.77	2.75	3.08
Open	75	1.79	2.78	3.12
Fully open	90	1.87	2.79	3.14

TABLE 15 - MAIN FAN TEST (3)

Velocity Pressure

Sand Bed 178 mm deep - Plenum cut off slide 1/4, 1/2 and fully open

Butterfly valve position	Degree	Slide 1/4 open Vel m/s	Slide 1/2 open Vel m/s	Slide fully open Vel m/s
Closed	0	3.03	3.14	3.19
Open	15	3.20	3.35	3.39
Open	30	3.42	3.60	3.66
Open	45	3.62	3.82	3.90
Open	60	3.70	4.08	4.24
Open	75	3.79	4.14	4.40
Fully open	90	3.89	4.32	4.62

CHAPTER 9

PRELIMINARY STRAW BURNING TRIAL

The propane gas burner was lit at the start of the trial which was planned to run at a bed temperature of 800°C; the straw being fed when 600°C had been reached.

The plenum temperature rose steadily but at 625°C, with the bed registering 318°C, it dropped rapidly to 562°C. Within a half an hour it had climbed back to 610°C, the bed temperature being 400°C but the body of the rig during this time had risen from 50° - 204°C. By the time the bed had reached 450°C, the plenum registered 720°C and the rig casing, 222°C.

Realising the trial could not carry on for any length of time, it was decided to start feeding straw to see what would happen at that temperature. In fact, the plenum temperature remained steady, the bed temperature rose to 498°C and the casing temperature went up to 252°C (Table 16). At these temperatures, the burner was turned off and the plenum temperature started to fall, although the casing temperature rose. The bed temperature continued to rise showing that heat was being released from the straw but after ten minutes with the casing temperature at 262°C, the straw feeder was stopped.

The heat loss was obviously too great to allow the trial to proceed with safety and the unit was allowed to cool down so that it could be examined. The actual running temperatures are shown in Table 16.

On dismantling the rig, it was found that the straw being fed, which was between 150 - 250 mm in length, had tended to wrap around the feed screw, causing some burn back. A further more serious problem was found, namely that a split had developed in the plenum across the thermocouple access hole and the porous tile had cracked. Heat was therefore being released

TABLE 16

PRELIMINARY STRAW COMBUSTION TRIAL

TIME		TEMP°C	
	PLENUM	BED	CASING.
1345 START UP			
50	185	25	20
55	295	45	26
1400	390	80	31
5	435	214	38
10	490	245	53
15	550	278	68
20	563	298	86
25	600	300	90
30	625	318	100
35	610	320	110
* 40	562	330	132
45	583	347	154
50	595	373	181
55	601	388	192
1500	610	400	204
5	618	409	211
10	642	421	218
15	726	439	224
** 20	720	450	229
25	721	461	234
30	721	469	240
35	723	481	245
*** 40	700	498	252
45	695	512	257
50	680	530	262
55 SHUT DOWN			

* INITIAL FRACTURE OF PLENUM

** STRAW FEEDER STARTED

*** BURNER TURNED OFF.

directly from the plenum on to the outer casing. An attempt to repair the damage with castable cement and steel banding the plenum was unsuccessful; it was also obvious that the porous tile was unable to stand the thermal shock it had been subjected to so modifications were required before the trial could be repeated.

The screw was also moved 40 mm into the tube reducing the distance from the end to 220 mm, and 42 mm from the inside wall of the fluidised bed.

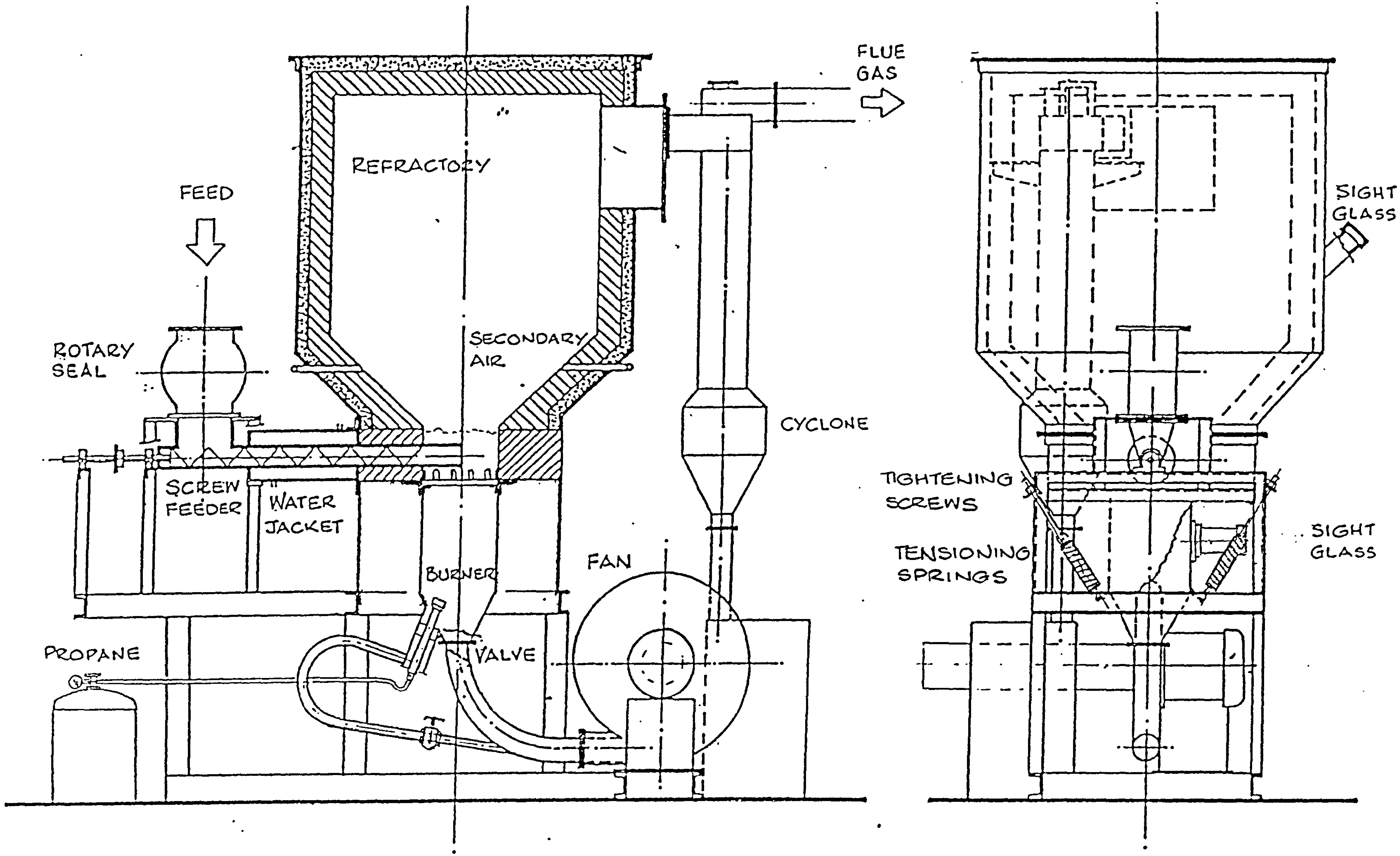
CHAPTER 10

MODIFICATIONS TO THE F.B.C. TEST RIG

Following the problems with the porous tile and refractory cast plenum, it was decided to replace them in stainless steel. A distributor with standpipes was also chosen instead of a perforated plate because of the history of cracking under thermal shock, and this was located inside the top of the new plenum. This was positioned using a Kos refractory cement and gasket beneath the combustion chamber and held in place by eight springs. These were attached to lugs welded on the outside of the plenum and connected to tensioning screws fitted to the body of the unit (Fig. 28). This method of attaching the plenum was used to accomodate changes in dimensions during heating and cooling processes.

The upward pressure after the screws had been tightened held the plenum tightly in position, eliminating air leakage.

The main fan connection and burner assembly were made and trials were again carried out.



GENERAL ARRANGEMENT OF EXPERIMENTAL FLUIDISED BED COMBUSTOR - FIG 28

CHAPTER 11

TRIALS WITH CHOPPED STRAW AS A FUEL IN THE UNINTERRUPTED AND INTERRUPTED FLUIDISATION MODE

The trial started when the propane burner was lit and introduced into the plenum with the air slide and butterfly valve both closed. The air slide was slowly opened and the butterfly valve cam adjusted by hand as the plenum temperature rose; in approximately fifteen minutes the plenum and bed temperatures both reached 520°C. At this point chopped straw was fed into the system at a predetermined rate of 22 kg/hr.

The bed temperature then rose quickly and the propane burner was shut off when it reached 675°C. The air slide was then fully opened and as the bed temperature reached 720°C, the cam was turned to set the butterfly valve into the fully open position. At this feed rate, the bed temperature fluctuated between 780 - 820°C. At this point, a problem developed namely that smoke started to emerge from the open rear end of the straw feed auger. It was obvious that a section of the feed tube behind the end of the auger had become hot enough to release volatiles from the straw in contact with it and had ignited and that their easiest escape route was back along the auger against the supply of straw. A partially successful temporary seal was made and was fitted to the revolving auger shaft without stopping the trial.

Flue gas composition observations were made and ash deposits were collected and weighed at regular intervals.

The cam operated butterfly valve was then started, the variable speed drive giving a five second cycle which included approximately one second when the bed was slumped; no difference in the bed temperature was recorded. (Table 17).

TABLE 17 SUMMARY OF RESULTS OF TEST RUNS.

FUEL - CHOPPED STRAW GROSS CALORIFIC VALUE 16.2 MJ/kg.

	BED TEMP °C	FUEL RATE kg/h	FUEL IMPUT MJ/h	PRIMARY AIR FLOW RATE m^3/h	PRIMARY AIR FLOW TOTAL m^3/h	EXCESS AIR %	FLUIDISING VEL. m/s	FLUE GAS CO_2 VOL %	FLUE GAS O_2 VOL %
(1)	800	22	356.4	165	165	52	1.3	8	4
(2)	800	22	356.4	* FLUIDISED 178	142.4	64.4	1.41	10	2
				NOT FLUIDISED 105	21.0	0	0.83		

	UNINTERRUPTED FLUIDISATION %	INTERRUPTED FLUIDISATION %
CARBON CONVERSION	99.14	99.35
COMBUSTION EFFICIENCY	99.30	99.47
THERMAL EFFICIENCY	64.52	68.52

* BASED ON THE BED BEING FLUIDISED FOR 80% OF THE TIME AND A RAPID CHANGE FROM THE FLUIDISED TO UNFLUIDISED STATE AND VICE VERSA.

The static pressure dropped each time the bed slumped and then recovered, reaching 28.5 ins W.G. (some 0.5 ins W.G. higher than when fluidisation was continuous).

Flue gas composition and ash sampling was again undertaken.

Eventually the seal failed and due to the back burning the subsequent build-up of smoke in the building housing the rig forced the termination of the test run. After shut-down the bed was examined and was found to be remarkably clean. The only fuel particles in it were a few small pieces of char found near the outlet to the feeder; these presumably came from the material left in the auger after it was stopped.

Whilst operating in the interrupted fluidisation mode it was noticeable from sampling that the cyclone catch was lower than when in the uninterrupted mode. The loss of bed material during the interrupted mode was also less. The rates of loss are shown in Appendix 2.

CHAPTER 12

TRIALS WITH PELLETED STRAW AS A FUEL

To evaluate combustion efficiency of pelleted straw, a trial was carried out at Olney on the 1.6m² bed F.B.C. hot gas generator. The fuel tested was composed of 95% barley straw (ground through a 6 mm screen) and 5% powdered bitumen; the bitumen acted as a binder in the pellets, which were produced in a ring press fitted with a 15 mm diameter hole die. The bitumen increased the gross calorific value of the straw from 15000 to 19700 kJ/kg but had no measurable effect on the relatively low ash fusion characteristics of the straw (Table 18)

It was decided to carry out the trials at various bed temperatures to determine the freeboard temperature at each bed temperature. Freeboard temperature is higher than bed temperature because of the combustion of fines and volatiles above the bed; it was also necessary to study the effect that different operating conditions had on the fusion of the ash particles.

Automatic control of the bed temperature level was achieved by the introduction of recycled gas which was regulated by a valve operating in conjunction with the bed temperature (set point selected on a Texas PM550 controller).

Pellets were fed by screw conveyor on to the top of the fluidised bed and to ensure complete combustion, secondary air was introduced above the bed. From experience gained with the unit while burning coal, with its lower volatile content, it was expected that when burning straw, there would be less heat release in the bed. Primary air was therefore reduced and the secondary air was increased. This increased the proportion of combustion taking place above the bed and the quantity of recycle gas was automatically reduced to compensate for the lower in-bed temperature.

TABLE 18

STRAW PELLET BURNING TRIAL ON COMMERCIAL 1.6 m^2 F.B.C. HOT GAS GENERATOR AT THE ENERGY EQUIPMENT Co LTD. - OLNEY

COMPOSITION OF FUEL PELLET 95% STRAW : 5% BITUMEN

ULTIMATE ANALYSIS BY B.C.R.A. - CHESTERFIELD.

	%
MOISTURE	7.9
ASH	8.8
VOLATILES	66.0
FIXED CARBON	17.3
	100.0

GROSS CALORIFIC VALUE 19700 kJ/kg

ASH FUSION CHARACTERISTICS

DEFORMATION	980°C
HEMISPHERE	1175°C
FLOW	1230°C

The trial commenced with a bed temperature of 825°C, the freeboard temperature rose to 1040°C before stabilising. On increasing the bed temperature to 850°C, the freeboard temperature rose to 1059°C and on further increasing the bed temperature to 900°C, it rose to 1060°C. Even at the lowest of these freeboard temperatures, fused ash deposits were produced above the bed on the walls and ducts of the HGG. The ash was soft and grey in colour initially but hardened on cooling, becoming friable and easy to remove.

From the experience gained, it seemed that the only practical way of operating the F.B.C. would be at a bed temperature no greater than 850°C, with excess air in the bed to burn off the volatiles and no secondary air.

A second trial was then carried out in which the bed temperature was kept between 800°C and 840°C; the supply of primary air was increased and no secondary air was used. At this setting, the freeboard temperature never exceeded 860°C - a level well below the ash fusion limit (Table 19).

It was obvious from the information obtained during these trials that straw, with its high volatile content, would have to be burned at a lower bed temperature than that normally associated with fuels of lower volatile content.

A third trial was therefore carried out on a larger, 3.5m² F.B.C. at the British Sugar Corporation, Brigg (Table 20) which had three bed observation ports fitted. Thermocouples were fitted beneath each of the two screw feeders positioned at one end and one thermocouple opposite to them. All three were connected to a Texas PM550 computer which automatically controlled bed attemperation by varying the flow of recycled fluegas to maintain the pre-determined set point temperature, and displayed an average bed temperature.

TABLE 19 SUMMARY OF RESULTS OF STRAW PELLET BURNING TRIALS (1) and (2)

AT THE ENERGY EQUIPMENT Co LTD, OLNEY'S $1.6 m^2$ HOT GAS GENERATOR

FUEL- BITUMEN BOUND STRAW PELLETS GROSS CALORIFIC VALUE 19.7 MJ/kg.

TRIAL	DATE	TIME	APPROX FEED RATE kg/h	APPROX FUEL INPUT GJ/h	TEMP. °C BED	TEMP. °C FREE-BOARD	GAS ANALYSIS CO_2 Vol. %	O_2 Vol. %	CO Vol. %	NO_x P.P.M.	SO_x P.P.M.	PRIMARY AIR kg/h	SECOND. AIR kg/h
1	26.9.85	1600	411	8.1	814	884	18	0.6	0.3	112	100	1768	869
		1630	411	8.1	821	970	17	1.9	0.3	125	75	1768	900
		1700	450	8.8	824	1004	17	2.5	0.3	137	55	1768	1068
		1740	436	8.6	853	1034	17	2.8	0.3	143	50	1768	1068
		1815	436	8.6	892	1050	16	3.0	0.3	156	60	1768	1068
		1900	436	8.6	895	1060	16	3.0	0.3	168	50	1768	1068
2	27.9.85	1140	436	8.6	700	591	16	3.8	0.3	N.M.	MM.	2288	0
		1150	444	8.7	866	820	16	6.3	1.5	"	"	2288	0
		1200	444	8.7	820	763	15	120	0.3	"	"	2288	1005
		1220	444	8.7	844	815	9	7.3	0.3	"	"	2288	0
		1240	444	8.7	846	861	13	7.0	0.3	"	"	2288	0
		1245	444	8.7	800	861	13	7.0	0.3	"	"	2288	0

TABLE 20 SUMMARY OF RESULTS OF STRAW PELLET BURNING TRIAL (3)

AT THE BRITISH SUGAR CORPORATION'S — BRIGG, $3.5 m^2$ HOT GAS GENERATOR

FUEL - BITUMEN BOUND STRAW PELLETS. GROSS CALORIFIC VALUE 16.3 MJ/kg.

TRIAL	DATE	TIME	APPROX FEED RATE kg/h	APPROX FEED INPUT GJ/h	TEMP °C BED	TEMP °C FREE-BOARD	CO_2 VOL %	PRIMARY AIR kg/h
3	12.12.85	2100	1000	16.3	799	982	3	7352
		2130	1000	16.3	799	982	3	7352
		2200	1000	16.3	799	983	4	7352
		2230	1000	16.3	800	984	7.5	7352
		2300	1000	16.3	801	989	8	7352
		2330	1000	16.3	799	994	8.5	7352
		2400	1000	16.3	800	1003	8.5	7352
	13.12.85	0030	1000	16.3	801	1001	8.5	7352
		0100	1000	16.3	800	1005	8.5	7352
		0130	1000	16.3	800	984	8.5	7720
		0200	1000	16.3	800	981	8.75	8106
		0230	1000	16.3	799	966	9	8106
		0300	750	12.2	800	959	9	8106
		0330	750	12.2	799	962	9.75	8106

It was realised early in the trial that an even temperature throughout the bed was not being obtained although the set point and average bed temperature recordings were similar. Individual thermocouple readings showed a difference of 300°C across the bed with the highest temperature being recorded beneath the feeders. The single thermocouple was therefore eliminated from the Texas programme (although still displayed) and the bed temperature controlled by averaging the temperatures indicated by thermocouples installed beneath the feeders.

The feeders could not be seen through the sight glasses as they were obscured by the burning volatiles which evolved within the feeder itself. Most of the burning pellets were to be seen in the area beneath the feeders, with some in the centre of the bed and very few at the end furthest from the feeders.

Fused particles of sand and ash could be clearly seen on the roof and walls above the feeders, indicating a local temperature above the ash fusion point although the outlet gas temperature was well below this. Obviously again an average mixed gas temperature was being recorded.

It became apparent that the relatively quickly burning straw pellets when fed on to this large area of fluidised bed, had pyrolised before migration across the bed was complete in complete contrast to that exhibited by the slower burning coal. It would therefore seem essential that on a larger FBC, straw pellets should be fed into the bed at several places to ensure an even distribution of fuel; this should then give relatively uniform bed temperatures, heat outputs and freeboard and outlet gas temperatures.

DISCUSSION

The majority of existing straw burning furnaces operate on a batch feed system which allows little control over the burning rate and almost none over the amount of excess air. The frequency of feeding is often determined solely by the need to avoid the unit burning out. Some units have been modified at considerable expense to allow continuous feeding or have been linked to an extremely large hot water tank (e.g. 20,000 l) in order to minimise fluctuations in water temperature.

As a fuel, straw is far from ideal in that it is not particularly dense, it has a very high volatile content which quickly releases heat when combusted, and an ash residue which is composed of fine, light particles. The ash has a high silica content and an initial fusing temperature of 1045°C. In any fluidised bed combustion efficiency is governed by a number of factors, the principle ones being:

Fuel characteristics
Bed temperature
Excess Air
Fluidising velocity
Bed depth
Freeboard height
System pressure

Combustion efficiency improves with fuel reactivity, bed temperature, bed depth, freeboard height, excess air and system pressure whilst increasing fluidising velocity reduces efficiency. To obtain accurate and instantaneous information on the conditions during the combustion of any fuel requires a high degree of instrumentation.

The cost of temperature recorders, gas analysers and fuel metering devices is very high and normally associated with commercial test plant such as the unit at the Energy Equipment Company Limited's Research and Development Centre in Olney (Fig. 29). In this unit the primary air fan has been replaced by compressed air stored in a cylinder metered by rotameter and controlled by a pressure reducing and regulating valve from 0-20 kg cm^2. This facility allows work to be extended to deeper beds than is possible using a fan; in consequence the unit is able to retain fuel longer, burn a higher proportion of volatiles in the bed and cut down losses by elutriation more than is normal.

In agriculture, we not only have to deal with foodstuffs but have stringent financial constraints to work to; in crop drying for example, economics dictate that the exhaust from the combustion system must be clean enough for direct drying because of the prohibitive expense and reduced thermal efficiency associated with a heat exchanger. Different industries and processes have different requirements - for example, the level of contamination allowable in hot gases passing into a brick kiln may be substantially different to those for use in a plant which is drying feedstuffs, similarly there will be differences between incinerators and units used in the metal processing industry.

