A STUDY OF THE BURNING CHARACTERISTICS OF SUSPENDED PEAT PARTICLES

BY

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A Thesis

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ABSTRACT

Burning characteristics of suspended devolatilised peat particles obtained from Khulna (Kola-Mouza) were studied under different conditions. The variables studied were : reaction temperature (400 - 800°C), dry air flowrate (60, 125, 180 l/h), particle size (d = 0.42 cm, 0.54 cm, 0.68 cm, 0.88 cm) and reaction time. The study showed that the burning rate of devolatilised particles depends largely on the variables studied. Experimental results reveal that specific reaction rate is directly proportional to temperature but inversely proportional to diameter. The combustion of peat particles is not diffusion controlled as indicated in the present investigation. The results obtained shows that known expression of surface reaction rate co-efficient for coal-char is invalid for devolatilised peat particles. Burning rate of peat in flow system decreases with time, initially rapidly and then slowly. This is largely due to diffusion of oxygen through the retaining solid ash becomes an added resistance in the process of fixed carbon burning.

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CHAPTER-1 INTRODUCTION

1. INTRODUCTION:

A large fraction of the export earnings is being used to import a small percentage of the total energy consumed in Bangladesh. The fact that Bangladesh is to spend about 45% of its yearly total export earnings to import commercial fuels like crude oil, petroleum products and some coal, which accounted for only 9.8% of the total yearly primary energy supply in 1983-84 (1), indicates the existence of serious problems in its energy sector as well as in its economic development. In order to relieve the economy from this strain the energy planners of the country are trying to evolve an alternative commercial energy mix that will have a greater share of presently unutilized or underutilized energy sources available in the country.

Biomass fuels often referred to as traditional or rural fuels provided 76.7% of the total energy requirement of the country while the remaining 13.5% was supplied by the indigenous natural gas (13.0%) and hydro energy (0.5%) in 1983-84 (2). The numbers quoted also includes non-energy use e.g. natural gas as feedstock for fertilizer production. When compared with the figures of the year 1973-74, one may see how Bangladesh has failed to switch over to the so called commercial fuels in the last 10 years which are considered essential for higher food production and industrilisation of the economy leading to improvement in the quality of life.

TABLE - 1.1: Comparative energy use in 1973-74 & 1983-84 in Bangladesh.

Type of energy	1973-74 (% of total)	1983-84 (% of total)
1. Imported fuels	18.4%	9.8%
2. Natural Gas	8.8%	13.0%
3. Hydro Power	0.3%	0.5%
4. Bio-mass fuel	72.5%	76.7%

Imported crude oil, refined Petroleum products and coal provided 18.4% of the total primary energy supply in 1973-74 demanding a much lower percentage (about 40-50%) of the total export earnings (3). Indigenous natural gas, hydro energy and biomass fuels supplied 8.8, 0.3 and 72.5% of the total respectively. So, contrary of expectations, traditional fuels are contributing relatively more now than before. In absolute terms, the yearly energy supply by biomass has increased from 211 PJ in 1973-74 to 489 PJ in 1983-84, with about three fold increase in agricultural and tree biomass supply. This has brought about deforestation in the country. On the otherhand, the increase in the yearly supply of commercial energy is from 80.32 PJ to 148 PJ in the same period, primarily in the supply of indigenous natural gas (more than 3 fold). The population in the country increased from around 77 million to about 100 million during the same period giving an increase in yearly per capita energy supply from 3.78 GJ. (Traditional - 2.74 G.J., commercial - 1.04 G.J) to 6.38 G.J. (Traditional - 4.89 G.J, Commercial - 1.49 G.J). This is less than 8.54 G.J needed at subsistance level (4). The per capita energy supply has increased primarily in the traditional sector.

If we turn to the rural areas where about 90% of the population live, the situation is even worse. Whatever little commercial energy is used, is primarily confined in the agricultural sector and household lighting by kerosene. Diesel engines are generally used to run irrigation pumps which irrigate only about 11.5% of the 22 million acres of cultivable land (5). Chemical fertilizers constitute an indirect commercial energy supply to agriculture. The use of fertilizer per acre is low compared to other countries of the world. In 1980-81, the total commercial energy used in the rural sector is estimated at 32.0 PJ, while the biomass fuels supplied about 416 PJ (mainly for cooking), giving a rural per capita consumption of about 5.41 GJ (Traditional 5.02 GJ, commercial - 0.39 GJ) (6).

At present the Government has emphasised on maximising use of natural gas. Gas transmission line expansion work programme has been taken up on a large scale. At present the estimated gas reserve in Bangladesh is about 10.5 trillion cubic feet. Gas utilization is limited to power stations, fertilizer plants, tea estates, industrial units, commercial and domestic purposes. The benefits of indigenous natural gas are being enjoyed by the areas in the eastern part of the country. But no gas could be provided to the western zone due to the problems of river crossing. On the otherhand a considerable reserve of peat exists in the districts of Faridpur and Khulna both situated on the western side of Bangladesh. A number of smaller deposits also exist in other parts of the country.

The total estimated peat reserve in Bangladesh (8) is 600 million tons (dry basis) spread over Faridpur area (125 million tons in an area of 116 sq. km, having peat thickness of 2m and overburden thickness of 1.5 m), Khulna area (8 million tons in an area of 39 sq. km, having peat thickness of 2-3m and overburden thickness of 0.3 - 0.6 m) and many other small areas of south west region of the country. The quality and physical properties of the peat as detailed by BEPP suggest that it may be considered for commercial extraction for power generation and for industrial and domestic use. This will lead to the decrease of deforestation rate and the conservation of indigenous natural gas.

In the past, peat had been used as a fuel for domestic purposes and small industries in many countries of the world to a limited scale. But with the advant of cheap premium fuels like oil and natural gas its use was greatly reduced. Recently, the interest in peat is again growing in countries like U.S.S.R., Finland and Ireland due to depletion and price rise of oil and natural gas.

Peat found in Bangladesh is marsh peat. Its ash content is high and variable (8). Samples drawn from various places were tested in the fuel laboratory of Chemical Engineering Department, BUET. The ash content of the samples were found to vary from 30 to 62%, whereas peat found in cold climate is fibrous in structure, local peat is structurally collidal (9).

In a recent report by Mahtab and Islam (10) on present status of local peat utilization, it has been mentioned that in certain rural location, locally available peat is extracted, dried and stored for using as fuel in household, cooking, in tea stalls and in brick kilns, mainly in the dry periods. In 1980, BMEDC (Bangladesh Mineral Exploration Development Corporation) conducted studies on the peat obtained from Kola Mouza in Khulna. The production cost of peat has been calculated as 253.15 taka per ton (dry peat). Whereas the prices of equivalent amount of Kerosene and fire wood are 2786 taka and 1680 taka respectively. This indicates that peat can offer far cheaper price than those of other fuels.

Power stations using fluidized bed combustion of peat are successfully working in Russia, Finland and Ireland (8). Howev er, the structure of Bangladeshi peat is collidal and its ash content is high and variable, and so its burning characteristics cannot be predicted from the results obtained in the above mentioned cases. Hence it is felt that the study of burning characteristics of local peat will be a valuable contribution on its effective utilization.

CHAPTER-2 LITERATURE SURVEY

2. LITERATURE SURVEY

2.1 PRINCIPLES OF COMBUSTION:

2.1.1 COMBUSTION PROCESS

By 'Combustion' is meant the combination of a material with oxygen accompanied by heat release at a temperature high enough to give useful heat. Under suitable circumstances many elements will combine with oxygen, but for practical purposes the elements met with in commercial fuels are carbon, hydrogen and their various compounds known as hydrocarbons. In all solid fuels carbon is the largest single component.

Primary aims in combustion control are:

i) To burn the fuel completely.

ii) To burn it in the minimum possible quantity of air.

In order to burn a fuel completely it must be raised to the ignition temperature and kept at that temperature until combustion is complete. To burn it in minimum air, the air and fuel must be mixed intimately by setting up turbulant conditions in the combustion chamber. These requisites are known as "Time" "Temperature" and "Turbulence" or the "Three T's".

Peat is a high volatile fuel and it must be burned under conditions suitable to such fuels. Although the ignition temperature is low, this does not mean that complete combustion can take place at very low temperatures. As soon as the peat ignites distillation of volatiles commences and the ignition temperature of these volatiles is much higher than that of the solid peat. The presence of water vapour and a large amount of excess air will render the ignition temperature even more difficult to reach. If these volatiles do not burn they invariably deposit in flues and chimneys. Thus excessive chimney deposits can follow from the use of wet peat and even from the use of quite good fuel in badly designed appliances. In order to avoid these conditions, the principle of the three "Ts" must be fully applied.

Time: The combustibles must spend a sufficient time under the proper conditions of temperature and Turbulence to allow completion of the burning reaction.