The products of combustion which are potential contaminents in materials to be dried are oxides of sulphur (SOx) and nitrogen (NOx), polycyclic aromatic hydrocarbons (P.A.H.) condensed pyrolytic volatiles (smoke) carbon particles (soot or char) and ash. With straw, SOx problems are relatively minor because the sulphur content of straw (0.1%) is very small and in any case, the inclusion of limestone

NOTE: ALL TIs ARE TO COMMON CHART RECORDER
ALL PIs AND dPIs WILL BE MANOMETERS (EXCEPT AIR SUPPLY)

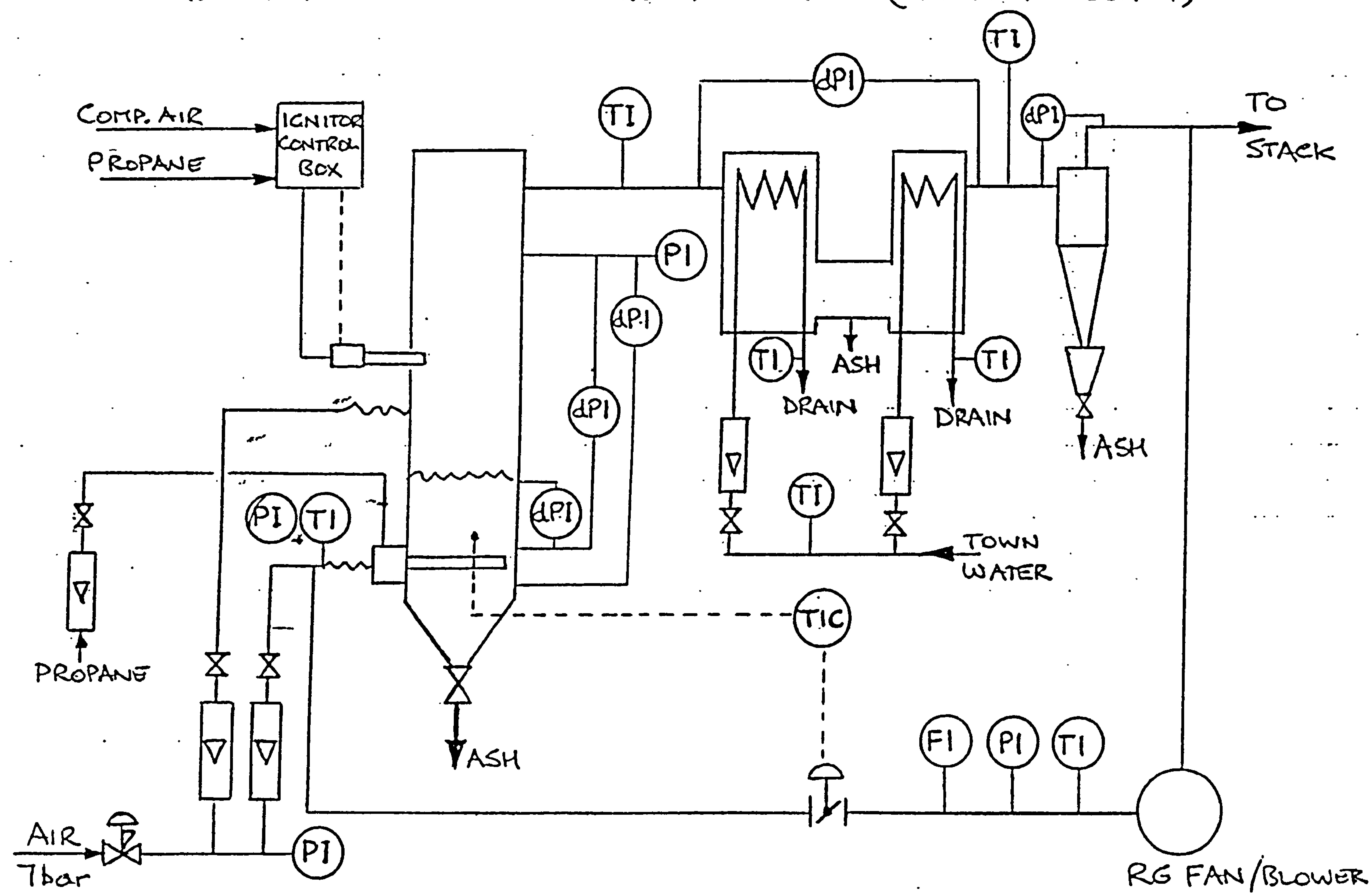


FIG 29 SMALL RIG INSTRUMENTATION

or dolomite in a fluidised bed, especially if it is a deep one, will reduce SO_x levels substantially. The oxides of nitrogen will also cause little concern at the relatively low temperature in most F.B.C.s where the oxides of nitrogen from atmospheric nitrogen account for only about 10% of the total nitrogen oxides produced, (JONKE et al 1969). The remainder is derived from the nitrogen contained in the fuel which, with straw, is again very low (0.8%). However, there is some cause for concern about the P.A.H.s. Some of these compounds are carcinogenic and carbon, char smoke and ash can contaminate products being dried. P.A.H. emission is influenced by the design and operation of the combustor, especially the combustion conditions, whilst soot formation is dependent upon bed residence time, temperature and turbulence, all of which have been given due consideration on the Test Rig. Work on the use of biomass as a fuel in the U.S. indicated that ash emission was proportional to the fuel ash content (PAYNE et al. 1981) which in the case of straw is very low (3.9%).

In a top feed fluidised bed combustor there is a strong tendency for light particles of fuel and ash to be immediately blown out of the combustion zone, so to control these problems it was decided to burn the straw with as low a bed temperature as possible so that any uplift by overbed burning of the volatiles would not raise the temperatures of the gases to the point of ash fusion. Previous experience with straw pellets in the Energy Equipment Company Limited's F.B.C. at Olney had shown that 850°C was a safe bed temperature and avoided the ash build up; this temperature was therefore used for initial combustion calculations. (It was later found in practice that 800°C was quite sufficient for burning the chopped straw). (Table 16).

A decision had to be made during the design stage on the size of the fluidised bed and it became apparent that this was going to be determined by the size of the feeding system to be used. From previous experience of conveying short chopped straw, a 75 mm screw auger was chosen as the minimum size to work with; it was also decided that the area of the feeder should occupy no more than 10% of the area of the bed. This resulted in a minimum bed size of 355 mm square.

The standard porous tile for air distribution was 400 mm square - large enough for it to be recessed between the combustion chamber and plenum and still leave a 355 mm square bed area.

The only combustion air fan available was a 5.5 kW 'Q' type centrifugal, producing 28 ins W.G. at full load; in practice this restricted the bed depth to 200 mm. It was felt that had a variable speed Rootes type blower or a backward blade impeller type fan been available, a much deeper bed could have been used. A 300 mm bed depth would have encouraged better reten tion of the carbon and volatiles within the bed which should have given a higher overall efficiency.

FUEL DISTRIBUTION

Experience with a $3.5 m^2$ commercial f.b.c. fired by straw pellets fed on to one side of the bed had shown that due to rapid combustion, there was insufficient time for fuel to migrate across the full width of the bed. On the test rig it was found that the slotted conveyor tube which extended into the centre of the fluidised bed well below the surface allowed the char to pack, increasing the density and helping to retain the mass in the bed and thereby giving a little more time for fuel distribution. From the upper sight glass it could be clearly seen that combustion was evenly distributed over the entire bed.

Samples for flue gas analysis were taken from a tapping in the duct between the cyclone and the stack; they were passed through Bacharach FYRITE analysers for CO_2 and O_2 and DRAEGER tubes for CO. During interrupted fluidisation, the sampling time was equivalent to approximately ten cycles which was perfectly adequate for overall evaluation. To obtain instantaneous readings, a probe above the bed attached to a continuous on-line gas analyser would have been necessary, but this facility was not provided.

In the flue gas calculations for interrupted fluidisation shown in Appendix 2, the CO_2 and O_2 readings used were the lowest recorded in the series of readings. Had the highest figures of 12% CO_2 and 4% O_2 been used, the overall efficiency would have been greater.

During the trial, only one cam shape was used on the butterfly valve operating mechanism and this had been designed using airflow data collected by moving the valve by hand through 10° increments. The mechanism was also used at only one speed; in any future investigations, an in-depth study of different ratios and proportions of the interrupted cycle should be conducted. It would be possible to compare changes in efficiency at selected total gas volumes produced by several different patterns of cam. Uninterrupted and interrupted bed runs could also be compared as could effects of varying the rate of change between the two states.

It could well be that two seconds slumped in every five would reduce emissions but that excess air during the other three seconds would create high air velocity, way above those required for minimum fluidisation, raise CO emission and would tend to elutriate fine sand. If, on the other hand, the cycle was increased to eight seconds with the bed slumped for two seconds of it, no great increase in excess air would be required.

AIR DISTRIBUTION

In practice, the porous tile gave good air distribution, but the thermal shock it received when turning off the propane gas flame when the bed had reached start-up temperature, caused it to crack. After this happened three times, causing considerable delays and cost, it was replaced by the stainless steel stand pipes mounted on a stainless steel plate. This also gave even distribution and no further problems in this respect were experienced.

START UP

From the beginning, it was the intention to operate the test rig with only one fan and this caused problems with the burner. It was difficult to keep it alight whilst simultaneously operating the air slide to the plenum, the burner air valve and the propane gas regulator. The flame had to be constantly watched through the lower sight glass to make sure it was alight; an automatic flame detector device was fitted at one stage but it was found to be unreliable so was discarded on safety grounds.

FUEL FEEDER

Heat transfer through the feed tube together with back pressure from the fuel feeder slots near the edge of the bed started volatilisation in the feeder prior to entry into the fluidised bed; this caused dense fumes to be expelled from the open end of the tube. Even with this sealed, there was still a problem as there was no means of making the rotary valve smoke-tight without a complete re-design. As a temporary measure, a collecting hood was positioned above the valve and this was connected to the secondary air fan which had the effect of transferring most of the smoke into the combustion chamber. This measure was only partially successful and it had no detectable effect on the efficiency of combustion.

On reflection, it would have been better to use a more positive feed system than an auger on the test rig. One possibility would have been to install a hydraulic ram feeder and this would also have eliminated the need for a rotary seal which, on occasions, jammed when long lengths of straw became trapped between the vanes and the casing.

It is well known that difficulties are likely to be encountered when introducing fuels by the bottom feed system into a pressurised bed; an approach used by many commercial companies is to fit a number of lock-hoppers above the feed system but this is only suitable for high density fuels such as coal. Fibrous materials are always difficult to handle, even when there are no pressure changes involved and it was intended to carry out a series of trials on various ratios of coal and straw but this was not possible because of the smoke problems. However, extended trials were carried out on such mixtures on a 2.5 m^2 fluidised bed hot gas generator installed at Dengie Crop Driers on a commercial grass drying plant. Three trials were carried out at coal : straw (Table 21) ratios of 1.75 : 1; 1 : 1; and 1 : 1.75. The unit utilised recycle gas for bed attemperation and for the dilution of the drying gases to control inlet temperature, particularly when drying low moisture crops. The analysis of the recycle gas (Table 22) altered little irrespective of the fuel's composition, but an increase in the proportion of straw in the fuel was mirrored by the content of ash in the dried grass.

With the high volatile effect of the straw, the larger the proportion used, the greater the increase in HGG exhaust gas outlet temperature due to the volatiles burning above the bed; there was also a subsequent decrease in bed temperature. (Table 23).

TABLE 21

COAL/STRAW PELLET BURNING TRIAL ON 1.6 m^2 F.B.C HOT GAS GENERATOR

ULTIMATE ANALYSIS

		MANSFIELD W.S. COAL	PELLETED STRAW.
MOISTURE	%	10.0	7.9
ASH	%	6.3	8.8
VOLATILES	%	33.9	66.0
FIXED CARBON	%	49.8	17.3
GROSS CALORIFIC VALUE	kJ/kg	28.0	19.7

TABLE 22

COAL/STRAW PELLET BURNING TRIAL (4) ON COMMERCIAL
$2.5 m^2$ F.B.C. HOT GAS GENERATOR
AT MESSRS DENGIE CROP DRIERS LTD. - DENGIE, ESSEX.

COMPOSITION OF RECYCLED FLUE GAS AND ASH CONTENT
AT VARIOUS COAL/STRAW PELLET RATIOS.

COAL : STRAW	CO %	CO_2 %	O_2 %	ASH IN DRIED GRASS %
1 : 0	0	7.5	14	.8
1.75 : 1	0	6.5	14	8.5
1.15 : 1	0	6.25	15	9
1 : 1.75	0	6.5	15	10

TABLE 23

COAL / STRAW PELLET BURNING TRIAL (4) ON COMMERCIAL
2.5 m^2 F.B.C. HOT GAS GENERATOR
AT MESSRS DENGIE CROP DRIERS LTD - DENGIE, ESSEX.

		Ratio of Coal : Straw (c.e.)		
		1.75 : 1	1.15 : 1	1 : 1.75
FUEL MJ/h	28800	18372 : 10473	15407 : 13393	10473 : 18372
BED TEMPERATURE °C	850	850	840	835
FREEBOARD TEMP. °C	1000	1005	1040	1050

The lower the temperature needed for the drier inlet, the larger the quantity of recycle gas required for control and with the amount already needed for bed attemperation, this quantity was limited. In this installation, the fuel was fed above the bed and the burning volatiles raised the freeboard temperature to within 100°C of the ash fusion point. It was impossible to use secondary air with any of the straw-based fuels because of the uplift in temperature and this clearly demonstrated the necessity for the underbed feeding of straw fuels, so that volatiles would be burned in the bed.

Although Fluidised Bed Combustion has been around commercially for twenty years and many of the units perform quite satisfactorily, a realistic assessment would come to the conclusion that they are capable of still further developments. Large units are working well but require highly sophisticated control equipment. Smaller units, in particular experimental ones which tend to be small, are without such equipment and tend to be erratic in operation.

In biomass combustion and combustion of light waste products, there are particular problems and if one excludes the possibility of densifying the fuel by a processing stage there appear to be few ways in which elutriation can be kept to acceptable levels.

The way explained in this work, namely that of intermittent fluidisation appears to offer a reasonably satisfactory method of meeting the requirements and throughout the period this work has covered, no potentially better way has been suggested.

It would be interesting if further work could be done to compare interrupted fluidisation results of straw burning with those of various types of coal or other solid fuels; this would be fairly easy to do as the cam can quickly be altered or changed and a comparative table of cycle and bed slump times could be prepared.

CONCLUSIONS

1. Straw is a potential fuel in this country and is not being efficiently burned at the moment. In the most recent review of the position of biofuels in the overall energy strategy of the E.E.C., stress has been laid on the need for more efficient systems to be developed with low emissions for combustion of wood and straw. (COOMBS - 1986).

2. Because of its bulk, it seems sensible to use this fuel in rural areas where many of the heating loads are relatively small.

3. Pelleted straw can be burned satisfactorily on small fluidised bed units but not on large ones unless there are several feed points.

4. All indications are that chopped straw should be fired below the surface of the bed but problems developed with burn back and the smoke leaked past the straw wad which it had been hoped would have contained it.

5. It would have been better to use a ram feeder, but this would not necessarily be smoke tight, so the only way to get around this would be to install a collecting hood adjacent to the ram and duct to the intake of the fluidising fan thereby returning emissions back into the system.

6. Fluidised bed temperatures of 800°C are sufficient to maintain satisfactory combustion. High temperatures lead to ash fusion above the bed.

7. Unwise to use secondary air because of increased freeboard temperatures again leading to ash fusion above the bed.

8. From the limited amount of evidence it would seem that the concept of intermittent fluidisation is worth pursuing. Carryover was reduced and the percentage of carbon in the carryover was also reduced.

9. Increased freeboard cross sectional area contained emissions also and contributed to the relatively low flue gas outlet temperature.

10. It has been shown that porous ceramic tiles will not stand up to stresses incurred during bed heating and cooling unless unreasonably long times are allowed for these processes.

11. Combustion was sustained with chopped straw, the first time this has been achieved.

SUGGESTIONS FOR FURTHER INVESTIGATION

1. Interrupted fluidisation using different cam shapes to give a range of cycle times.

2. Repetition of 1 using different depths of fluidised bed and mean particle size of sand.

3. Replacement of screw feeder with pneumatic ram to eliminate as far as possible the emissions created by burn back.

4. Manometers fitted to gas and air lines to improve burner efficiency and establish flows.

5. Repeat all trials using coal or other waste derived fuels to establish comparisons with straw.

6. Repeat trials using high moisture content straw and other biofuels.

APPENDIX 1

CALCULATIONS FOR SIMPLE FLUIDISED BED COMBUSTOR

Figures Used in Calculations for FBC at 850°C

Silica Sand S.G. 2.65 Calculated in Text. Mean particle size	$d_p = 0.984^{-3}$ m
Particle Density	$p_s = 2650$ kgm³
Gravitational Acceleration	$g = 9.81$ m/s²

Density of Fluidising Gas at Bed Temp:

$$\frac{1\ \text{ATM}\ (\text{kN/m}^2)}{R \times {}^\circ K} = \frac{101.325}{0.2871 \times [273 + 850]} \qquad p_g = 0.314\ \text{kg/m}^3$$

Viscosity of Fluidising Gas (Air) at Bed Temp:

1 Centipose = 0.001 kg/m-s
From Viscosity Nomograph 850°C = 0.0445c.p. $\mu_g = 4.45^{-5}$ kg/m.s.

Thermal Conductivity of Fluidising Gas (Air) at Bed Temp. from Average Products of Combustion Standard Graph for Coal and Oil $k = 7.318^{-2}$ x

Density of Fluidising Gas at Atmos. Temp. (20°C):

$$\frac{1\ \text{ATM}\ (\text{kN/m}^2)}{R \times {}^\circ K} = \frac{101.325}{0.2871 \times [273 + 20]} \qquad p_a = 1.20\ \text{kg/m}^3$$

Discharge Coeffficient for Distributor Nozzle $c_d = 0.5$

Mass Flow of Fuel:

$$\frac{\text{Theor. Output FBC}}{\text{G.C.V. Fuel}} = \frac{0.1\ \text{MW}}{16200 \times 10^3} \qquad m_f = 0.00617\ \text{kg/s}$$

Enthalpy of Water Vapour at 850°C from Liquid at 20°C:

h_f from 20°C-100°C	= 419.1 - 83.9		
	= 335.2 kj/kg	= 80.05 kcal	
h_{fg} at 100°C	= 22567 kj/kg	= 538.90 kcal	
h_j from 100°C-850°C	= 0.497 x 750		
	= 372.75 kcal	372.75 kcal	h_g = 991.7 kcal

FLUIDISED BED COMBUSTOR TEST RIG

Requirement To operate at atmospheric pressure of 1 atmosphere, (= 1.013 bar) at 850°C and be fired with chopped straw

The combustion system is required to produce an output of 0.1 MW, and the bed area is to be 0.1265 m^2 and the bed depth 0.229 m.