Temperature: The combustibles must be kept over the ignition temperature until burning is firmly established. The volatiles in peat, however, will be distilled off at temperatures lower than their ignition temperature and care must be taken to distribute the available heat so that they will burn completely.

Turbulence: In many cases, although more than enough air for complete combustion is supplied, unburnt combustibles are lost up the chimney. This is due very often to absence of the foregoing conditions of time and temperature but is also commonly a result of "Stratification". This means that the combustion air and the rising volatiles pass through the plant in separate streams without mixing. Steps should be taken to break these streams and cause the unburnt gases and air to mix efficiently, or in other words to set up turbulent conditions.

2.2 SCIENCE OF SOLID FUEL COMBUSTION:

Most of the practical combustion phenomena belong to one of the following three categories:

i) Phenomena which are primarily controlled by chemical kinetics.

ii) Phenomena which are primarily controlled by diffusion.

iii) Phenomena in which the roles played by chemical kinetics and physical mixing are more or less of equal importance.

Ignition, explosion, extinction and quenching of flames serve as examples of phenomena in the first category. The burning of a gaseous fuel jet, of a liquid fuel spill, spray, or drop, of a carbon sphere and of a candle illustrate the diffusionally controlled combustion phenomena. Flames in a gasoline engine, a Bunson burner and other situation in which the fuel and oxident are premixed belong to the third category (12).

Some solids are distilled when subjected to heat. Examples are coal, wood, paper, cotton etc. These solids disintegrate irreversibly in response to heat. Products of such disintegration are combustible volatile gases and a carbonaceous solid residue. This is known as PYROLYSIS and the solids undergoing such disintegration are called PYROLYZING or CHARRING solids. Once the volatiles of these are released, the carbonaceous solid residue is attacked by oxygen in much the same manner like a simple solid such as carbon.

2.2.1 NUSSELT'S SHRINKING DROP THEORY FOR PYROLYZING SOLIDS:

As shown in figure 2.2.1, a sphere of a pyrolyzing solid fuel such as wood or coal. Its diameter d=2R. Let to be its initial uniform temperature. At time equal to zero, let the surface (12) temperature of the solid is abruptly raised to T_{α} and held these. Heat then flows into the interior of the solid by conduction to raise the temperature of any portion continuously with

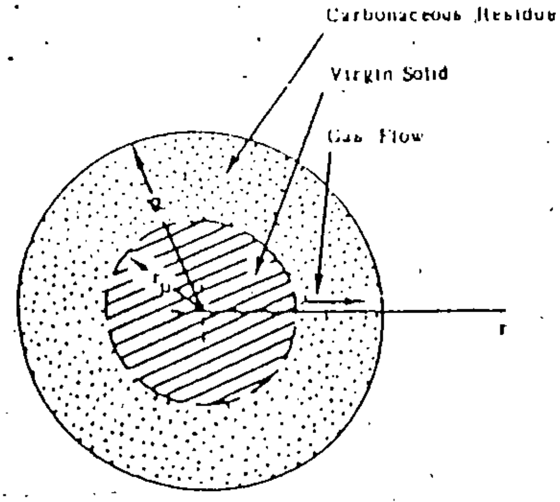


Figure 2.2.1: Nusselt's shrinking drop model.

time. The internal temperature history is governed by the following transient conduction equation and conditions:

$$\frac{\partial}{\partial r}\left(K'_s r^2 \frac{\partial T}{\partial r}\right) = \rho_s C_s r^2 \frac{\partial T}{\partial t} \quad \ldots \quad (2.2.1)$$

$$t < 0 \quad T = T_o \quad \text{in } 0 \leq r \leq R$$

$$t \geq 0 \quad T = T_\alpha \quad \text{at} \quad r = R \quad \ldots \quad (2.2.2)$$

$$\frac{\partial T}{\partial r} = 0 \text{ at} \quad r = 0$$

Solution T = T (r,t) of Eqs. 2.2.1 and 2.2.2 is

$$\frac{T(r,t) - T_o}{T_\alpha - T_o} = \frac{2R}{\pi r} \sum_{n=1}^{\alpha} \frac{(-1)^n}{n} \sin \frac{n\pi r}{R} \exp\left[\frac{-n^2 \pi^2 \alpha_s t}{R^2}\right] \quad (2.2.3)$$

Where α_s is the thermal diffusivity of the solid. Nusselt assumed that at any station solid releases its volatiles abruptly when it reaches a temperature equal to a characteristic pyrolysis temperature T_p. The pyrolysis front i.e., the radius at which $T = T_p$ continously progresses towards the origin of the sphere as time increases. The rate of volatile production, equal to the rate of weight loss of the sphere, is given by the speed of inward propagation of the pyrolysis isotherm $T = T_p$.

$$W(t) = (\rho_s - \rho_c)\ 4\pi\left(r^2 \frac{\partial r}{\partial t}\right)\Bigg|_{T=T_p} \qquad \ldots \qquad (2.2.4)$$

r^2 and $\left(\frac{\partial r}{\partial t}\right)\Bigg|_{T_p}$ are obtained from equation (2.2.3) by setting $T = T_p$, a constant. Such a calculation yields the following relation between time-averaged pyrolysis rate and sphere diameter:

$$W = K_p W_o d^{-2} \qquad \ldots \qquad (2.2.5)$$

where W_o is the initial weight of the sphere and $d = 2R$. K_p is the pyrolysis constant. Its magnitude for cellulosic substances lies between 0.005 and 0.01. Its unit is cm^2/Sec.

It is a function of thermal properties of the solid, the pyrolysis temperature T_p, the surface temperature T_α and the solid initial temperature T_o.

2.2.2 KINETIC MODELS FOR CARBON COMBUSTION:

The following discussions refer to heterogeneous reactions that take place at the surface of a carbon particle and it will be assumed that carbon particle is impervious to diffusion. Certainly this is not the general case. But when the objective is to separate out the gas phase effect the assumption is quite acceptable.

Any heterogeneous combustion reaction over a non porous solid surface involves the following six steps in series (13).

1. Diffusion of gases to the reaction surface.
2. Adsorption of the gaseous reactant.
3. Reaction between the solid and gaseous reactants.
4. Desorption of the gaseous products from the surface.
5. Diffusion of the products away from the surface.
6. Homogeneous gas phase reaction

These steps occur in series, the slowest of them determines the overall rate. In the case of carbon combustion, steps 2 and 4 are known to be extremely fast. When the particle temperature is low, the particle is small and the flow around it is feeble, step 3 is known to be much slower than step 1 or step 5. The burning rate then is determined by the chemical kinetics and therefore the process is kinetically controlled. In kinetically controlled regime, the burning rate depends upon the temperature exponentially. Since the process of diffusion is a function of particle size and flow, and since it is irrelevant in the kinetic regime, the burning rate is independent of the particle size and flow around it.

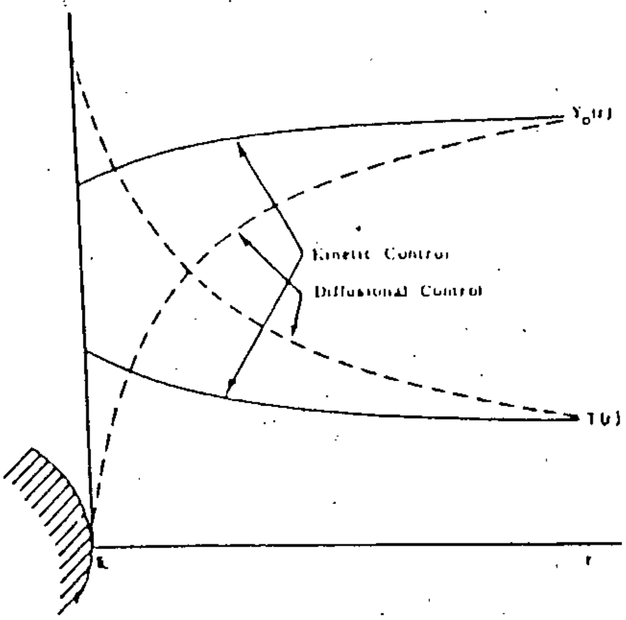


Figure 2.2.2: Kinetically controlled and diffusionally controlled combustion of a solid particle.

When the particle and flow velocities are large and temperature is high, step 3 is known to be much faster than step 1 and step 5. The burning rate is then controlled by the diffusion rate of oxygen to the particle. In the diffussionally controlled regime, the burning rate depends weakly ($T^{0.5-1.0}$) on temperature and strongly on the particle size.

The oxygen concentration at the reacting surface is negligibly small. This description is also illustrated qualitatively in Figure (2.2.2) & (2.2.3).