FUEL ANALYSIS

C	39.9%
H	4.9%
O	33.4%
N	0.8%
S	0.1%
Ash	3.9%
H_2O	17.0%

ULTIMATE FUEL ANALYSIS

C	0.399
H	0.049
O	0.334
N	0.008
S	0.001
Ash	0.039
H_2O	0.170
	1.000

PROXIMATE FUEL ANALYSIS

Volatiles	0.430
Fixed Carbon	0.361
Total Moisture	0.170
Total Ash	0.039
	1.000

CALORIFIC VALUE

$$\text{Nett C.V.} = 81.37C + 345[H - \frac{(O+N)-1}{8}] + 22.2\,S$$

$$= 3246.66 + 258.75 + 2.22$$

$$= \underline{3507.63 \text{ kcal/kg}}$$

14689 KJ/kg

$$\text{Nett C.V.} = \text{Gross C.V.} - 586\left[\frac{\%H_2O + 9 \times \%H}{100}\right]$$

$$\text{N.C.V.} = \text{G.C.V.} - 358.046$$

$$\therefore \text{G.C.V.} = 3507.63 + 358.046$$

$$= \underline{3866 \text{ kcal/kg}}$$

16200 KJ/kg

$$\text{Fuel Rate} = \frac{\text{F.B.C. Theor.Output(0.1MW)}}{3866}$$

$$= \frac{85966}{3866}$$

$$= \underline{22.24 \text{ kg/w}}$$

COMPOSITION OF AIR

Used in these calculations

Dry Air = O - 0.2314

N - 0.7681

CO - 0.0005

1.000

$$\text{Specific Humidity} = w = \frac{\text{wt of water vapour}}{\text{wt of dry air containing the vapout}}$$

Molecular weights derived from table of International Atomic Weights

CALCULATION OF STOICHIOMETRIC AIR/FUEL RATIO

Oxygen Balance (per kg of fuel)

$$C + O_2 \dashrightarrow CO_2 = 0.399 \times \frac{32.000}{12.014} = 1.0630 \text{ kg/kg fuel}$$

$$4H + O_2 \dashrightarrow 2H_2O = 0.049 \times \frac{32.000}{4.036} = 0.3885 \text{ kg/kg fuel}$$

$$S + O_2 \dashrightarrow SO_2 = 0.001 \times \frac{32.000}{32.006} = 0.0010 \text{ kg/kg fuel}$$

Less oxygen supplied by fuel = -0.3340 kg/kg fuel

$\therefore$ Stoichiometric oxygen required = $\underline{1.1185 \text{ kg/kg fuel}}$

Stoichiometric Dry Air Requirement $\frac{1.1185}{0.2314}$ = 4.8336 kg/kg fuel

but at 20°C relative humidity 60% W = 0.009 kg vapour/kg dry air

$\therefore$ Stoichiometric as received air

= 4.8336 x 1.009

= 4.8771 kg/kg fuel

CALCULATION OF PRODUCTS OF COMBUSTION

The following assumptions are made:

(a) 100% combustion (giving maximum gas flow)

(b) because of unknown high volatile fuel reactions, maximum excess air quantity provided to be 50%

$\therefore$ Maximum Air flow rate = 1.5 x 4.8771	= 7.31565
Moisture content = $\frac{0.009}{1.009}$ x 7.3157	= 0.0653
$\therefore$ Dry Air content = 7.3157 - 0.0653	= 7.2504
CO_2 in air = 0.0005 x 7.2504	= 0.0036
O_2 in air = 0.2314 x 7.2504	= 1.6777
N_2 in air = 0.7681 x 7.2504	= 5.5690

CARBON BALANCE

Wt of CO_2 in air	= 0.0036
Wt of unburnt carbon	= 0
Wt of CO_2 = (0.399-0) x $\frac{44.011}{12.011}$	= 1.4620
TOTAL CO_2	= 1.4656

NITROGEN BALANCE

Nitrogen in Fuel	=	0.0080
Nitrogen in Air	=	5.5690
TOTAL N_2	=	5.5770

FREE OXYGEN

Wt of O_2 in CO_2	$= (0.399 - 0) \times \frac{32.000}{12.011}$	=	1.0630
Wt of O_2 in SO_2	$= 0.001 \times \frac{32.000}{32.066}$	=	0.0010
Wt of O_2 in H_2O	$= 0.049 \times \frac{32.000}{4.036}$	=	0.3885
Assume NOx = O and CO = O			1.4525

Free O_2 = O_2 in air + O_2 in fuel - O_2 used

$= 1.6771 + 0.3340 - 1.4525 \quad = \quad 0.5586$

MOISTURE BALANCE

H_2O in air		=	0.0653
H_2O in Fuel		=	0.1700
H_2O from hydrogen	$= 0.049 \times \frac{18.016}{2.016}$	=	0.4379
	TOTAL H_2O	=	0.6732

SULPHUR BALANCE

NO sulphur retention

$\therefore \; SO_2 = 0.001 \times \frac{64.066}{32.066} \quad = \quad 0.0020$

OVERALL MASS BALANCE

Input

Fuel	=	1.0000	kg
Humid Air	=	7.3157	kg
TOTAL	=	8.3157	kg

Output

CO_2	=	1.4656
N_2	=	5.5770
O_2	=	0.5586
H_2O	=	0.6732
SO_2	=	0.0020
NOx	=	0
CO	=	0
Ash	=	0.0390
TOTAL	=	8.3154

VOLUMETRIC ANALYSIS

Constituent	Mass kg/kg Fuel	Molecular Weight	No. of Mols = Mass / Mol. wt.	Vol of 1 kg Mol @ S.T.P. m^3/kgmol	Volume m^3/kg fuel	Volumetric Fraction = Partial Pressure	% by Volume
CO_2	1.4656	44.011	0.033301	22.41	0.746275	0.11638	11.638
SO_2	0.0020	64.066	0.000031	22.41	0.000695	0.00011	0.011
O_2	0.5592	32.000	0.017456	22.41	0.391189	0.06100	6.100
N_2	5.5770	28.168	0.197990	22.41	4.436956	0.69192	69.192
H_2O	0.6732	18.016	0.037367	22.41	0.837394	0.13059	13.059
TOTAL	8.2770		0.286145	22.41	6.412509	1.00000	100.000

GENERAL FLUE GAS DATA

(a) Density $= \frac{8.2764}{6.412509} = 1.2907$ kg/m^3

(b) Specific Volume $= \frac{1}{1.2907} = 0.7748$ m^3/kg

(c) Mean Molecular Wt. $= \frac{8.2764}{0.286164} = 28.924$

SUMMARY

	Gravimetric Analysis %	Volumetric Analysis %	Vol. Dry Analysis %
CO_2	17.708	11.638	13.386
SO_2	0.024	0.011	0.013
O_2	6.749	6.100	7.016
N_2	67.385	69.192	79.585
H_2O	8.134	13.059	0.000
TOTALS	100.000	100.000	100.000

BED VOIDAGE ε

Silica sand of mean particle size 0.984 mm and S.G. 2.65 was weighed for slumped bed 0.229 m deep. Bed depth at U_{mf} = 0.254 m deep.

Wt of bed = 46 kg

Vol. of fluidised bed = 0.032 m^3

$\therefore$ Wt. per m^3 = 1494 kg

Sand density = 2650 kg/m^3

$\therefore$ Voidage ε $= 1 - \frac{1496}{2650}$ = 0.435

PRESSURE LOSS

Through bed Δp_b $= (p_p - p_f)(1 - \ \ ) gh$

p_p = 2650 kgm^3 $= 2649.69 \times 0.565 \times 9.81 \times 0.229$

p_f = 0.314 kg/m^3

= 0.435 = 3363 Nm^3

g = 9.81 m/s^2 = 13.5 INS WG

h = 0.254 m

Through distributor Δp_d = 1.5mm WG per mm static depth (BOTTERILL)

= 1.5 x 228.6 = 343 mmWG = 13.5 INS WG

$\therefore \Delta p_b \equiv \Delta p_d$ Total = 27 INS WG

Δp_1 assume pressure drop through duct of 1 INS WG

Δp_1 and pressure drop through valve of i INS WG

(Actual measured loss by slcak manometer was 29 INS WG)

FAN POWER

$$p = \frac{M_a}{P_a}(\Delta p_b + \Delta p_d + \Delta p_l)$$

where $\dot{m}_f = 0.00617$ kg/s

$$\dot{m}_a = (1 + x_s)\ Sm_f = 1.5 \times 4.8771 \times 0.00617 = 0.045 \text{ kg/s}$$

$$\therefore\ p = \frac{0.045}{1.20} \times (29 \times 240.08) = 2.61 \text{ KW}$$

but centrifugal fan is $\simeq$ 50% efficient

$$\therefore\ \text{MOTOR POWER} = \frac{2.61}{0.50} = 5.22 \text{ KW}$$

MOTOR POWER ON FAN QP 6110 = 5.50 KW

DISTRIBUTOR DESIGN

$$\frac{\dot{m}_a}{p_a} = C_d\, A\sqrt{\frac{2\ p_d}{p_a}}$$

$$\therefore\quad A = \frac{\dot{m}_a}{p_a} \times C_d\sqrt{\frac{p_a}{2 \times p_d}}$$

$$A = \frac{0.045}{1.20} \times .5 \times 0.0134$$

$$= 2.5 \text{ m}^2$$

Number of 2mm diameter holed $= A = n \frac{\Pi}{4} d_o^2$

$= \underline{80}$

HEAT LOSS THROUGH REFRACTORY

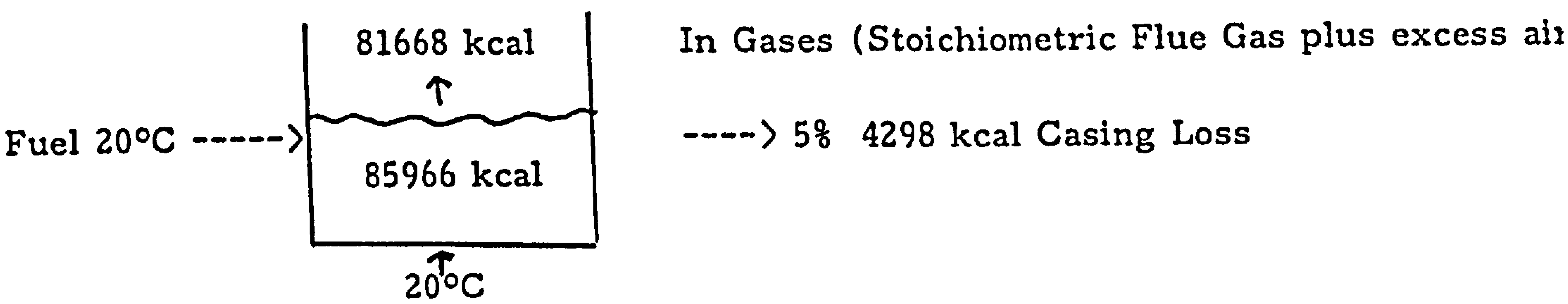
$$Q = k_{refrac} \times \text{Bed Area} \times \text{Temp gradient}$$

$$= 1.4 \times 0.71 \times 0.254 \times \frac{850}{50} \times 1000 \text{ W}$$

$$= 4292 \text{ W}$$

Loss through Refractory $= \frac{4292}{100,000} \times 100 = 4.29\%$ say 5%

TOTAL HEAT LOSSES



$CO_2 + SO_2$ $(1.4656 + 0.002) \times 22.24 \times 0.226 \times (850 - 20)$

$= 6123$ kcal/hr

N_2 $5.5770 \times 22.24 \times 0.264 \times (850 - 20)$

$= 27178$ kcal/hr

H_2O $0.6732 \times 22.24 \times$ (Vapour 850 - Liquid 20)

14.972×991.7

$= 14848$ kcal/hr

Ash $0.039 \times 22.24 \times 0.2 \times (850 - 20)$

$= 144$ kcal/hr

TOTAL

$CO_2 + SO_2$	6123	kcal/hr
N_2	27178	kcal/hr
H_2O	14848	kcal/hr
Ash	144	kcal/hr
	48293	kcal/hr

Heat content of stoichiometric flue gas plus ash leaving bed at 850°C = 48293 kcal/hr

We have to remove from bed 81668 - 48293 kcal/hr which leaves a surplus of 33375 kcal/hr.

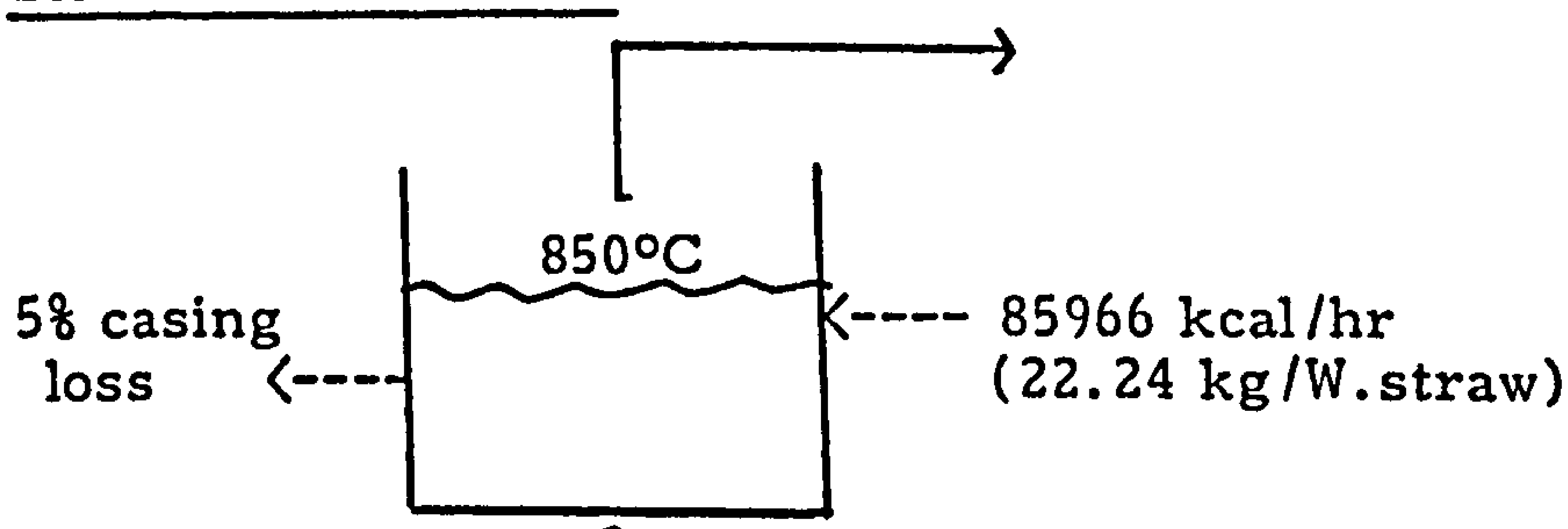
To calculate excess air to absorb surplus heat let x = kg/W of excess air
(SP Heat of air at 850°C=0.257)

$$\therefore \; x = \frac{33375}{0.257 \times (850-20)} = 156.46 \text{ kg/W air}$$

Actual stoichiometric air = 4.8771 x 22.24 = 108.33 kg/W

$$\therefore \; X_s \text{ air} = \frac{156.46}{108.33} = 144.43\%$$

THERMAL BALANCE



INPUT = Air plus straw

CO_2	32.59 kg
N_2	124.03 "
O_2	12.42 "
H_2O	14.97 "
SO_2	0.04 "
$X_s$air	156.46 "
Ash	0.87 "
	341.38 "

156.46 + (7.3156 x 22.24) + 22.24 = 341.40 kg

Note Above calculations are correct in theory but may not be correct in practice. It assumes that the 22.24 kg of straw releases all its 85966 kcal/hr of heat to the bed. In fact, with a fuel of such high volatile content, we can expect that a significant proportion of the heat will go to the freeboard.

Assume 30% of heat is lost to freeboard
plus 5% of heat is lost through casing

$\therefore$ heat to bed = 0.65 x 85966 = 55878 kcal/W

Now the products of combustion (stoichiometric) removes 48202 kcal/W

$\therefore$ the new surplus = 55878 - 48202 = 7695 kcal/W

$$\text{and } X_s \text{ air} = \frac{7695}{0.257 \times (850-20)} = 36.00 \text{ kg/W}$$

$$= \frac{36.08}{4.8771 \times 22.24} = 33.2\%$$

INGRESS AIR

However, it is theoretically possible that air within the straw mass will be fed into the FBC.

Conveying volume of screw/W = area x pitch x RPM x 60

$= 3.53^{-3} m^2 \times 0.07 \times 26 \times 60$ $= \underline{0.385\ m^3}$

Weight of compacted straw m^3 (see INTRODUCTION) $\underline{66.73\ kg}$

Weight of chopped straw as fed = $22.24 \div 0.385\ m^3$ $\underline{57.77\ kg}$

$\therefore$ Wt of air as fed = 66.73×0.385
-57.77×0.385 $= \underline{3.45\ kg}$

$\therefore$ New X_s air = $36.08 - 3.45$ $= \underline{32.63\ kg/hr}$

$$\frac{32.63}{4.8771 \times 22.24} \times 100 = \underline{30.08\%}$$

A P P E N D I X 2

CALCULATIONS FOR TRIALS ON SIMPLE FLUIDISED BED COMBUSTOR

355 mm x 355 mm

Figures and data used in calculations for FBC at 800°C.

Weight of straw per hour	22 kg
Gross Calorific Value	16200 MJ/kg

Ultimate analysis of fuel by mass

		%
M	Moisture	17.0
A	Ash	3.9
C	Carbon	39.9
H	Hydrogen	4.9
N	Nitrogen	0.8
S	Sulphur	0.1
O	Oxygen	33.4
		100.0

Volumetric analysis of flue gas (CO_2 & O_2 by fireite, CO by draeger tube & N_2 by difference

	Uninterrupted Fluidisation %	Interrupted Fluidisation %
CO_2 carbon dioxide	8.0	10.0
O_2 oxygen	4.0	2.0
CO carbon monoxide	0.1	0.1
N_2 nitrogen	89.9	87.9
	100.0	100.0

Ultimate analysis of cyclone ash

	Uninterrupted Fluidisation %	Interrrupted Fluidisation %
Volatile Matter	6.27	5.50
Ash	85.80	87.95
Fixed Carbon	6.08	5.10
Total Moisture	1.85	1.45
	100.00	100.00

Assume 80% of volatiles in carbon ∴ total carbon becomes	0.111 kg/h	0.095 kg/h
Weighed cyclone catch	0.622 kg/h	0.548 kg/h
Assume cyclone efficiency is 90% ∴ total loss becomes	0.684 kg/h	0.603 kg/h
Thermal value of cyclone ash		
Gross calorific value	3310 kJ/kg	2950 kJ/kg
Net calorific value	3233 kJ/kg	2881 kJ/kg

Temperature of inlet air ta = 25°C (298°K)

Flue gas temperature tg = 280°C (553°K)

Carbon Efficiency

$$\text{C.E.} = \frac{\text{Wt C in fuel - Wt C in cyclone ash}}{\text{Wt C in fuel}} \times 100 \quad \%$$

Uninterrupted fluidisation

$$\text{C.E.} = \frac{[22\text{x}0.399]-[0.684\text{x}0.111]}{22 \text{ x } 0.399} \text{ x } 100$$

$$= \underline{99.14\%}$$

Interrupted fluidisation

$$\text{C.E.} = \frac{[22\text{x}0.399]-[0.603\text{x}0.095]}{22 \text{ x } 0.399} \text{ x } 100$$

$$= \underline{99.35\%}$$

Combustion Efficiency

$$\text{Com.E.} = \frac{\text{Heat content of fuel - C loss}}{\text{Heat content of fuel}} \times 100 \quad \%$$

Uninterrupted fluidisation

$$\text{Com.E.} = \frac{[22\text{x}16200]-[0.684\text{x}0.111\text{x}32792]}{22 \text{ x } 16200} \text{x}100$$

$$= \underline{99.30\%}$$

Interrupted fluidisation

$$\text{Com.E} = \frac{[22\text{x}16200]-[0.603\text{x}0.095\text{x}32792]}{22 \text{ x } 16200} \text{x}100$$

$$= \underline{99.47\%}$$

Efficiency of F.B.C.

The losses method

F.B.C. efficiency can be obtained by subtracting the sum of the following percentage losses from 100.

L_1 Loss due to moisture in fuel

L_2 Loss due to water in the flue gas resulting from hydrogen in fuel

L_3 Loss due to elevated temperature of the dry flue gas

L_4 Loss due to incomplete combustion of the fuel resulting in carbon monoxide in the flue gas

L_5 Loss due to incomplete combustion of the fuel resulting in carbon in the refuse

L_6 Loss due to elevated temperature of the refuse

L_7 Loss due to radiation and other unaccounted for heat transfer

Losses

$$L_1 = M\left[\frac{2.0\left(\frac{tg}{k} - 100\right) + 4.2\left(100 - \frac{tg}{k}\right)}{G.C.V.}\right]$$

$$= 17\left[\frac{2.0\left(\frac{280+273}{273} - 100\right) + 2260 + 4.2\left(100 - \frac{25+273}{273}\right)}{16200}\right]$$

$$= \underline{2.6\%}$$

$$L_2 = 9 \times H\left[\frac{2.0\left(\frac{tg}{k} - 100\right) + 2260 + 4.2\left(100 - \frac{tg}{k}\right)}{G.C.V.}\right]$$

$$= 9 \times 4.9\left[\frac{2.0\left(\frac{280+273}{273} - 100\right) + 2260 + 4.2\left(100 - \frac{25+273}{273}\right)}{16200}\right]$$

$$= \underline{6.75\%}$$

$$L_3 = \frac{Wd(tg - ta)Cp}{G.C.V.} \times 100$$

where $$Wd = \frac{(11xCO_2)+(8xO_2)+7(N_2+CO)}{3(CO_2 + CO) \times 100} \times C - \frac{UCXA}{100-UC} + \frac{3 \times S}{8}$$

Uninterrupted fluidisation

$$Wd = \frac{(11x8)+(8x4)+7(87.9+0.1)}{3(8+0.1) \times 100} \times 39.9 - \frac{(11.1x3.9)}{100 - 11.1} + \frac{(3x0.1)}{8}$$

$$= 11.926$$

$$\therefore L_3 = \frac{11.926(553-298) \times 1.032 \times 100}{16200}$$

$$= \underline{19.38\%}$$

$$L_4 = \frac{0.1\ CO}{(8CO_2+0.1CO)x100}\ \frac{39.9C-(11.1x3.9)}{100-11.1} \times \frac{23600 \times 100}{16200}$$

$$= \underline{0.71\%}$$

$$L_5 = \frac{A \times UC}{100(100-UC)} \times \frac{32800x100}{GCV}$$

$$= \frac{3.9 \times 11.1}{100x88.9} \times \frac{3280000}{16200}$$

$$= \underline{0.99\%}$$

Interrupted fluidisation

$$Wd = \frac{(11x10)+(8x2)+7(87.9+0.1)}{3(10+0.1) \times 100} \times 39.9 - \frac{(9.5x3.9)}{100 - 9.5} + \frac{(3x0.1)}{8}$$

$$= 9.661$$

$$\frac{9.661(553-298) \times 1.031 \times 100}{16200}$$

$$= \underline{15.68\%}$$

$$\frac{0.1\ CO}{(10CO_2+0.1CO)}\ \frac{39.9C-(9.5x3.9)}{100-9.5} \times \frac{23600 \times 100}{16200}$$

$$= \underline{0.57\%}$$

$$\frac{A \times UC}{100(100-UC} \times \frac{32800 \times 100}{GCV}$$

$$= \frac{3.9x9.5}{100x90.5} \times \frac{3280000}{16200}$$

$$= \underline{0.83\%}$$

$L_6 = \frac{A(tg-ta)x0.84x100}{100 \times GCV}$	ditto
$= \frac{3.9(553-298)x0.84x100}{100 \times 16200}$	
$= \underline{0.05\%}$	$= \underline{0.05\%}$
$L_7 = \underline{5.0\%}$	$=\underline{5.0\%}$

From refractory losses calculation in Appendix 1

Total losses 35.48%	31.48%
Efficiency 100 - losses	
$= \underline{64.52\%}$	$= \underline{68.52\%}$

Flue gas specific heat Cp for uninterrupted fluidisation

	mol. wt	%	total wt	%	Cp Mean	Actual Cp
CO_2	44.011	8.0	352.088	11.90	0.955	0.1136
O_2	32.000	4.0	128.000	4.33	0.951	0.0412
N_2	28.168	87.9	2475.967	83.68	1.047	0.8761
CO	28.000	0.1	2.800	0.09	1.055	0.0009
		100.0	2958.855	100.00		1.0318

Flue gas specific heat Cp for interrupted fluidisation

	mol. wt	%	total wt	%	Cp Mean	Actual Cp
CO_2	44.011	10.0	440.110	14.75	0.955	0.1409
O_2	32.000	2.0	64.000	2.15	0.951	0.0204
N_2	28.168	87.9	2475.967	83.01	1.047	0.8691
CO	28.000	0.1	2.800	0.09	1.055	0.0009
		100.0	2982.877	100.00		1.0313

The enthalpy of combustion of carbon monoxide per mass of carbon monoxide is taken as 23600 KJ/kg

The enthalpy of combustion of carbon per mass of carbon is taken as 32800 KJ/kg

The mean isobaric specific heat capacity of the refuse is taken as 0.84 KJ/kg

(From selected principles and methods of Technical Calculations)

A = ash % infueld as per ultimate analysis
UC = carbon in dust as % of dust
Wd = mass of dry gas per mass of fuel

APPENDIX 3

EXTRACT FROM

FLUIDISED COMBUSTION OF CHAR AND VOLATILES

FOR COAL

BY

TURNBULL, E.