Diffusional combustion of a carbon sphere depends upon the extra variable of stoichiometry. Three possibilities arise:

Firstly, at moderately high temperatures, oxygen diffuses to the carbon surface and reaction occurs to form CO_2.

$$C + O_2 \longrightarrow CO_2 \quad (f = 12/32) \quad \ldots. \qquad (2.2.6)$$

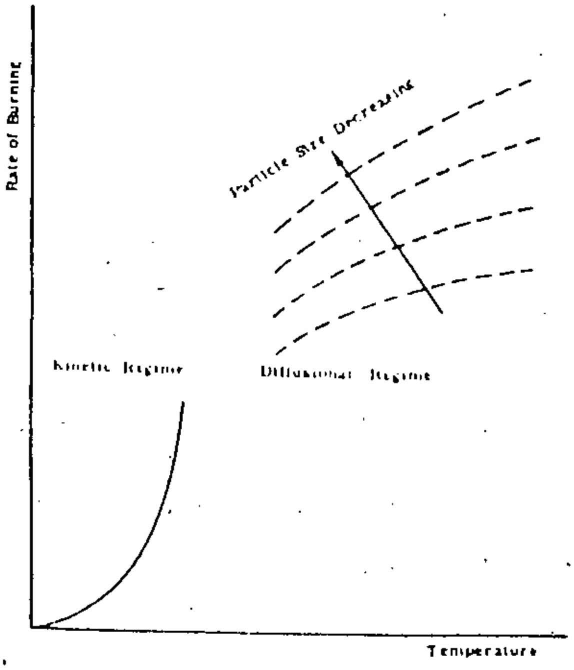


Figure 2.2.3: Rate of burning Vs temperature for the kinetically and diffusionally controlled regimes.

The fuel oxygen ratio, f, for this mechanism is 12/32. The product CO_2 diffuses away from the surface. Figure 2.2.4. shows the expected profiles of Y_o, Y_{co_2} and T for this scheme.

Secondly, when there is no free oxygen present, the reaction between O_2 and carbon at the surface might yield a richer oxide, viz: CO, instead of the leaner oxide.

$$2C + O_2 \longrightarrow 2CO \quad (f = 12/16) \qquad .. \qquad (2.2.7)$$

The stoichiometric fuel oxygen ratio for this mechanism is naturally higher, f = 12/16.

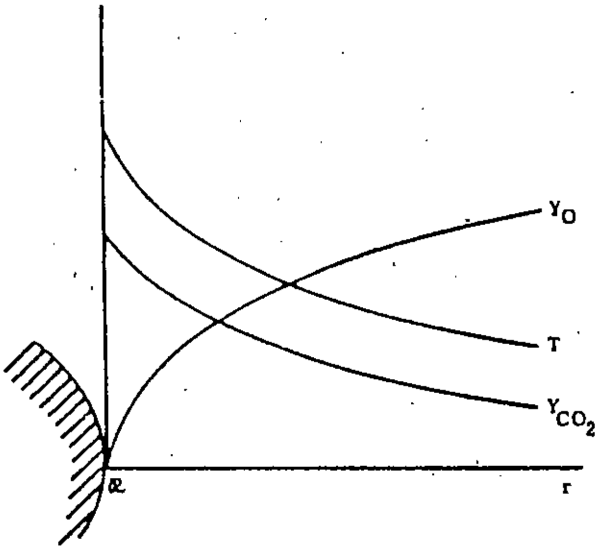


Figure 2.2.4: Combustion of carbon according to $C + O_2 \rightarrow CO_2$.

The third mechanism, which occurs when the temperature is high, is one provoked by experimental observations. Coffin (19) has conducted experiments which showed that there is negligible amount of both CO_2 and O_2 present in the gas phase near the carbon surface and that the temperature and CO_2 profiles exhibit a maximum at some distance away from the surface in the gas phase. Experiments of Hottel and Davis (15) show a definite gas phase reaction in which carbon monoxide is thought to be burnt by the counter-diffusing oxygen to yield a thin, weakly visible, bluish, envelope flame separated from the carbon surface by a distinct dark zone. A realistic mechanism of carbon burning will therefore be as follows: Carbon monoxide diffuses from the carbon surface and burns in gas phase as it meets the oxygen. Part of the carbon dioxide produced in this reaction diffuses back to the carbon surface where it is reduced to CO by the carbon.