Unpublished PhD Thesis - Trinity College

Cambridge

BURNOUT OF STRAW WAFERS IN A FLUIDIZED BED COMBUSTOR

Experimental Apparatus : A 100 mm diameter fluidized bed combustor was used to obtain the CO and CO_2 traces, and a 75 mm diameter silica fluidized bed used for visual observation.

Results : The experimental conditions were :

Temperature	:	850°C
Air Velocity	:	100 mm bed : 1.50 m/s at 850°C
		75 mm bed : 0.30 m/s at 850°C
Wafer diameters:		traces 1,2,3 : 25 mm
		traces 4,5,6 : 52 mm
Wafer densities:		200-600 kg/m^3

Two types of behaviour were identified :

(i) The low density wafers displayed two stage combustion. During the first stage :

(a) the wafer remained on top of the bed,
(b) devolatilization occurred,
(c) the wafer expanded considerably - therefore increasing the surface area available for reaction

As a result, a long diffusion flame appeared above the bed.

The second stage commenced when the wafer had broken up into many small pieces and combustion of the solid carbon commenced. The small pieces were distributed throughout the bed and combustion occurred in the bed.

This type of behaviour is characterised by :

(a) production of large amounts of carbon monoxide - which could represent a significant energy loss.

See traces 1,2,4,5.

(b) high elutriation rates in the initial stages of combustion.

(ii) The most dense wafers basically remained intact.

Initially the wafers remained on top of the bed, but because the centre was not exposed to oxygen the rate of combustion was much slower than for the previous cases - hence less CO is produced (see traces 3,6).

Eventually the wafers broke into two pieces (see humps in curves) and these stayed intact. The smaller pieces were distributed and burnt throughout the bed - they looked very much like high volatile coal particles.

The points to be noted are :

(i) The initial results are very promising - the rates of combustion compare very favourably with those of coal.

(ii) In larger beds the wafers may not initially remain on top of the bed.

(iii) The more dense wafers may be better suited to combustion in a fluidized bed because

(a) there is less overbed burning

and

(b) there is less elutriation.

However, if (ii) is true then combustion of the less dense wafers may be just as efficient.


15% CO2
2.9% CO
5% CO2
CO2
TRACE 3 : 26* : 25mm
15% CO2
2.9% CO
5% CO2
CO
CO2
TRACE 2 : 33*
25mm
15% CO2
2.9% CO
5% CO2
CO
CO2
TRACE 1 : 38* : 25mm


TRACE 5 : 17* 54mm
TRACE 4 : 19* : 54mm
15% CO2
2.9% CO
5% CO2


TRACE 6 : 15* : 54mm
CO2
CO
15% CO2
2.9% CO
5% CO2
TRACE 5 : 17*
54mm
CO
CO2

APPENDIX 4

THE DIVIDED BED (MCV) GASIFICATION PROCESS

E.E.C. ENERGY DEMONSTRATION PROJECT

BY

J. F. WASHBOURNE

ENERGY CONFERENCE - BRUSSELS, JUNE 1984

INTRODUCTION.

During the 1960's, most Energy Planners in the U.K. believed that the future of power generation lay in the efficient use of low grade coals with high or variable ash contents.

It was recognised that fluidised combustion was the answer to most of the problems inherent with this type of fuel, and in 1979 the Energy Equipment Company Limited built a test facility at Olney in Buckinghamshire.

Originally, a 10.55 GJ/hr (10 x 10^6 Btu/hr) 'Energy' designed Hot Gas Generator was installed, whereby totally combusted gas between 1200 - 1400°C could be produced.

It was known that gas of this type could be used for direct and indirect drying and heat applications and, using the control parameters, the hot gas production would follow precisely the load demanded by the process.

Originally, difficulties were found when trying to control the bed temperature at a safe level (950°C) solely by regulating the air quantity passed to it. This lead to the Energy Equipment Company Limited's patented technique being evolved whereby cool waste gas was recycled through the bed restricting the oxygen content of the fluidising gas and allowing only partial combustion of the fuel. Secondary air was added above the bed to complete combustion.

Further, by the use of this recycled waste gas, control of the inlet temperature to the process could also be automatically maintained. (Figure 1)

The mechanical design of the Energy Equipment's fluidised bed low calorific value gasifier is similar to the 'Energy' hot gas generator in which fuel is fluidised by air through sparge pipes in a bed of sand, but secondary air is not introduced above the bed as we do not wish to complete combustion. (Figure 2)

By using a mixture of air and highly superheated steam, with a proportion of the made gas recycled through the sparge pipes, a combustible gas, having virtually no vaporised tar content, is available with a low calorific value of 4.6 to 5.0 MJ/m^3 (123 to 134 Btu/ft^3). Being hot at about 950°C, it can be cooled to any temperature suitable for the end use.

It was realised that the plant would have to be self-reliant in respect of the high temperature steam requirement and so a waste heat boiler and superheater were built into the system.

In 1982, the test facility was rebuilt and a new 15.8 GJ/hr (15×10^6 Btu/hr) unit was installed. This was designed so that it can operate either as an H.G.G. or as an L.C.V. gasifier.

Because the Energy Equipment's F.B.C. process can accept a wide range of fuels such as lignites, bituminous coals, anthracites, poor quality high ash fuels, biomass and waste derived fuels (W.D.F.), the new facility is ideal for the testing and analysis of potential customer's fuels. In fact, it is a much more versatile system than the town gas and producer gas processes which required quality fuels because they both were susceptible to variations in fuel characteristics. (Figures 3 and 4)

The atmosphere in an Energy Equipment bed is a reducing one where carbon is converted to CO rather than CO2, and sulphur to H2S rather than SO2. Also, as final combustion is completed at a later stage in the process, this leads, in practice, to a considerable proportion of the sulphur in the fuel being retained in the ash in the bed.

The relatively low temperature at which our F.B.C. operates ensures that the emissions of oxides of nitrogen are reduced to a very low level, less than 200 ppm by volume. Again, in practice, we find that in most cases it is lower than would be expected in theory, simply from the nitrogen content of the fuel. Some nitrogen is retained in the ash.

THE MEDIUM CALORIFIC VALUE (M.C.V.) GAS PRODUCER.

The M.C.V. demonstration project was started effectively in 1983 with financial support from the C.E.C.

The M.C.V. gasifier itself has been developed from the Energy Equipment system where partial combustion takes place in one area of the bed and where gasification takes place in another due to the addition of only steam and recycle gas to form a synthesis gas.

The plant gasifies solid fuel to generate a combustible gas at a nominal rate of 26.4 GJ/hr (25×10^6 Btu/hr). The gas has a calorific value in the range 9 to 13 MJ/Nm^3 (241 to 349 Btu/s ft^3) and temperature in the range of 300 - 450°C.

The plant consists of the coal handling, ash removal, ash handling, steam generation, water treatment, fuel oil, compressed air, thermal oxidiser systems and the gasifiers. (Figure 5)

Fuel is elevated from the feed hopper by a belt conveyor passing over a belt weigher and fed into the gas generator by a fuel feeder lock, and screw feeder system. The hopper lock utilises a double valve system to seal against the back pressure of gas in the gas generator. This consists of three chambers with gas proof isolating valves and level probes, all mounted within a spring loaded frame for ease of fuel flow. The unit operates sequentially and vibrators are incorporated on each chamber to avoid 'ratholing', arching or other interference with the fuel flow to the screw feeders.

The fuel is fed into a sand bed which is fluidised by air, steam and recycle gas, through separate pipes. When lighting up the plant, the air is preheated by the direct combustion of light oil, which is achieved by a high intensity 3.1 GJ/hr (3 x 10 Btu/hr) burner utilising 35 second Redwood No 1 oil, which is atomised by 0.7 bar compressed air from a variable speed blower.

Final dilution to achieve a combusted temperature of 950°C consists of mixing primary and tertiary air.

The area above the bed is divided into three separate zones - two combustion and one gasification. These zones are mirrored under the bed by the distribution system which is also split into three zones. (Figures 6 and 7)

Superheated steam enters the gasification zone of sand and fuel by sparge pipes, and recycle gas can also be added by a recycle gas fan in order to maintain fluidisation at low fuel rates. (Figure 8)

Steam can also be added to the combustion zones to control the zones bed temperatures. The zones are divided above the bed by a membrane water wall boiler.

From the combustion zones of the gasifier, the flue gas passes through the flue gas superheater and waste heat boiler which cools it to between 300 and 450°C and generates process steam at nominally 17 bar. Any excess steam is vented to atmosphere. This boiler also has an oil burner to raise steam for start-up purposes.

Ash is carried with the flue gas stream through the superheater and boiler. After the boiler, the flue gas passes to a multi-cyclone system which removes the ash. The ash is dropped into a skip for disposal by road transport.

After the ash removal, the flue gas is discharged to atmosphere via a stack.

The upper section of the gasifier has a reaction/disengagement zone to allow the entrained sand or large unburned fuel particles to fall back to the bed.

From the gasification zone of the gasifier, the product gas passes through a superheater and waste heat boiler, which cools it to between 300 and 450°C which generates process steam at nominally 17 bar, 450°C.

Ash is carried with the product gas stream through the product gas superheater and boiler, through a smoke box to a twin cyclone system which removes the ash. This is transferred by pneumatic conveying to a storage hopper from where it is discharged into a skip for disposal by road transport.

The combustible gases are burned in a test furnace before being discharged to atmosphere via a stack.

A major advantage of the fluid bed system is the elimination of the ash clinker removal and handling associated with fixed grates. However, with very high ash content materials, and those that contain slatey shale, ash is extracted from below the bed. The extracted material can be screened and the sand recycled to the bed. The separated ash and shale is disposed of by road transport.

The extracted material is screened and the carbon particles and sand are returned to the bed via the fuel feeding system.

TECHNICAL DESCRIPTION.

The Energy Equipment's divided fluidised bed gasifier generates a nitrogen-free synthesis gas from solid hydrocarbons using air rather than oxygen. Present medium calorific value (M.C.V.) gasifiers are limited to size and type of fuel and almost all require industrially pure oxygen.

Those others which are intended to operate without oxygen use some kind of pumped recirculation of the hot bed material from the heat producing (exothermic) part of the plant which absorbs heat (endothermic).

The Energy Equipment system has a common fluidised bed dynamic base for both sections, and heat is carried by the transverse progression of flüidised particles (transmigration) through the bed. (Figure 9)

The product gas and waste gas streams are separated by the membrane walls above the bed dividing the chambers, whilst the expanded dynamic fluidised bed provides the seal for the divided chambers, incorporating steam within the bed for the lower divided curtain.

The energy to move the particles is supplied solely from the dynamic energy of the fluidising gases, air and process steam. There is no mechanical handling or pumping of hot bed material. (Figure 10).

IN-BED GAS DIFFUSION.

Diffusion within the bed has to be considered on the following basis:-

(1) Diffusion of gases is limited and, with lower bed velocities, i.e. 0.5 m/s (based on standard temperatures and pressure), diffusion within the bed is such that there is likely to be a 30% slip of reactive gases out of the bed into the splash zone.

(2) When the bed operating conditions approach full load, the gas velocities within the bed approach 1 m/s perhaps rising to 1.3 m/s (related to standard conditions of temperature and

pressure) then diffusion is greater and the slip of reactive gases into the splash zone is likely to reduce to less than 10%.

However, diffusion of gases has a very limited effect upon the transfer of heat across the bed as this relates more particularly to the transmigration of particles of solids that constitute the bed.

Transmigration of solids within the bed is found to be effectively homogeneous, and the mechanics of this transmigration relates to the active columns of mixed gases and solids that operate in the bed during the boiling fluidisation.

DIFFERENTIAL PRESSURE.

Arrangements are made to balance the back pressure of the two gas streams since waste gas pressure in the surrounding second-phase chamber is controlled by the induced draught fan. To promote a bias for flow towards the endothermic section of the bed, we control at a slightly higher pressure in the waste gas area, i.e. 1.2 mbar higher than the pressure in the make gas second-phase gas chamber.

There are slight variations of pressure (up to 0.6 mbar) but these are not a problem as an active steam curtain within the bed separates the zones, approximately 0.25 m wide, where the prevailing dynamic effect will be for the gas to surface at a velocity of 1 m/s standard.

When considering the true velocities in the bed and out of the bed, the gas velocities must be multiplied by a temperature factor which, dependent on the bed temperature, will always be a factor of 3 or above.

It is important to understand that the common fluid bed containment vessel below the two zones has no supplementary means of bed material pumping.

ILLUSTRATIONS.

18. Photo - this is an outside view of the MCV unit showing the main body with the inspection door open to the right hand side exothermic chamber.

Below can be seen the main bed extractor screw.

On the left of the photo can be seen the recycle gas fan inside its sound proof housing.

19. Photo - This is a view inside the endothermic chamber showing both dividing membrane walls and below them the sparge pipes through which superheated steam enters the sand bed.

20. Photo - this is another view of the MCV unit taken from the roof of the recycle gas fan sound proof chamber.

21. Photo - an outside view showing the flue gas waste heat boiler and multi-cyclone.

22. Photo - another outside view showing the oil and water tanks.

23. Photo - this is an outside view of the test furnace installation and exhaust stack.

TYPICAL FUEL ANALYSIS: (WASHED SMALLS.)

AS FIRED

	%
Fixed Carbon	47
Volatile Matter	31
Ash	10
Total H_2O	12
	100

Combustible Sulphur 1·4

Gross Calorific Value 25·7 MJ/Kg (11050 Btu/lb)

Ash initial deformation temperature, > 1100 °C.

OTHER POTENTIAL FUELS

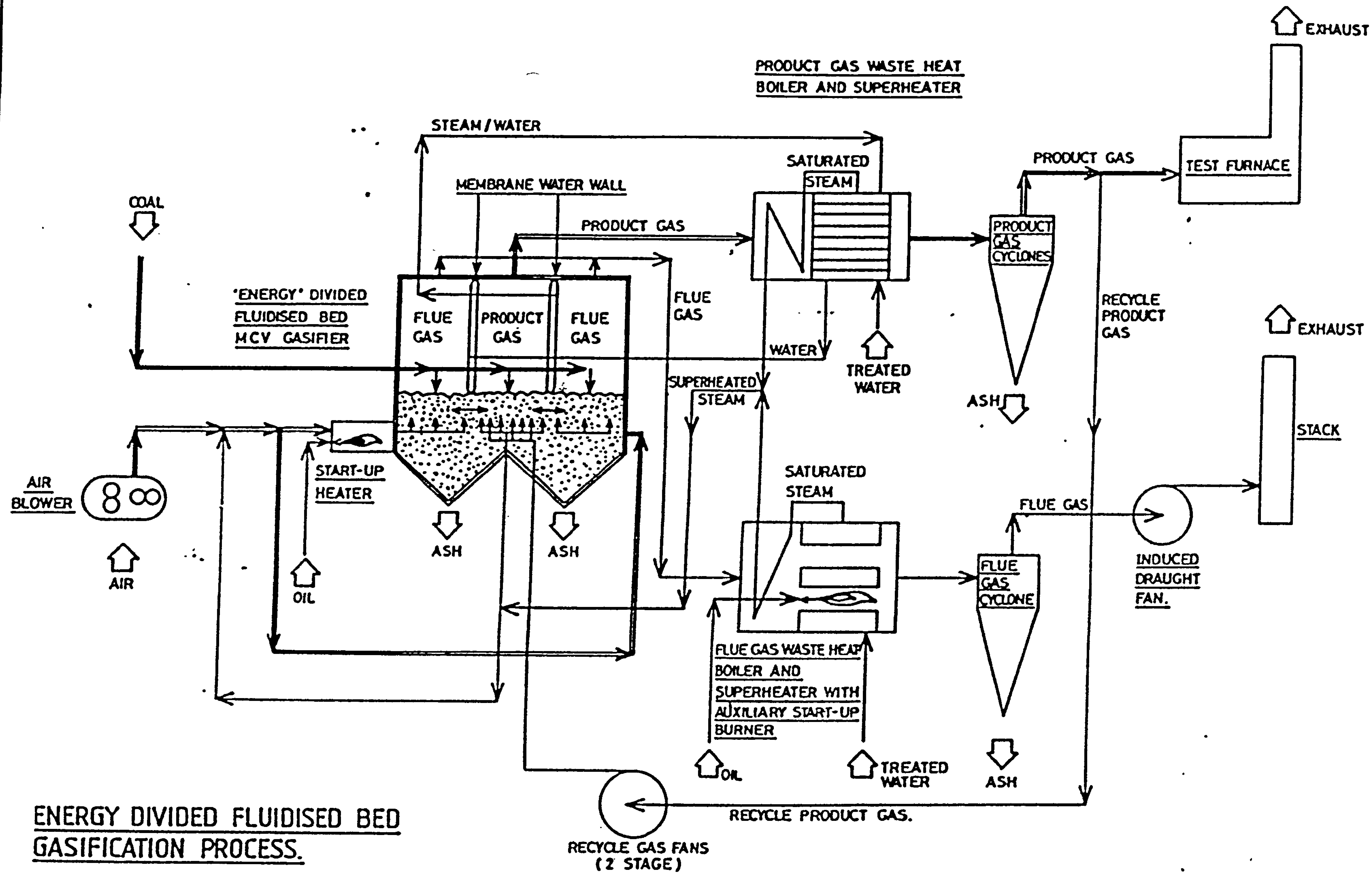
Lignite

Brown coal

High ash and fine particle size fuels: tailings, duff and shale fuels.

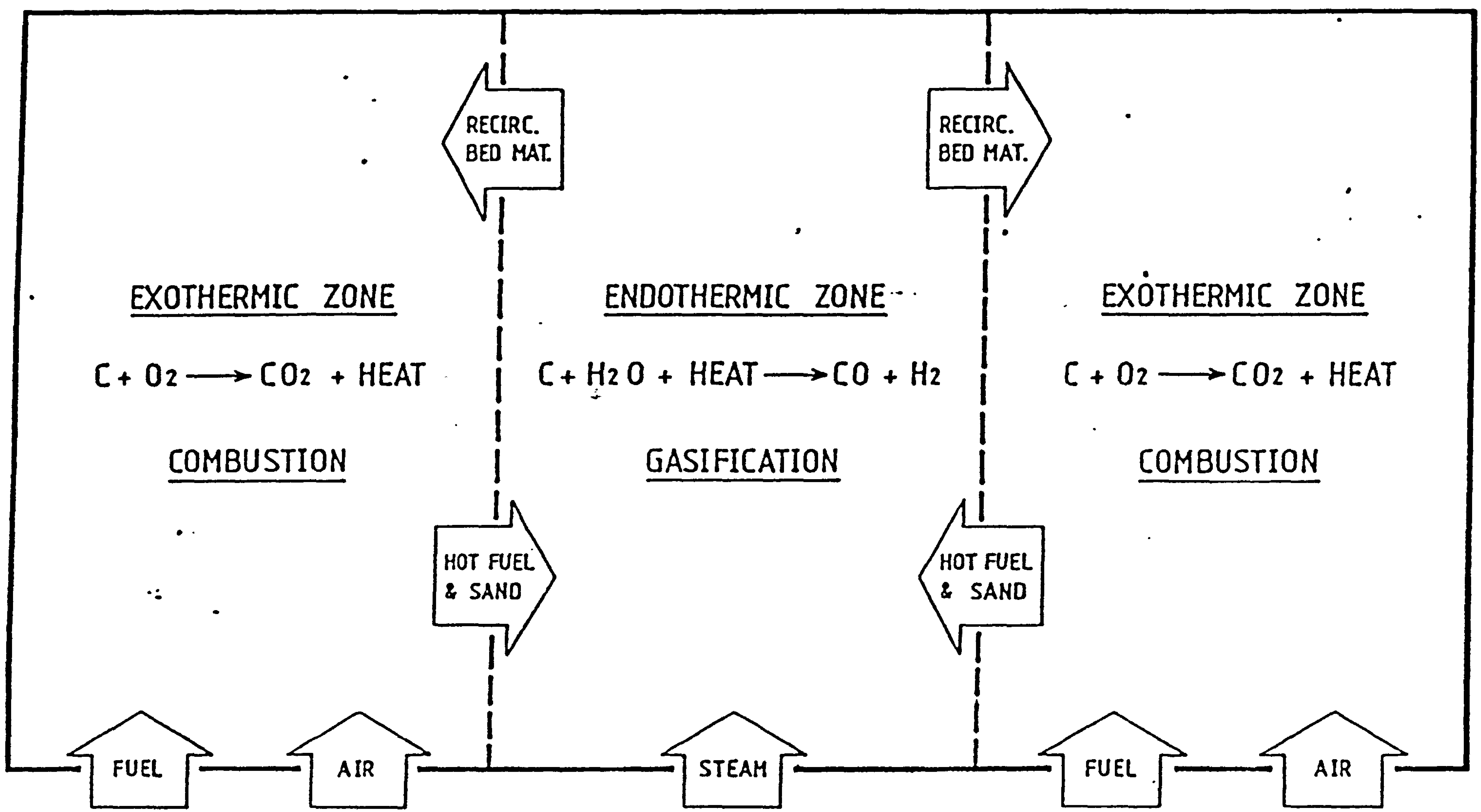
Wood chips, peat, coffee ground and other biomass fuels.

Municipal waste - derived fuel (plastic - free)

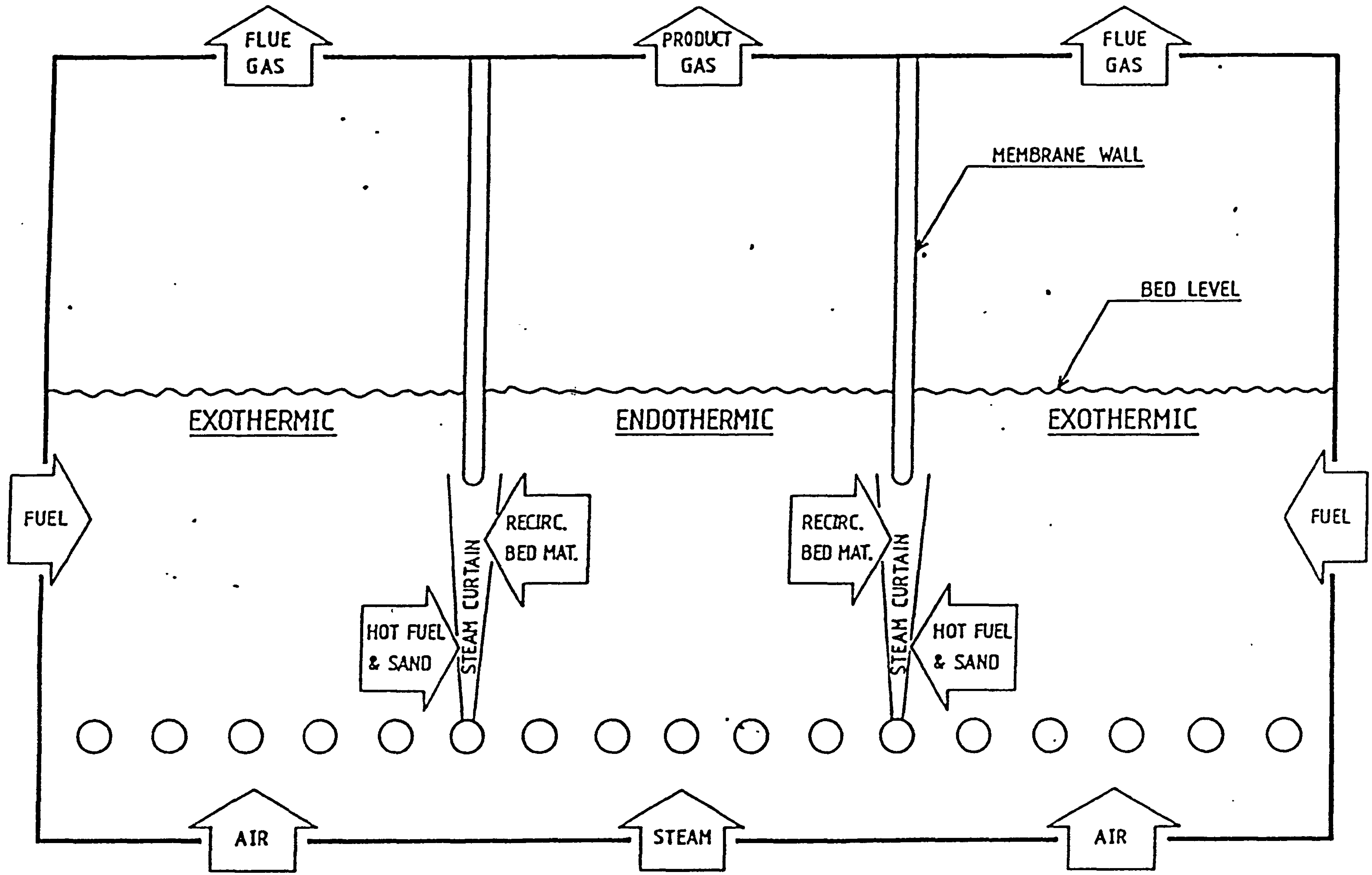


ENERGY DIVIDED FLUIDISED BED
GASIFICATION PROCESS.

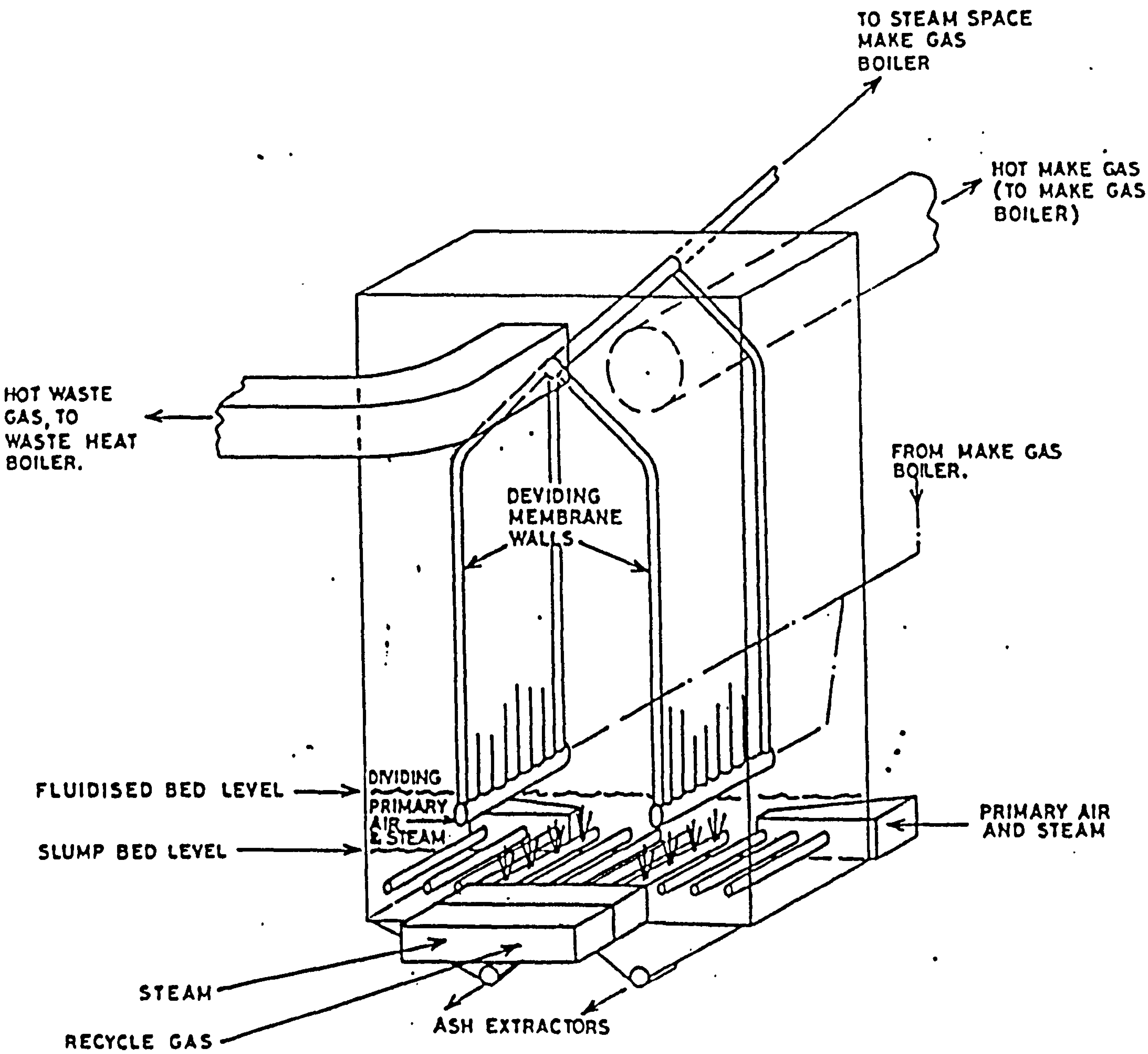
FLOW DIAGRAM.



SIMPLIFIED SCHEMATIC OF 'ENERGY' DIVIDED BED M.C.V. GASIFIER — PLAN VIEW

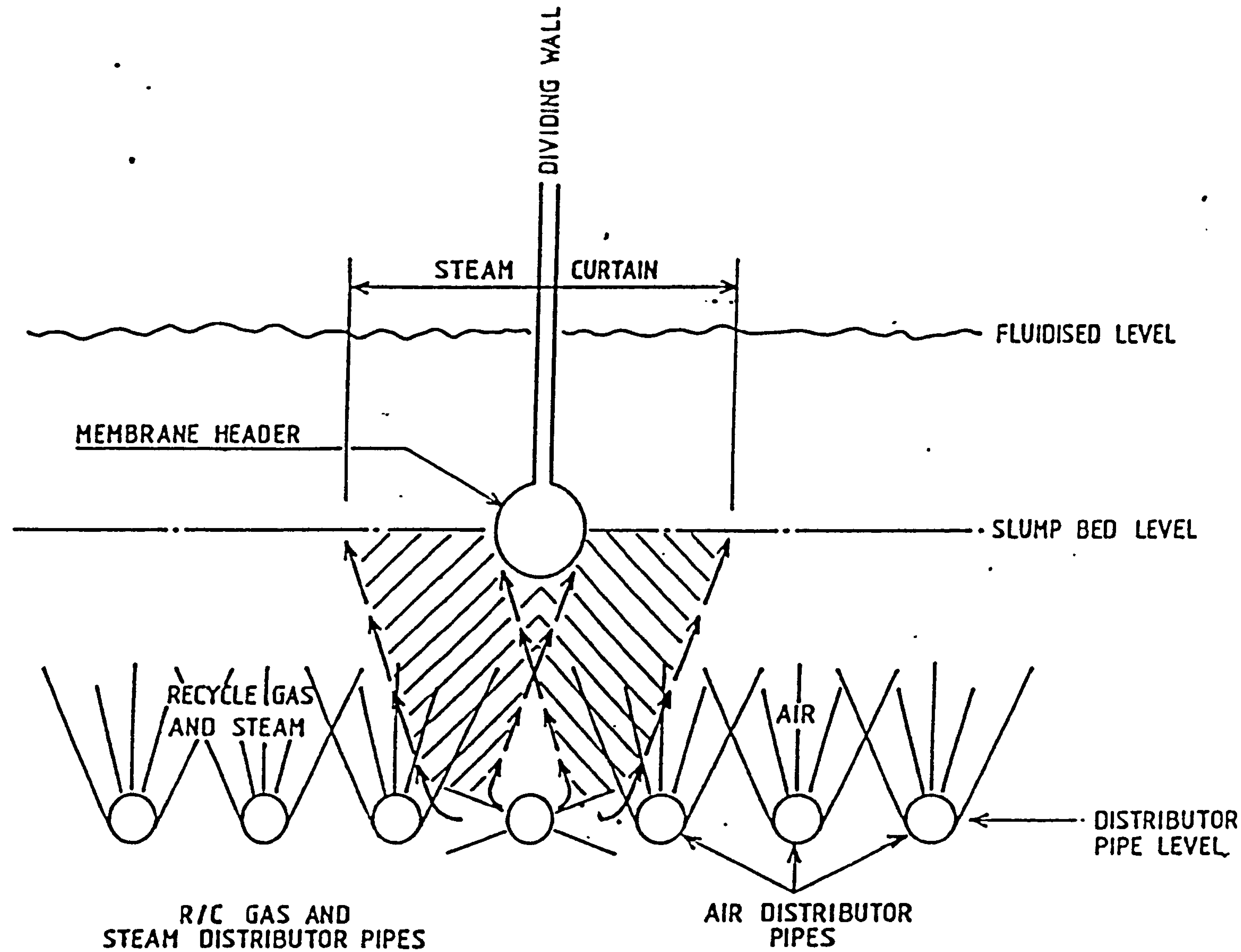


SIMPLIFIED SCHEMATIC OF 'ENERGY' DIVIDED BED M.C.V. GASIFIER – ELEVATION



THE ENERGY EQUIPMENT COMPANY LIMITED

M.C.V. GAS PRODUCER


DIVIDING WALL
STEAM CURTAIN
FLUIDISED LEVEL
MEMBRANE HEADER
SLUMP BED LEVEL
RECYCLE GAS AND STEAM
AIR
DISTRIBUTOR PIPE LEVEL
R/C GAS AND STEAM DISTRIBUTOR PIPES
AIR DISTRIBUTOR PIPES

TRANS MIGRATION FLOW PATTERN

TYPICAL GAS COMPOSITION

H_2O free basis

	L.C.V. PROCESS ACTUAL	M.C.V. PROCESS PREDICTED
VOLUME %		
CO	12	37
CO_2	15	Balance
CH_4	2	2
H_2	22	56
Higher hydrocarbons	Trace	Trace
H_2S	0·06	0·13
N_2	Balance	Trace
GROSS CALORIFIC VALUE		
MJ/m^3	4·90	12·00
Btu/ft^3	131	322

M.C.V. PROCESS EFFICIENCY

Ignoring carbon losses.

COLD GASIFICATION EFFICIENCY

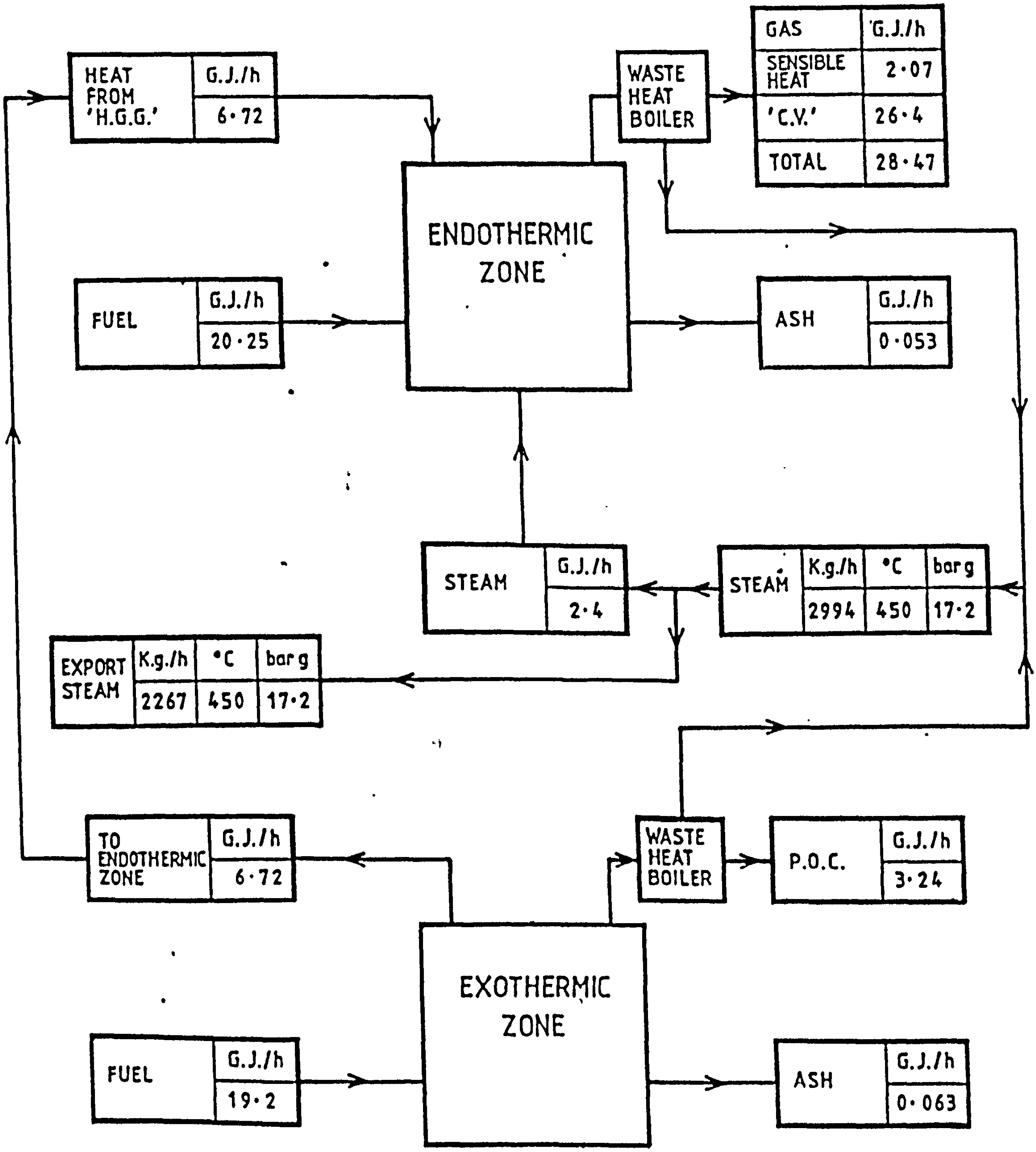
$$\frac{\text{Gross Chemical heating value of gas}}{\text{Gross heating value of fuel.}} \times 100\% = 66\%$$

HOT GASIFICATION EFFICIENCY

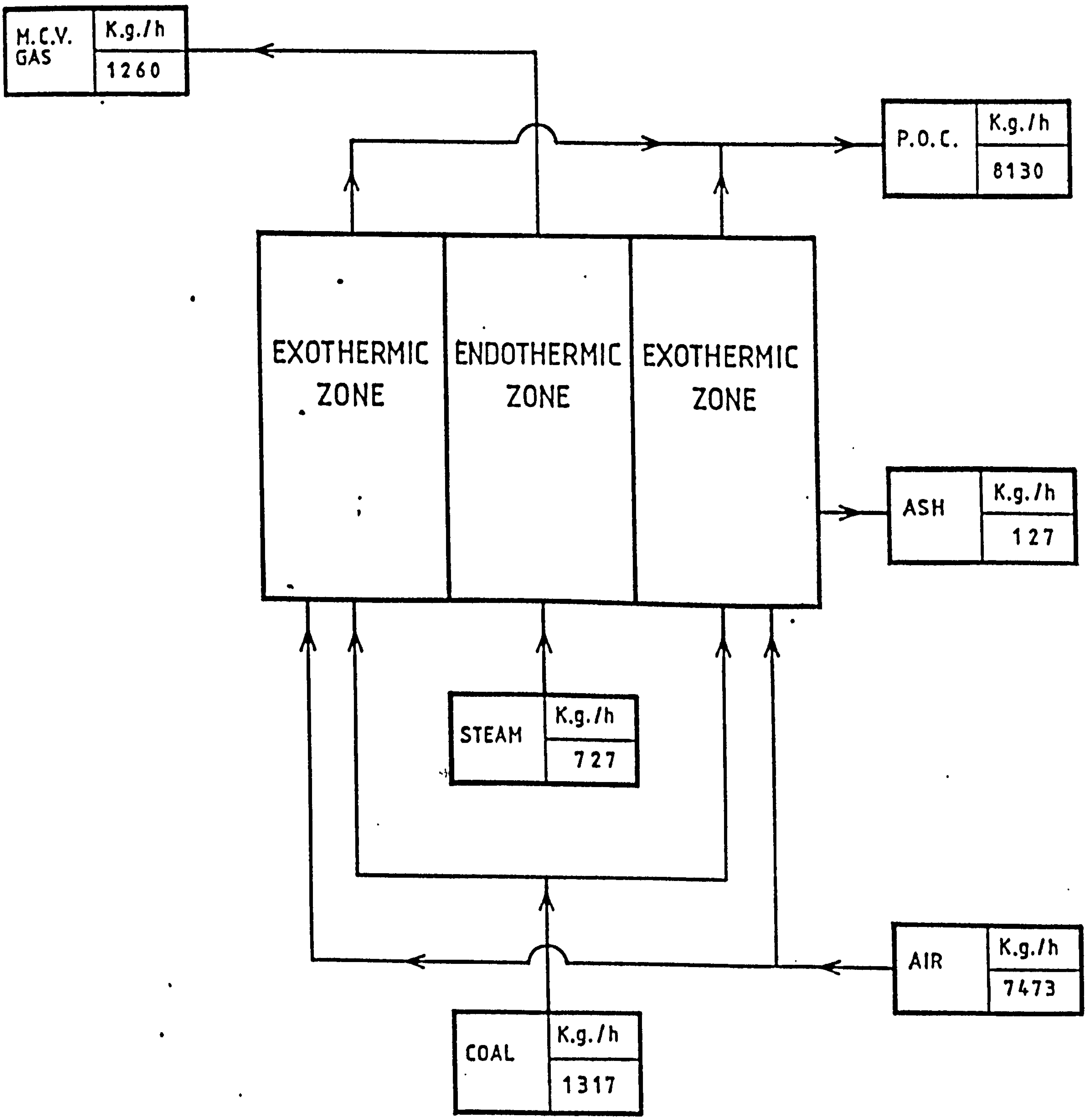
$$\frac{\text{Chemical \& sensible heat of gas}}{\text{Gross heating value of fuel.}} \times 100\% = 72\%$$