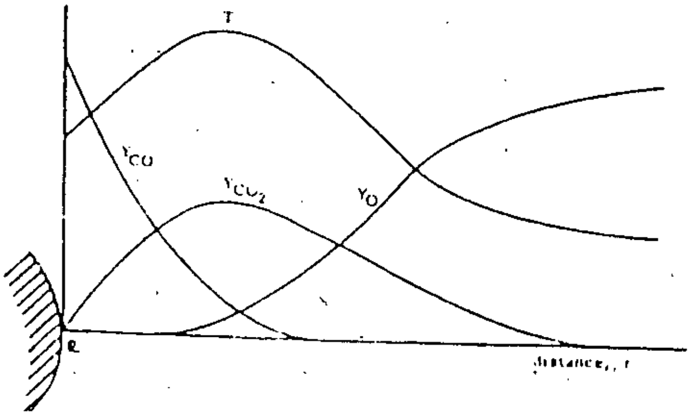


Figure 2.2.5: Combustion of carbon according to Coffin's hypothesis.

The distribution patterns of T, Y_o, Y_{co} and Y_{co_2} are then similar to those of liquid fuel combustion as shown in Figure (2.2.5).

2.3 COMBUSTION OF SMALL PARTICLES OF SOLID FUELS:

2.3.1 RESTRICTED MODEL OF A BURNING SPHERICAL PARTICLE:

Let us consider a restricted model (16) of a burning solid fuel particle as illustrated in Figure 2.3.1. This model is very similar to Nusselt's Model. A spherical carbon particle is assumed to be surrounded by a boundary layer of stagnant gas several particle diameters thick, through which oxygen diffuses from the free stream, reacting at the surface to form products which diffuse outwards to pass into the free stream. The products may be a mixture of carbon

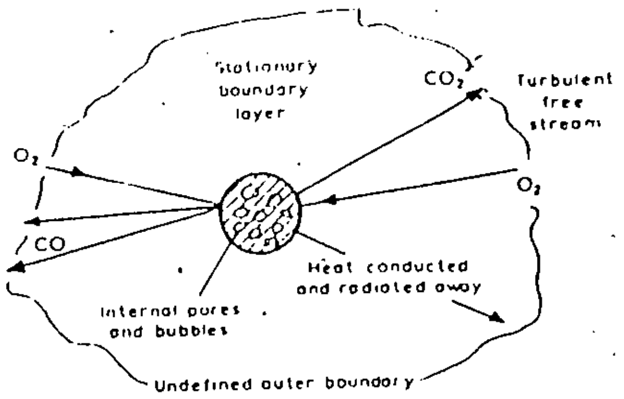


Figure 2.3.1: Restricted model of burning particle.

monoxide and carbon dioxide, but it will be assumed that the oxidation of carbon monoxide takes place entirely either very close to the surface, in which case the product is effectively carbon dioxide, or alternativ ely in the free stream.

In the boundary layer, on these assumptions, oxygen is conserved and therefore the rate at which oxygen is transported through an imaginary surface at any radius is equal to the rate at which oxygen is transported to the particle surface:

$$4\pi R^2 G(R) = 4\pi r^2 G(r) \qquad \ldots \qquad (2.3.1)$$

where

r = radial distance from the centre of particles, cm

R = radius of particle, cm and

G(r) = Flux of oxygen at radius r, gm/cm^2.s

The flux of oxygen is given by the diffusional transport equation:

$$G(r) = \frac{DM}{R'T} \cdot \frac{dp}{dr} \qquad \ldots \qquad (2.3.2)$$

where,

D = diffusional co-efficient of oxygen in the gas cm^2/s.

M = Molecular weight of oxygen.

R' = gas constant (= 82.06 atm cm^3/mole ^{o}k)

T = gas temperature, ^{o}k, and

P = Partial pressure of oxygen, atm.

Here convective transport due to bulk flow is ignored. A bulk flow may exist, especially if the carbon is transported away as carbon monoxide, because there is then a net molar outflow, but this bulk flow represents only a small correction to the diffusional equations. Beside the more obvious reason for a bulk flow there is the less obvious Stefan flow (17), discussed by Frank-Kamenetskii (1955), which arises if the transport of each species is calculated as though it diffused independently. The Stefan flow is perhaps less a physical reality, and more a mathematical convenience to overcome discrepancies which arise from the use of diffusion equations that are only approximately correct. For practical purposes where the bulk of the gas is the inert diluent nitrogen, introduced in the combustion air, both of these flows represent only small corrections to the calculations.

Substituting for the oxygen flux in Equation 2.3.1 yields:

$$R^2 G(R) = \frac{DM}{R'T}\, r^2 \, \frac{dp}{dr} \qquad \ldots \qquad (2.3.3)$$

In order to integrate this equation exactly it would be necessary to take account of the variation of temperature, and thus also of the diffusion co-efficient, with radial distance. However, it is more convenient to neglect this radial variation and assume a mean value T_m for the temperature in the boundary layer and a corresponding value for the diffusion co-efficient. Equation 2.3.3 may then be integrated between any radius r and the surface to give:

$$R.G(R)\ \frac{R'T_m}{DM}\ (1-R/r) = P(r) - P(R) \quad \ldots. \qquad (2.3.4)$$

Far from the particle, i.e. at large values of r/R, R/r approaches zero and the oxygen concentration tends to the value in the free stream, giving:

$$R.G(R)\ \frac{R'T_m}{DM} = P_g - P_s \quad \ldots \qquad (2.3.5)$$

Where P_g and P_s represent the oxygen partial pressures in the free stream and at the surface.

This equation gives the rate of transport of oxygen to the particle surface in terms of the concentration of oxygen in the free stream and at the surface. The ratio of carbon assumed to oxygen transported to the surface depends on the product which is transported away. For generality the relation may be written:

$$q = \frac{12\phi\ G(R)}{M} \quad \ldots \qquad (2.3.6)$$

where,

q = Rate of consumption of carbon, gm/cm^2.s and ϕ is a mechanism factor which takes the value 1 when carbon dioxide is transported away and 2 when carbon monoxide is transported away.

Combining Equations (2.3.5 and 2.3.6) yields an expression for the rate of consumption of carbon:

$$q = 24\phi D(P_g - P_s)/ dR'T_m = k_{diff}(P_g - P_s) \quad \ldots \quad (2.3.7)$$

where,

d = 2R = particle diameter, cm, and

$$K_{diff} = 24\phi D/dR'T_m \quad .. \quad (2.3.8)$$

There is an im portant special case of Equation (2.3.7) which is worth noting : that of diffusion control. If the surface reaction rate co-efficient is sufficiently high, then the oxygen partial pressure at the surface may be small compared with the value in the free stream. The reaction is then said to be diffusion controlled and the rate of reaction is giv-en by setting P_s=0 in Equation (2.3.7).

$$q_{diff} = k_{diff}P_g \quad \ldots \quad (2.3.9)$$

K_{diff} is called the diffusional reaction rate co-efficient. It depends on the particle diameter, the mechanism factor and the mean temperature, but is completely independent of fuel type. Combining Equations (2.3.7) and (2.3.9) yields

$$\frac{q}{q_{diff}} = \frac{P_g - P_s}{P_g} \qquad \ldots \qquad (2.3.10)$$

After simplification, Equation 2.3.10 becomes.

$$P_s = P_g - \frac{q}{q_{diff}} P_g \qquad \ldots \qquad (2.3.11)$$

A part from this special case, the partial pressure of oxygen at the surface has to be found in order to evaluate the reaction rate from Equation (2.3.7). This involves additionally the kinetics of the surface reaction, including internal reaction if any, and is assumed here to be given by:

$$q = k_s p_s \qquad \ldots \qquad (2.3.12)$$

where k_s is the surface reaction rate co-efficient.

The partial pressure at the surface may thus be eliminated between Equation (2.3.7) and (2.3.12) to give:

$$q = \frac{P_g}{(1/k_{diff}+1/k_s)} \qquad \ldots \qquad (2.3.13)$$

The overall reaction rate depends on both k_{diff} and k_s but tends to be dominated by whichever is the smaller. Thus the diffusional term tends to dominate for large particles and at high temperatures, while the chemical term, as it is often called, tends to dominate at low temperatures and for small particles or highly unreactive fuels.

Given the particle size, the gas temperature, the particle temperature, the combustion mechanism factor and the oxygen concentration in the free stream, the reaction rate may be calculated from Equation (2.3.13).

2.3.2 CHOICE OF MECHANISM FACTOR (ϕ)

The fact that twice as much oxygen is required to burn carbon to form carbon dioxide as to form carbon monoxide means that in conditions approaching diffusion control the reaction rate is uncertain by a factor of two ($\phi = 2$) if the combustion product is not known. The question is not only which product is formed, but also, if carbon monoxide is formed, where the subsequent oxidation of the carbon monoxide takes place. This is due to the fact that if this occurs close to the surface, then just as much oxygen has to be transported to the surface as if the actual product had been carbon dioxide. No firm conclusion can be reached regarding this question; but calculating the reaction rate that for small particles it is probably reasonable to assume that carbon monoxide is formed, and that the subsequent oxidation takes place mainly outside what may be considered to be the boundary layer to the particle.