THERMAL EFFICIENCY

$$\frac{\text{Total. heat of gas \& export steam}}{\text{Gross heating value of fuel.}} \times 100\% \quad 85\%$$

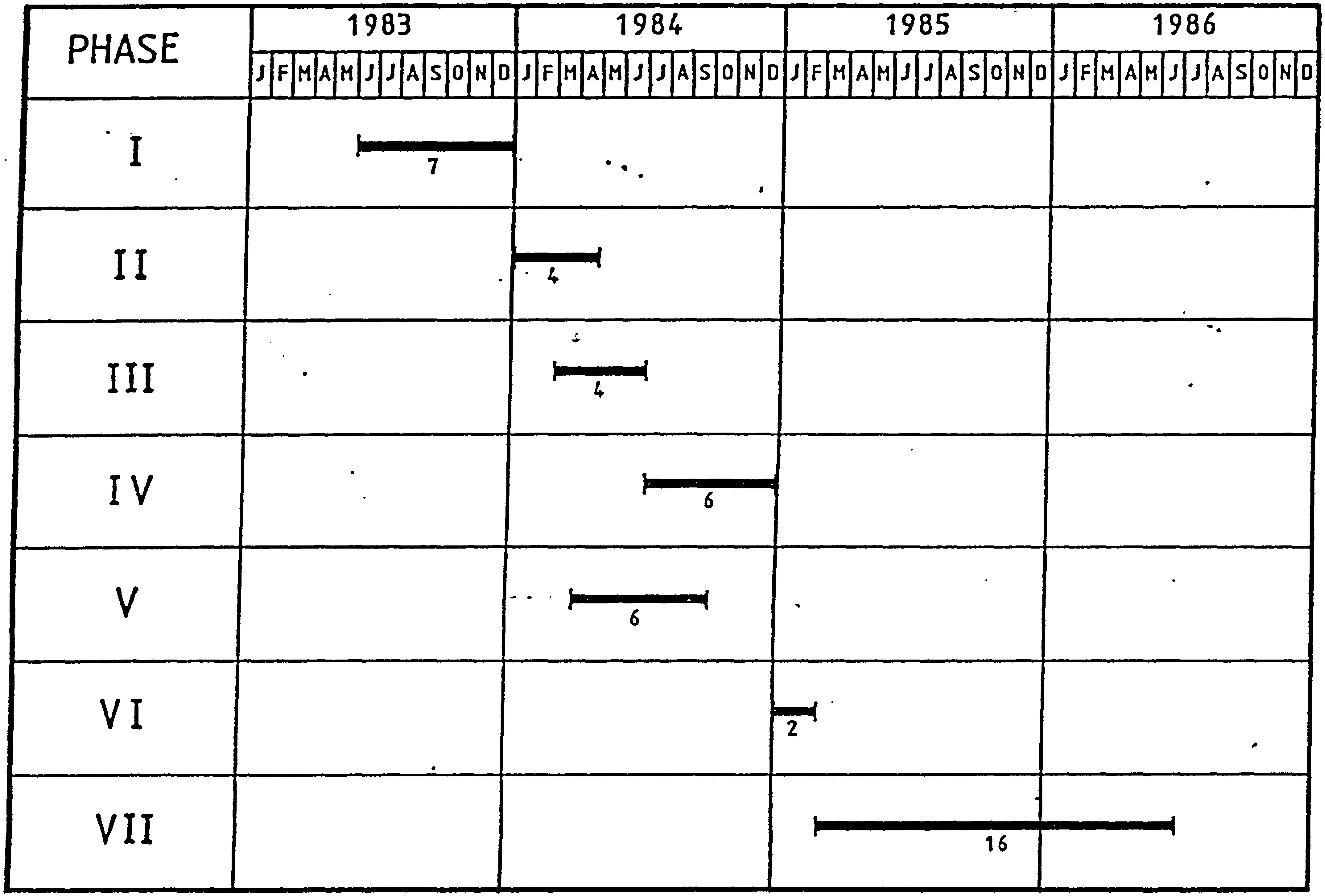


OVERALL ENERGY BALANCE



OVERALL MASS BALANCE

SCHEDULE FOR IMPLEMENTATION OF THE PROJECT



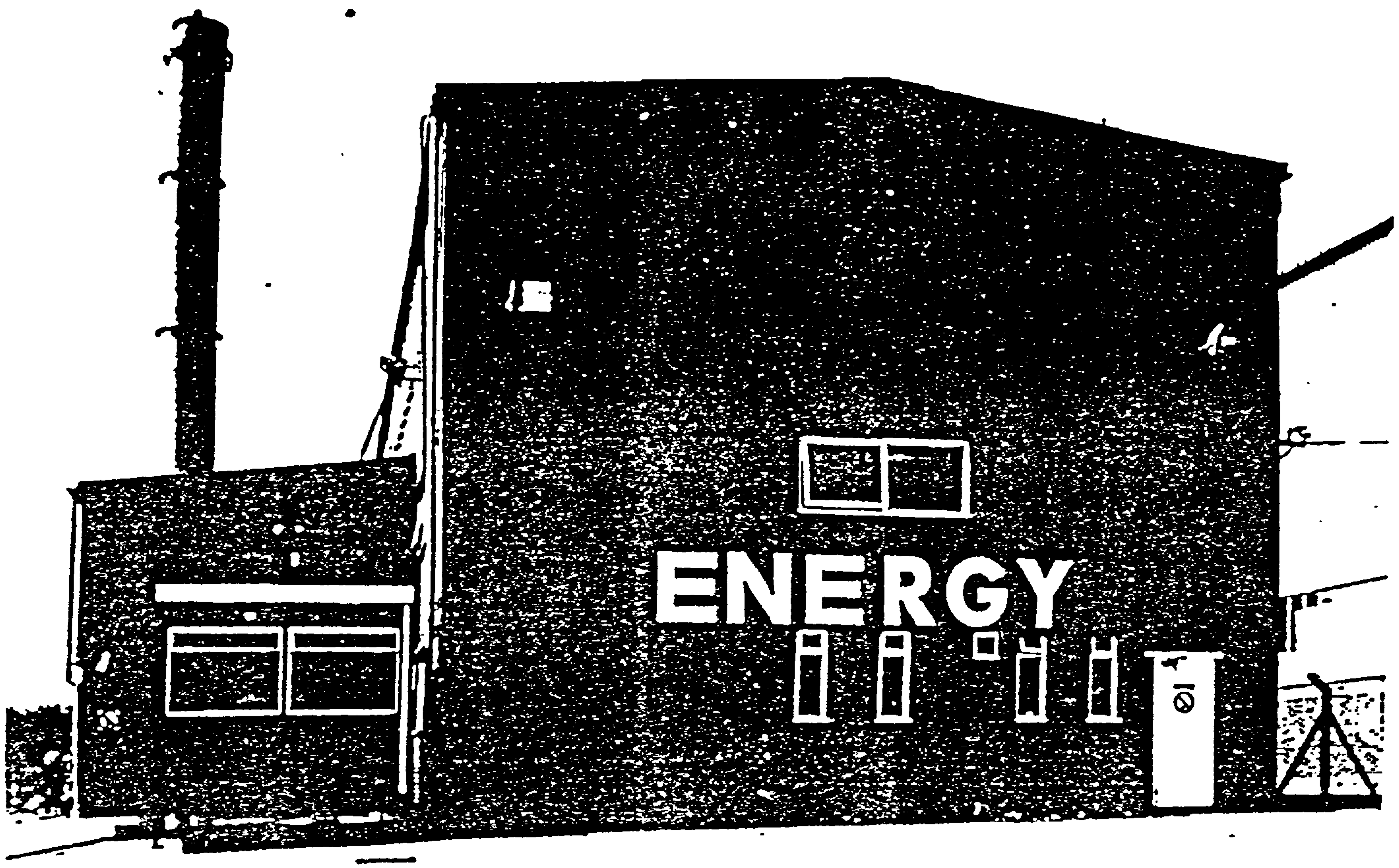


FIGURE 16.

This shows the outside view of the rebuilt Research and Development Centre at Olney.

The offices and control room are at this end of the building with the M.C.V. plant at the far end.

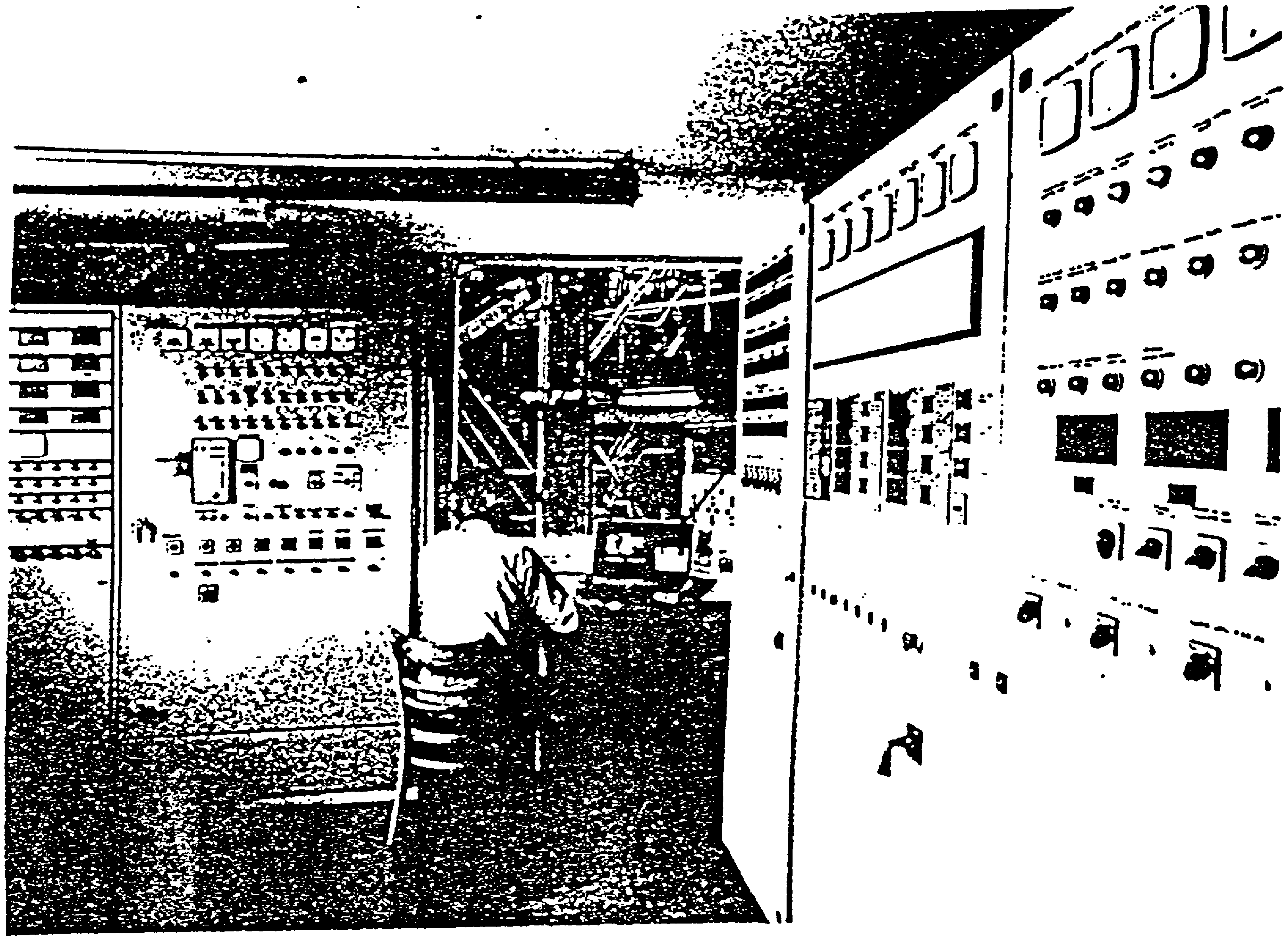


FIGURE 17.

This view shows the inside of the control room. The panel on the left houses the operational switchgear for the H.G.G. and L.C.V. units, whilst the panel on the right does the same for the M.C.V. unit.

Not shown are the infra-red spectrometer, the gas chromatograph and Sigma calorimeter.

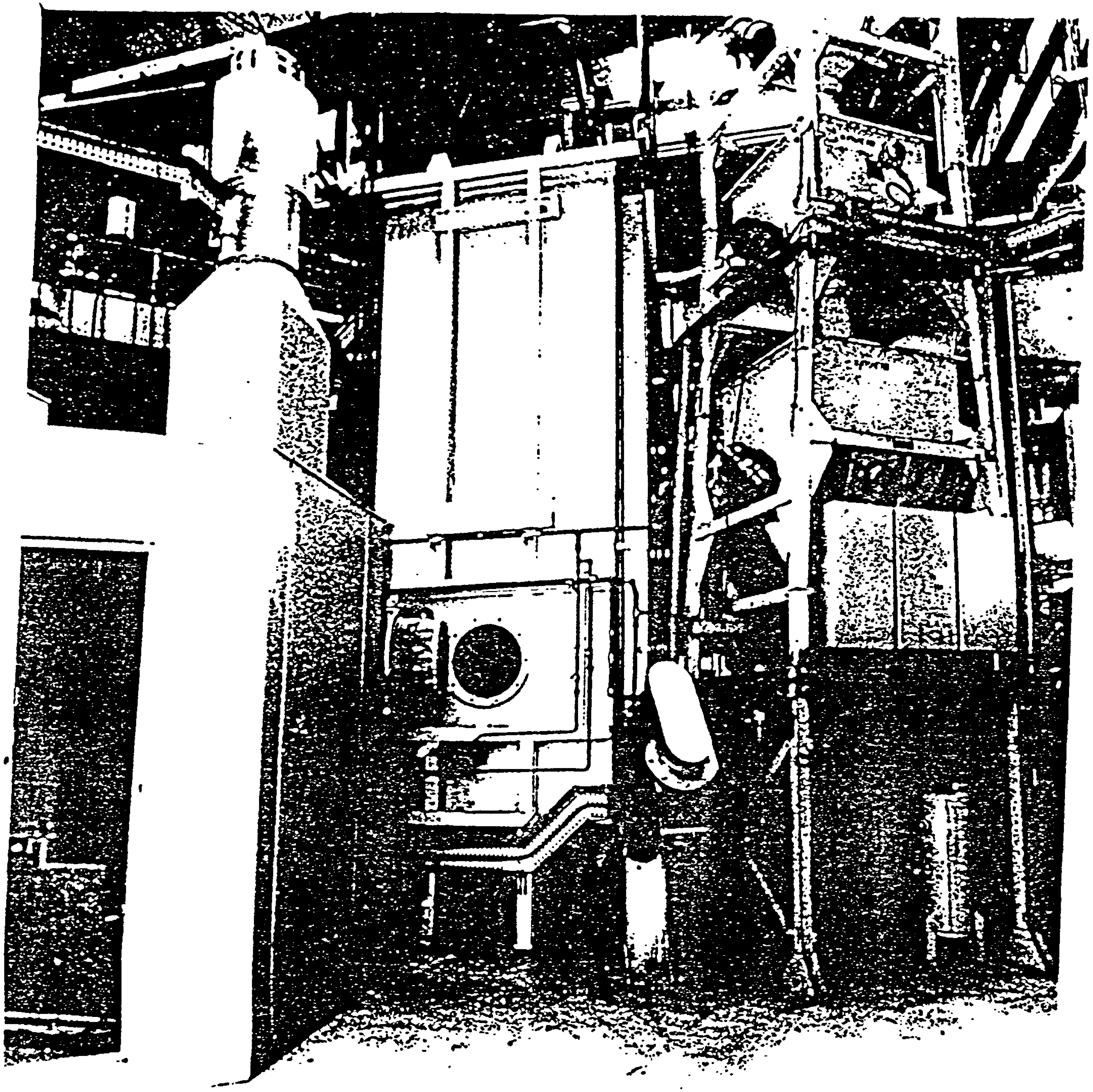


FIGURE 18.

This is an outside view of the M.C.V. unit showing the main body with the inspection door open to the right hand side exothermic chamber.

Below can be seen the main bed extractor screw.

On the left hand side of the photo can be seen the recycle gas fan inside its sound proof housing.

FIGURE 19.

This is a view inside the endothermic chamber showing both dividing membrane walls and below them the sparge pipes through which superheated steam enters the sand bed.

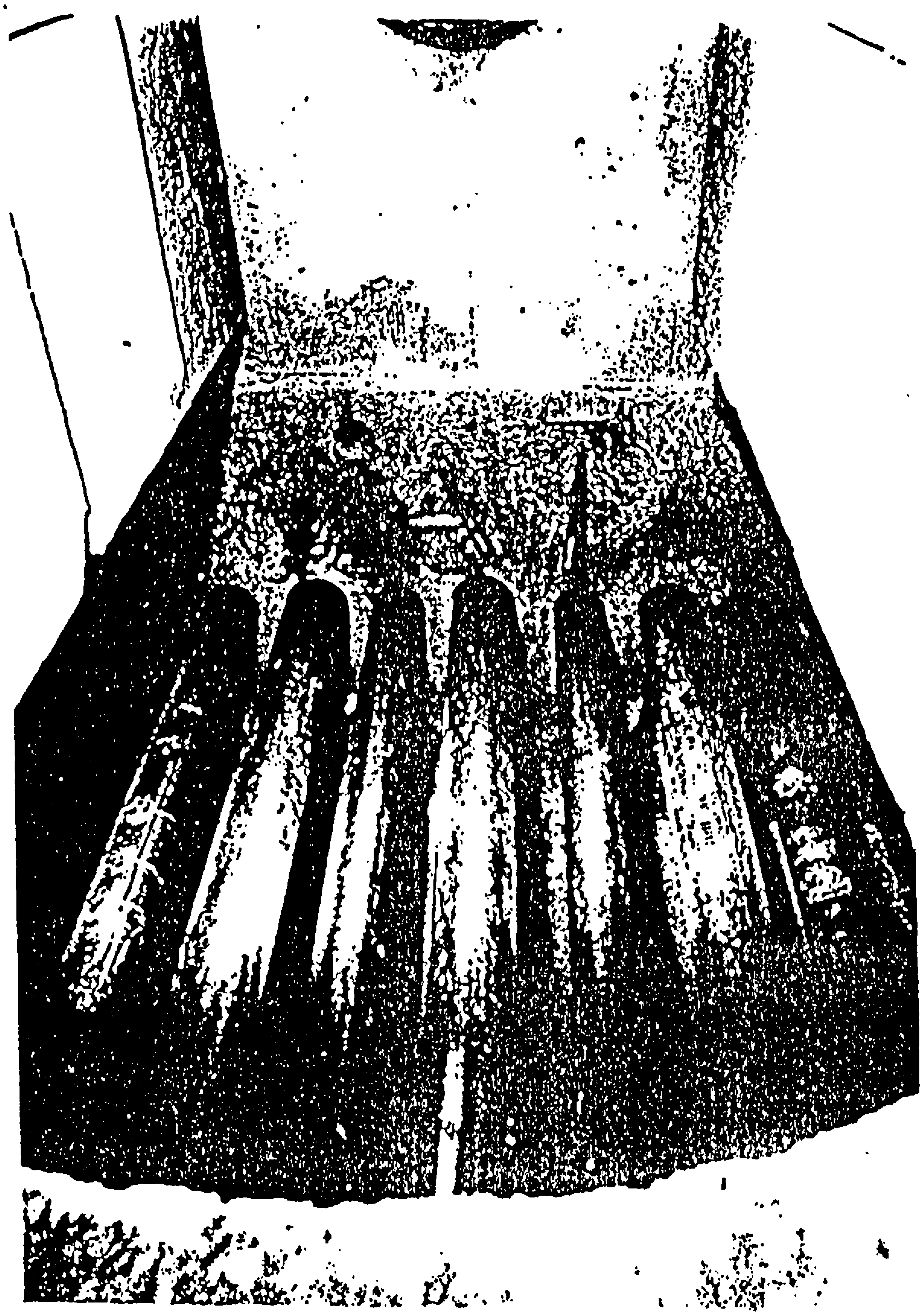
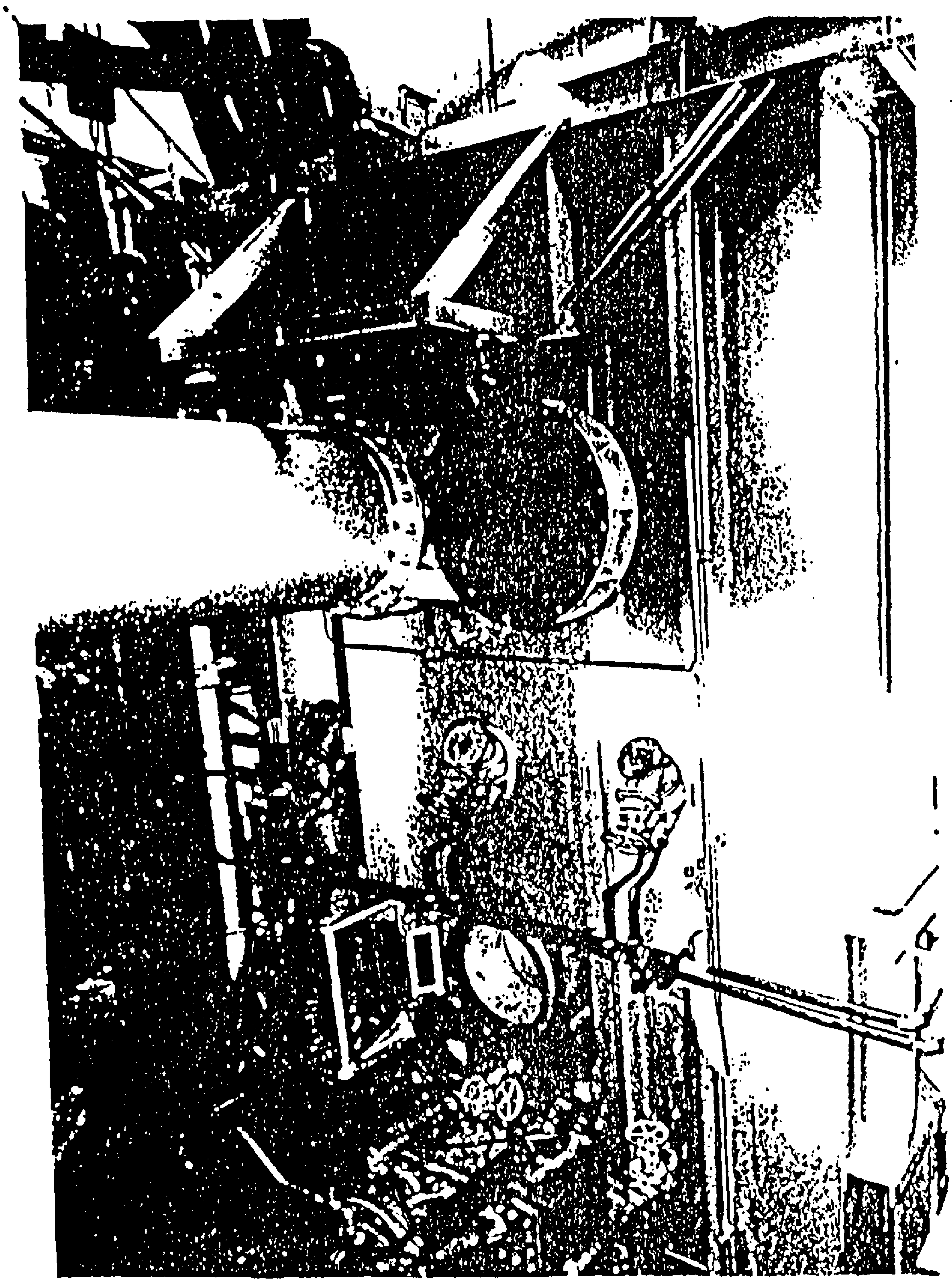


FIGURE 20.

This is another view of the M.C.V. unit taken from the roof of the recycle gas fan sound proof chamber.



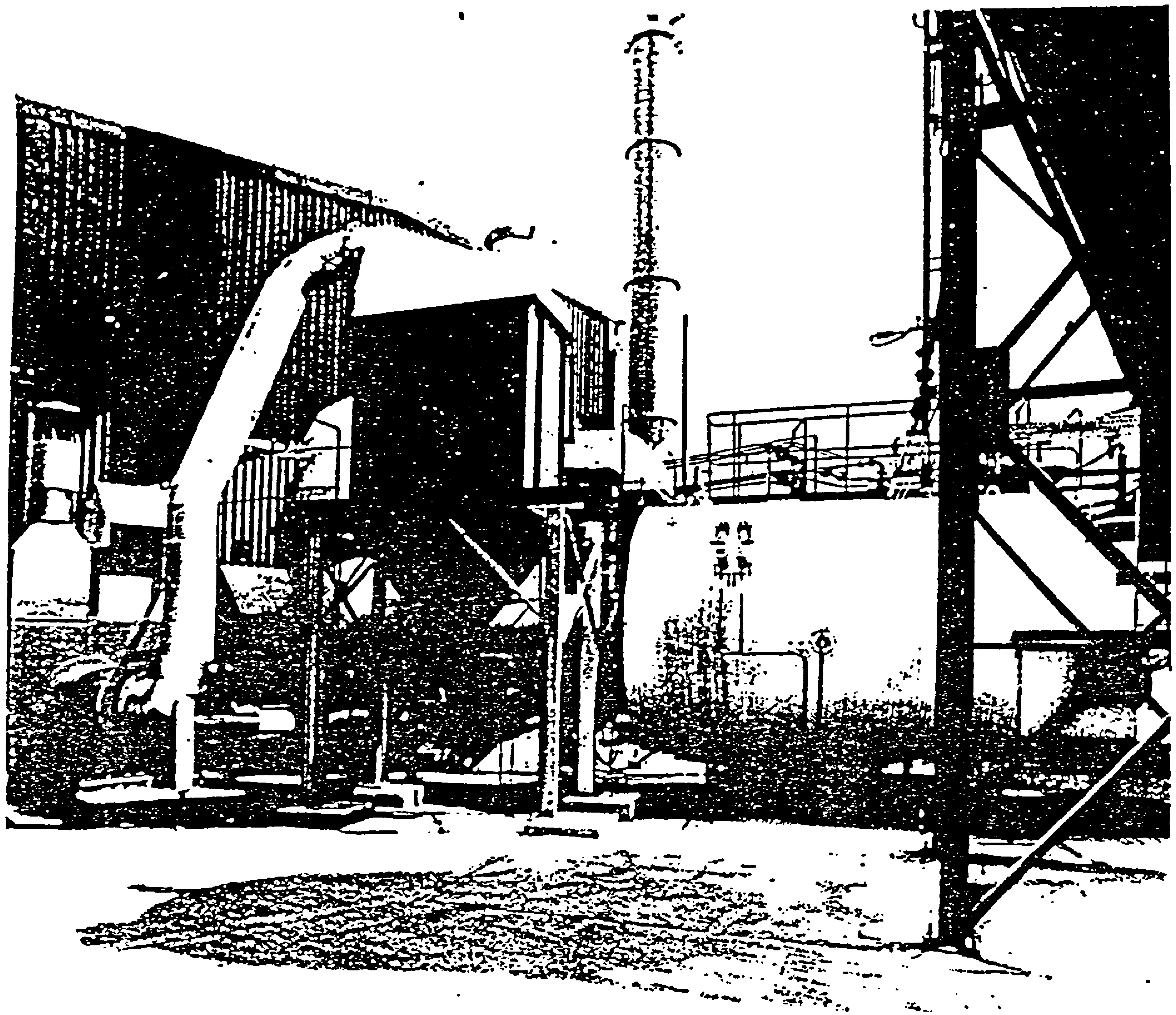


FIGURE 21.

An outside view showing the flue gas waste heat boiler and multi-cyclone.

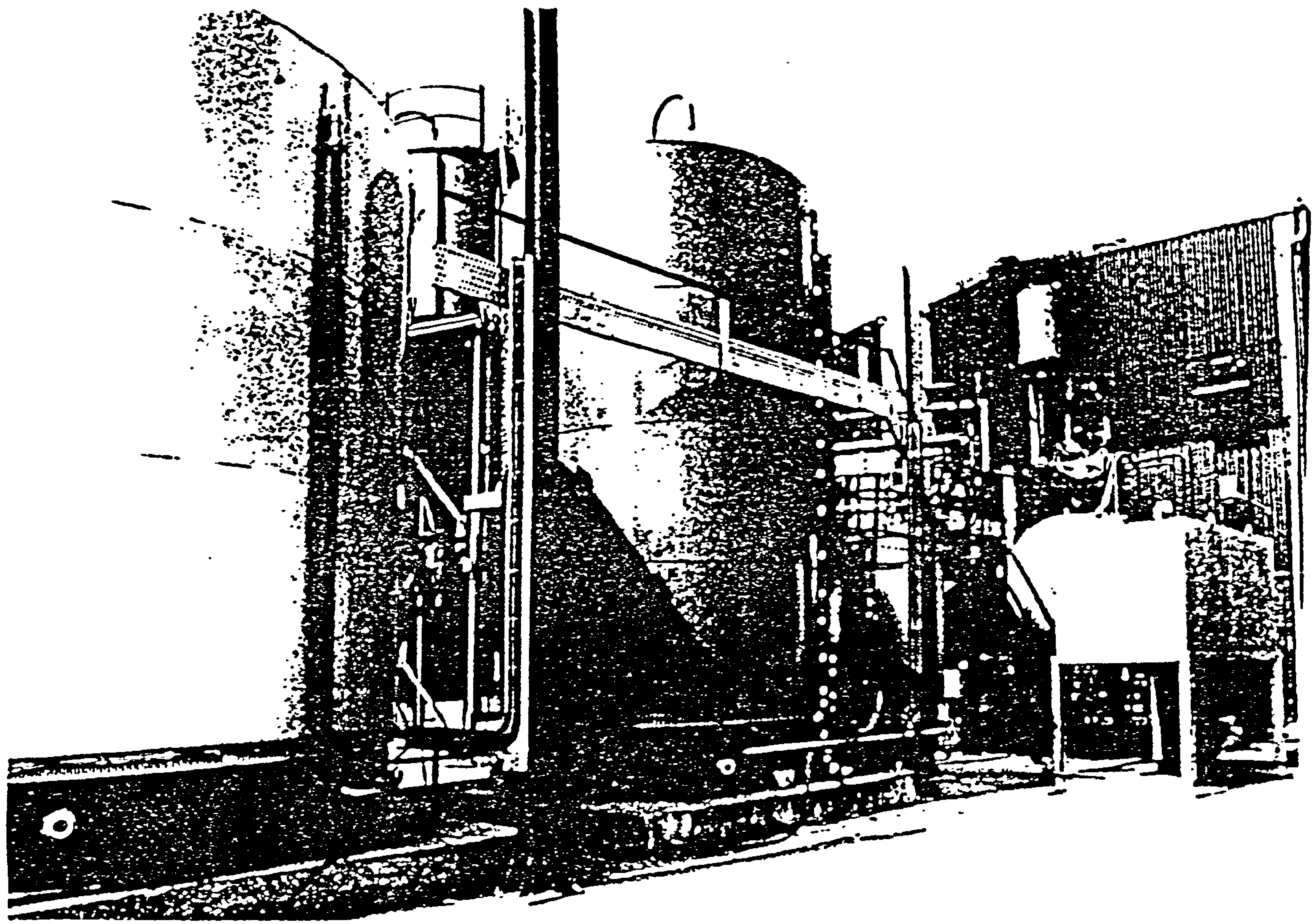
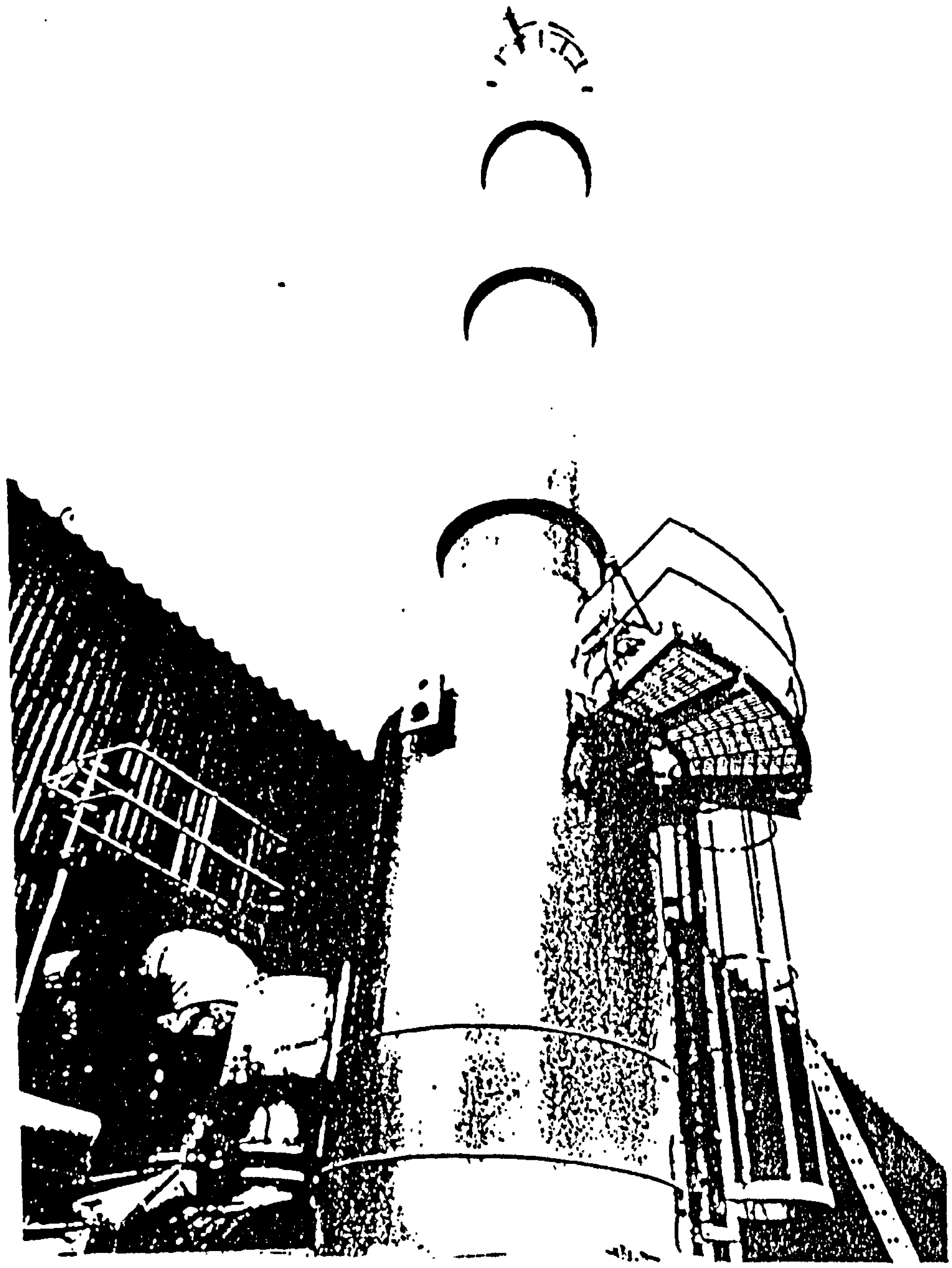


FIGURE 22.

Another outside view showing the oil and water tanks.

FIGURE 23.

This is an outside view of the test furnace installation and exhaust stack.



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APPENDIX 5

FLUIDISED BED COMBUSTION OF BOTH LIGHT AND WET BIOMASS

BY

B. WILTON & J.F. WASHBOURNE

ENERGY FROM BIOMASS CONFERENCE - VENICE, MARCH 1985

FLUIDISED BED COMBUSTION OF BOTH LIGHT AND WET BIOMASS

B. WILTON, University of Nottingham
J.F. WASHBOURNE, Energy Equipment Co. Ltd.

Summary

Although combustion of coals in a fluidised bed of sand is recognised as being a particularly efficient way of using poor quality fuel there has been little work on the fluidised bed combustion of biomass.
Problems of elutriation of fuel and sand may be encountered in fluidised bed units so to minimise these effects intermittant fluidisation and under-bed feeding is being used, together with a large expansion chamber in which some centrifugal treatment of the flue gases will be possible.
It is expected that moist biomass will be able to be used and this should be an attractive feature as it will reduce the need to store and/or dry biomass before use.
Fluidised bed combustion should also produce flue gases that are less polluting than those given off by other methods of combusting biomass.

When compared with most other fuels biomass is at best less convenient, while at worst it can be almost impossible to use. It is less energy-dense than oil, coal or fuel gases, it will not flow through narrow pipes (although it can be conveyed in fluids through large ones) and biomass harvesting, handling and storage can all present problems. Despite all these drawbacks it undoubtedly has a part to play in helping to meet the world's energy demands.

Two further major problems with biomass are that it can be extremely variable and in some cases it can be very wet. In general dry biomass fuels are utilised by combustion or gasification, whereas the only sensible way of using some of the wettest ones is by fermentation: as so often happens it is the in-between moisture content materials which cause problems.

Another feature of biomass that has to be considered is that the period of availability can vary considerably. Some materials are produced more-or-less continuously: if they happen to be wet, like waste animal slurries, then fermentation is obviously the most promising pathway to follow. If they are produced annually it is likely that there will be an optimum season for collection and this will almost invariably be followed by storage to even out the supply. It may also be found necessary to dry the material to some extent to prevent deterioration in store.

Perennial crops or their by-products are different: they can often be taken at more-or-less any season, however once again they will normally need to be dried either before or during storage

It would be useful to have available methods of rapidly extracting energy from moist biomass materials without the need to dry and store or ferment them. This would allow collection (harvesting) and utilisation to be consecutive operations, thus tending to minimise cost. Moist fuels can be used in gasification plants and in very large furnaces, however the cost of these appliances is high, so in the current work at Nottingham the aim is to develop a fairly small, and reasonably cheap unit capable of using these intermediate moisture content fuels.

The method selected - to use a fluidised bed combustion unit - is also expected to minimise the risk of producing polycyclic aromatic hydrocarbons in the flue gases. This is an important factor in biomass combustion because of the carcinogenic properties of these compounds. The addition of limestone or dolomite to the bed will also retain much of the sulphur present in the fuel.

With the exception of pulverised fuel combustion systems, it is usual in conventional combustion equipment for the fuel to be fairly static in the combustion zone; in such conditions moist fuel will dry out only slowly. In fluidised bed units, however, the rate of heat transfer between particles is high; moist fuels should dry out quickly, their volatile components should be given off rapidly and combustion of the non-volatiles should be achieved in a short time.

One of the chracteristics of the fluidised combustion of coal is that of carryover of ash and, in some designs of plant, of the elutriation of the bed material. Some biomass materials have relatively finely divided components, for example leaves and pieces of fibre, so elutriation of these fractions could cause a problem. The longer that such materials can be retained in the bed the better, so it was decided to feed the fuel into the bed below the surface rather than adopt the simpler approach of dropping it onto the bed (see Figure 1).

Several other design features should minimise the problems of elutriation and moist fuel; these include

(i) having a slotted fuel feed tube which runs through the bed,

(ii) using a fluctuating velocity air supply to the bed so that for several periods of a few seconds duration each minute the bed is not quite fluidised, and

(iii) having a large expansion chamber above the bed, with secondary air being introduced tangentially to encourage centrifugal separation of particles entrained in the flue gases.

There is very little published work in this area and a large number of materials and design features need to be studied. Among the materials that will be used first are chopped cereal straw, freshly produced wood chips and bagasse pith: once the preliminary work on these three widely differing potential fuels has been started, work on mixtures of biomass fuels may well be undertaken.

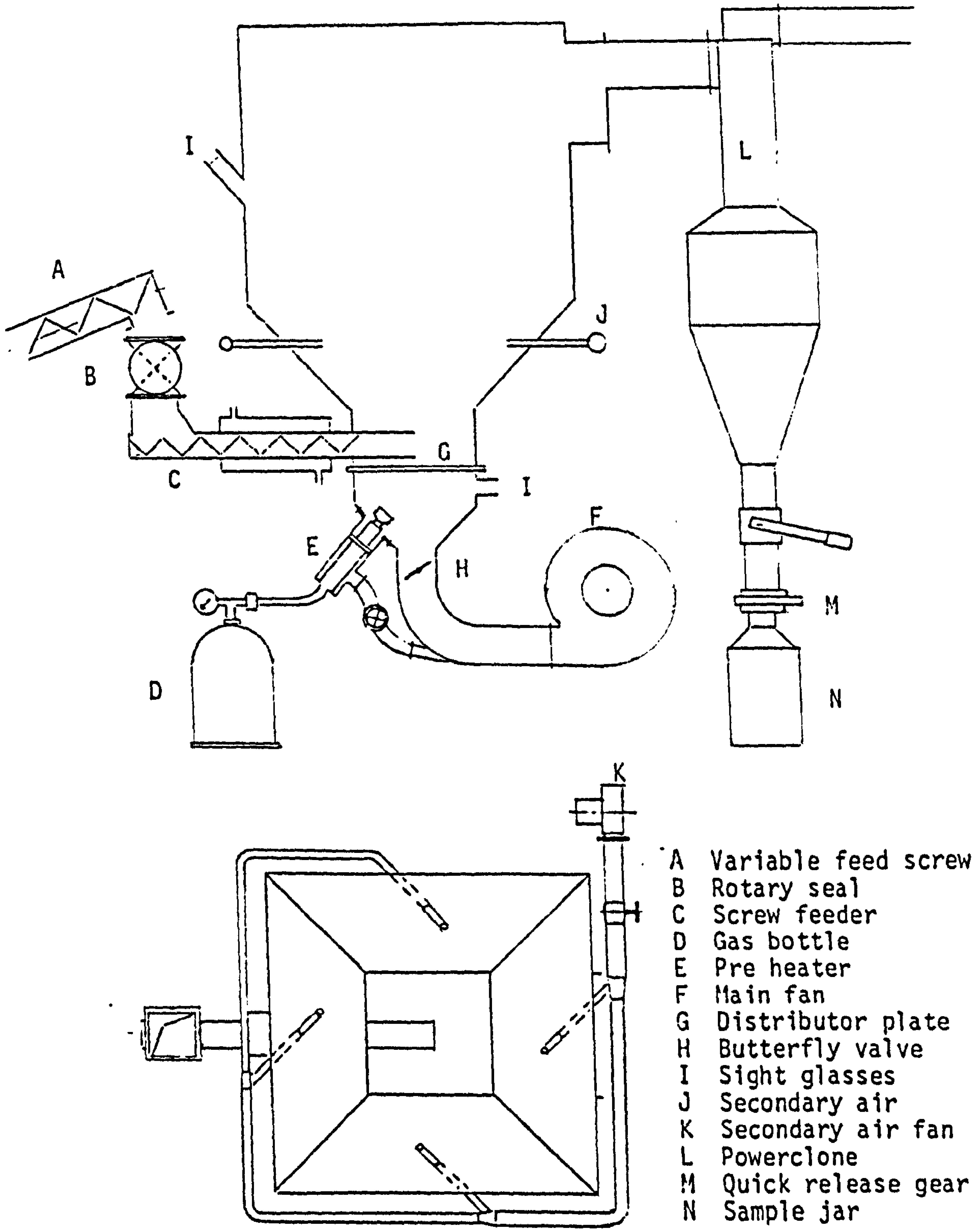

A
B
C
D
E
F
G
H
I
J
K
L
M
N
A Variable feed screw
B Rotary seal
C Screw feeder
D Gas bottle
E Pre heater
F Main fan
G Distributor plate
H Butterfly valve
I Sight glasses
J Secondary air
K Secondary air fan
L Powerclone
M Quick release gear
N Sample jar

Figure 1

0 1m

A P P E N D I X 6

THE CASE FOR THE DESIGN CONTRACTOR

BY

J. F. WASHBOURNE

ENERGY MANAGER - VOL 5, NO 10 - OCTOBER, 1982

The case for a design contractor

J. Washbourne explains how careful plant design will pay dividends later

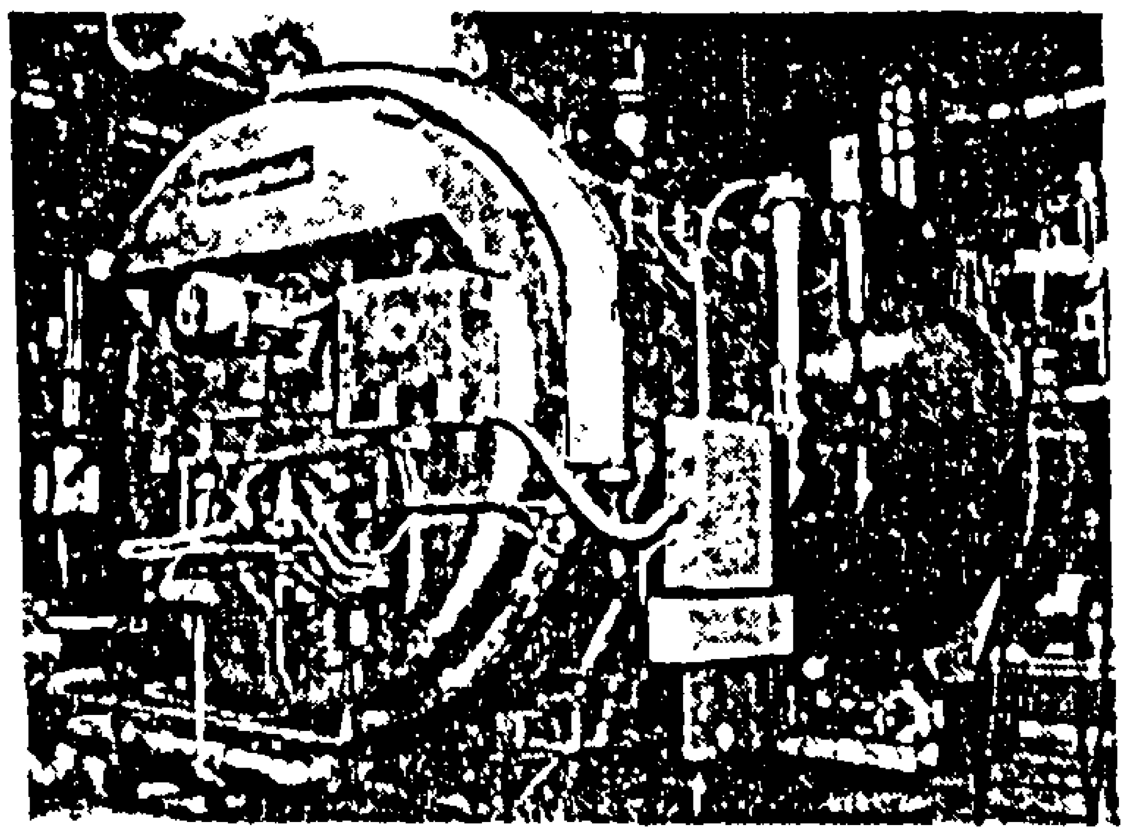
Design and installation of a steam boiler plant involves expertise from all the engineering disciplines. The basic and most important factor is to interpret correctly the true requirement by careful study of the potential steam usage under possible maximum and minimum steam demand. This may involve removing conditions set by the customer, that are difficult to meet satisfactorily. Choice of fuel is very important. The prospect of having to operate a coal-fired boiler plant tends to be unpopular with engineers who have always had the convenience and reliability of oil fuel or gas. This is where the experience of the design contractor can instil confidence.

Closely associated will be the choice of combustion equipment and the type of boiler. For example, the decision may well be that the plant will be oil fired, however, with an eye to the future, the new plant has to be capable of being converted for use with solid fuel. Boiler selection will require knowledge of the probable development in solid fuel firing equipment and the potential for conversion of presently available boilers. A decision in favour of coal firing means additions to be incorporated into the boiler plant for coal storage and coal and ash handling. The choice between belt conveyors, bucket elevators, pneumatic systems employing fans or air compressors can involve expensive mistakes.

The requirements for the proper storage and handling of liquid fuels are subjects of British Standards, nevertheless, there is a considerable margin for the discretion of the designer – especially in securing correct oil temperature at the burner – a very vital necessity with currently available Heavy Fuel Oil grades. Similarly, the accurate sizing of pipework can minimise the cost of expensive natural gas regulators.

The choice has to be made between the variety of boiler types and make available. Important factors must be budget, delivery dates and the possibility of the user being able to visit a boiler plant that has some similarity with the plant he expects to operate.

The type of water treatment plant will be dictated by the boiler manufacturers' recommendation on the quality of feedwater and the characteristics of the raw water available. This is very much a case where practical experience will temper salesmen's enthusiasm, especially when the client's budget is limited. There may well have to be a choice between two or three

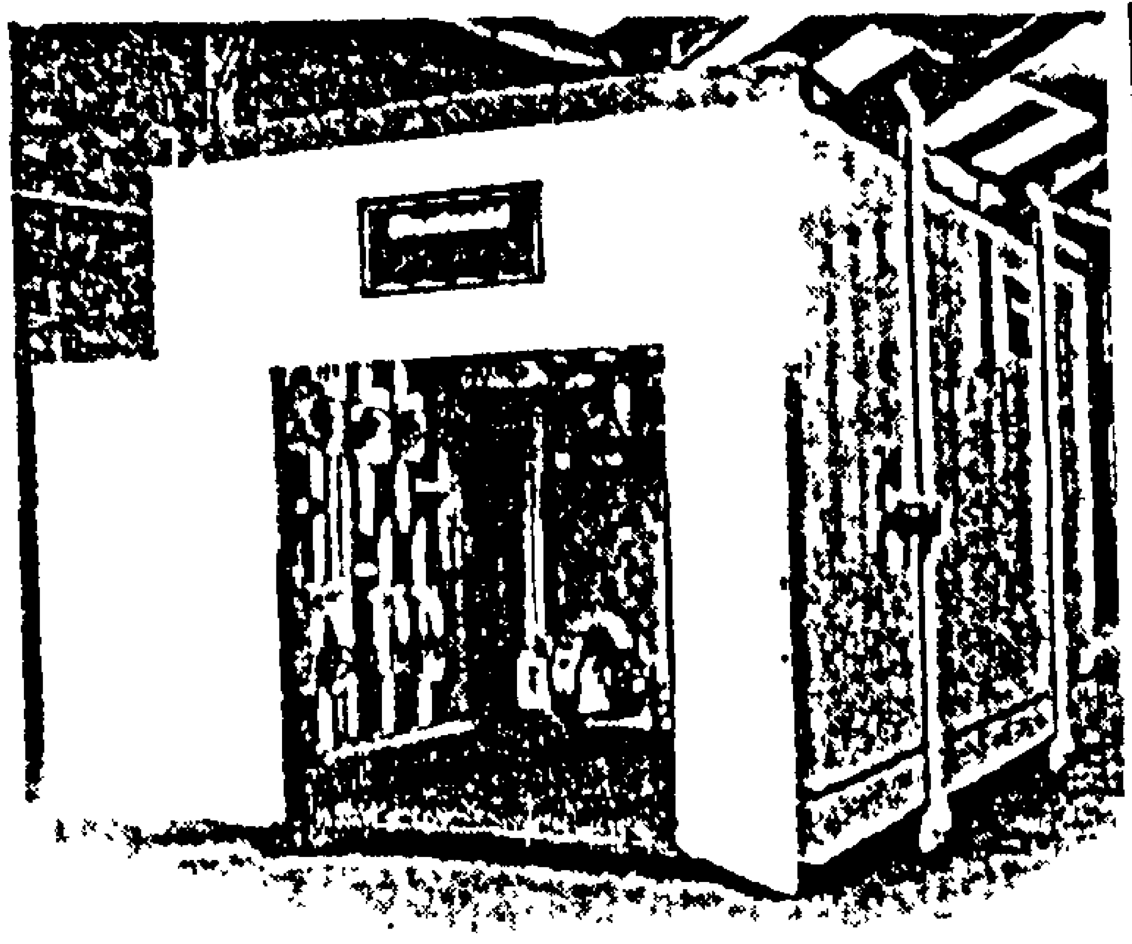


The new boiler house at British Rail, Clacton. The acoustic hood has been swung back on one of the Allen Ygnis boilers to reveal a special burner assembly. Circle 149.

options which will influence other equipment. With a wide selection of types and makes of feedwater pumps and the choice between on/off or modulating control of feedwater supply to the boiler(s), choosing correctly by taking account of the nature of steam demand will have a significant effect on the wetness of steam leaving the boiler. By making the most suitable choice of pump capacity and of type and make, the use of electric power will be minimised and with proper provision made to cater for breakdown, the utmost reliability is built into the boiler plant. The disposition of hotwell (and deaerator, if required), is important and dependent on a number of inter-related factors. The pressure head provided by the hotwell can dictate acceptable feedwater temperature. Total water capacity must be matched to a maximum rate of steam demand and the capacity of the water treatment plant over regeneration periods. The provision made for the return of condensate and its maximum and minimum temperature and flow rate will be amongst the factors to be taken into account.

Circulation at the intake of a feedwater pump will result in constant difficulty due to operation of low water level safety control of the combustion system. Similarly, inadequate reserve of treated water at times of maximum steam demand can result in unacceptable build ups of TDS due to excessive use of raw water. On the other hand excessive water capacity can be an expensive luxury over the life of the boiler plant, especially with boiler plant that is operated on an intermittent basis due to heat losses from the stored water.

Continuous, automatically controlled, blow-down with heat recovery to boiler feedwater can be cost effective where there is only a relatively low proportion of returned condensate but may be of little benefit where steam is used in circumstances of high condensate recovery such as space heating. The choice between blow-down pit or blow-down vessel or for the recovery of flash steam are again matters requiring judgement based upon experience. A badly constructed blow-down pit can result in undetected erosion of chamber and drains which may become



The picture shows the Macdonald Heatcentre that has been purchased by the Scottish Development Agency. It is a specially designed factory built packaged hot water boiler house that has an output of 2.6 million BHu's/hr. Circle 148.

dangerous before being discovered. Poor siting of vents from pit or vessel can be expensive to rectify.

The design of the waste gas handling system must meet the local Planning Authorities requirements, the provisions of the Clean Air Act and aim to avoid emission of dust or smuts.

For a new installation the choice has to be made between independant flues and chimney liners for each boiler, or for a common waste gas handling system to serve all boilers, or independant flues from each boiler to a common chimney. For the correct decision the range over which steam demand can alter must be assessed accurately in order to establish the variation in the quantities of waste gases to be handled.

The choice of chimney fabric and exit sealing dampers should take account of thermal storage and heating up time.

Temperatures of surfaces of flues and chimney liners should be above the dewpoint temperature of the gases. The provision of drains for the removal of condensate from a chimney serving intermittently operated gas or oil fired boilers is often overlooked. Other practices to avoid include lining a brick chimney with porous insulating refractory for use with gas fired boilers or using a loose fill insulating material in the windshield of a multi-core chimney.

The re-emergence of coal for use with industrial steam boiler plant introduces the factor of grit emission from chimneys – especially with existing waste gas systems – and the probable need for induced draught fans to cope with grit arrestment equipment. For the inexperienced the potential for less than optimum plant performance or expensive mistakes is a real hazard.

Excessive emission of grit and dust can be expensive to cure and will result in difficulties with neighbours and Local Authorities.

J. Washbourne is Sales Manager of the Energy Equipment Co.

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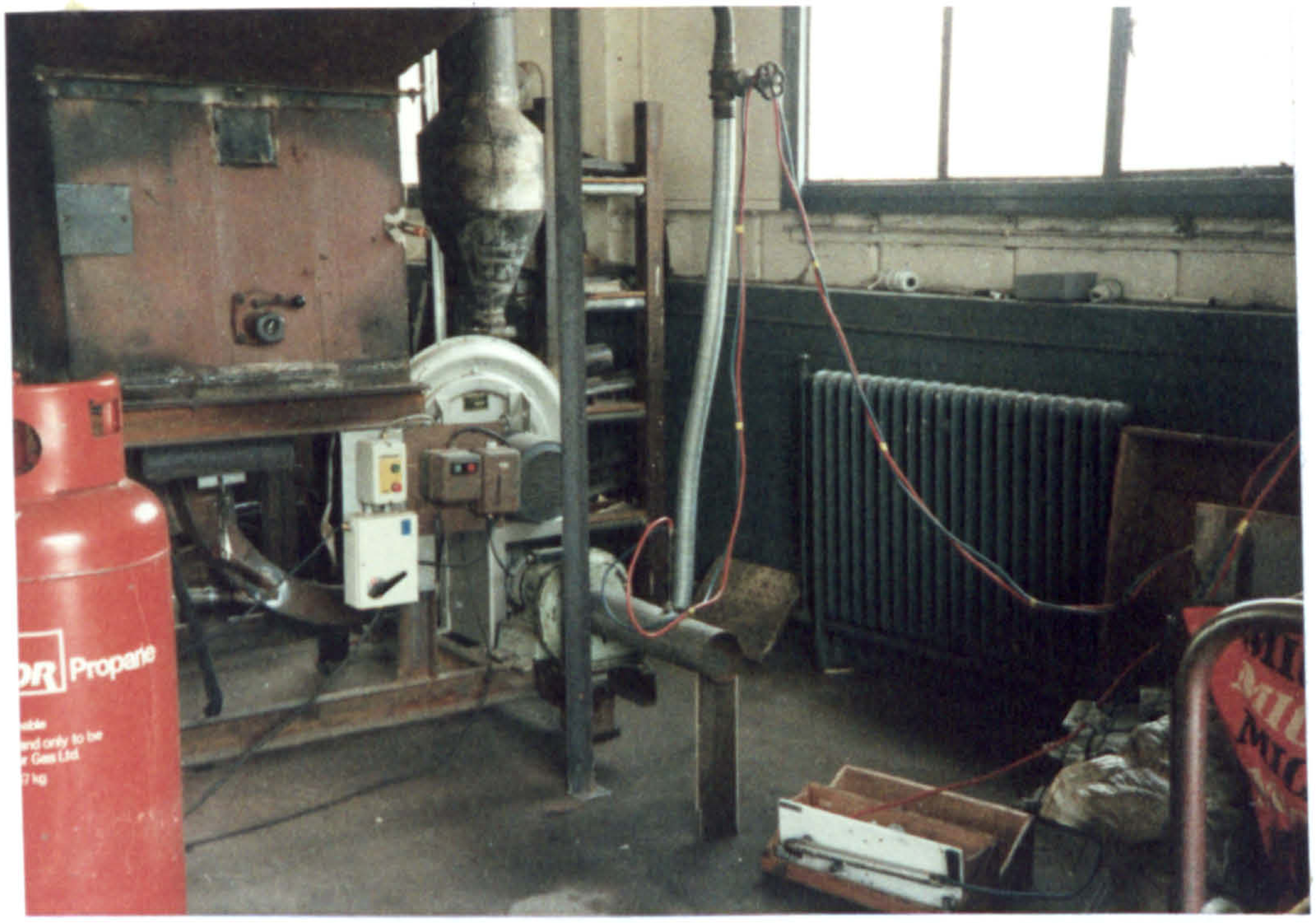
VIEW OF FUEL INTAKE

(1)



MAIN FAN, SECONDARY AIR FAN AND CYCLONE

(2)

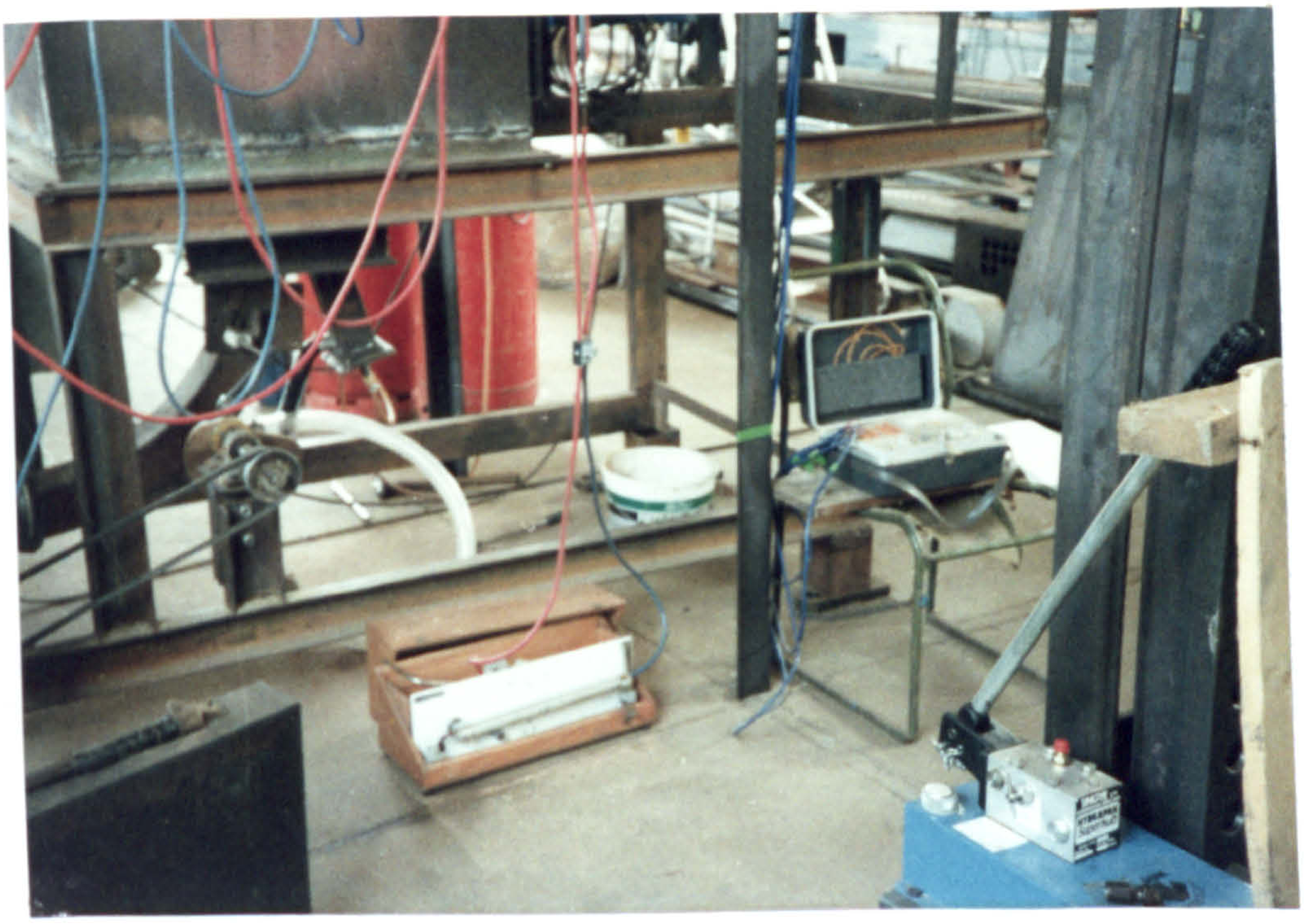
STRAW CONVEYOR

(3)

VIEW SHOWING INTAKE FEEDER

AND THERMOCOUPLES

(4)



AIR FLOW AND TEMPERATURE RECORDING EQUIPMENT

(5)

CAM OPERATED BUTTERFLY VALVE

AND START UP BURNER

(6)