The larger the particle, the less likely is it that the carbon monoxide will escape from the boundary layer without being burnt. The reaction rate is such that a large proportion of the carbon monoxide is burnt within the boundary layer, but even for particles as large as 100 micron it would still not be reasonable to assume that the combustion of carbon monoxide takes place close to the surface. Such an assumption is probably only valid for particles larger than 1 mm. Therefore mechanism factor, ϕ can be taken to be unity for particles, whose diameter is larger than 1 mm.

2.4 RESEARCH ACTIVITIES ON PEAT IN BANGLADESH

There have been some work by different persons on the utilization of locally available peat. Short descriptions of their activities are as follows:

A study of burning characteristics of different sizes of locally available peat was carried out by Farooq(9). The study was mainly related to generate rate data on peat particles (d=0.5-2.21 cm), collected from Demra, Kishoreganj and Kola mouza in static environment at 828° - 914°C. Most of the experiments were carried out in a Muffle furnace. The main findings of his study were as follows+

i) Peat ash retains its size and structure fafter burning to completion in static environment.

ii) Since peat ash retains size and structure, diffusion of oxygen into the solid becomes an added resistance in the process of fixed carbon burning. Hence, burning of peat in large lamps will be difficult.

iii) Burning rate of peat decreases with time. Rate of Pyrolysis is much higher than the burning rate of fixed carbon in peat.

iv) For a definite interval burning rate of peat increases with decreasing particle size.

Some experiments were also carried out(9) on the combustion of peat particle in a fluidized bed operating at atmospheric pressure and 943°C. During these investigations, it was observed that the burnt peat ash was not friable and the ash particle accumulated in the bed. With increased addition of peat particles the bed ceased to fluidize.

Begum and Hossain (26) carried out systematic study on characterization of peat available in Kola Mouza, Chanda Beel and Baghia Beel areas. The results obtained led to the following conclusions.

i) Bangladeshi peat is marsh, swamp type and highly decomposed. The colour is in between brown and dark. It is not hygroscopic and does not catch fire in contact with air.

ii) Bangladeshi Peat is poor raw material for chemical processing because of high ash content.

iii) The peats of Baghia and Chanda Beels are not graphitizable.

CHAPTER-3 PROGRAMME OF RESEARCH

3. PROGRAMME OF RESEARCH:

3.1 OUTLINE OF APPROACH

At present Bangladesh is suffering from twin energy problem. A major share of our export earnings is being spent every year for the import of commercial fuels. On the other hand traditional fuels (fire wood, agricultural residues, cow dung etc) are presently being consumed at a much higher rate than their regenerative supply rate causing environmental degradation in the form of deforestation and loss of organic matter of the soil.

Bangladesh has a considerable reserve of peat in the districts of Faridpur and Khulna. A number of smaller deposits also exist in other parts of Bangladesh. The total estimated peat reserve in Faridpur and Khulna is about 133 million tons (8). If the smalleer deposits are included the estimated figure rises to about 600 million tons (8).

Rational use of available peat resources has an important role to play in meeting the total energy demand of the country. Preliminary techno-economic feasibility study for mining of the peat deposits of Kola Mouza which was carried out by Bangladesh Mineral Exploration and Development Corporation (BMEDC) during the period from February, 1979 to July 1980, had proved the economical viability of peat extraction. At present BOGMC (formerly BMEDC) is trying to establish pilot plants for briquette manufacturing and power generation (10-15) MW capacity) with the help of foreign organisations. An agreement has already been signed between BOGMC (Bangladesh Oil, Gas and Mineral Corporation) and CIDA (Canadian International Development Authority) regarding feasibili ty study of peat power generation and the designing of a 10-15 MW Power Plant (7).

It is well known that the rate of combustion of solid fuels depends on heat transfer and mass transfer rates, rate of devolatilisation, burning rate of volatile matters and of fixed carbon. In accordance with the literature (19), the carbon burning rate is the slowest among the above mentioned rate. Preliminary investigations on peat sphere burning have also proved that rate of carbon burning is the limiting stage (Sec. Appendix-G).

Known expressions for heat and mass transfer, devolatilization and volatile matter burning rate used in coal burning rate may be used in the case of peat burning with minor change using the physical and thermodynamic constants of peat. On the other hand rate of burning of solid carbon depends largely on carbon type. This is why, rate expression for carbon burning should be developed first. Expressions of mass transfer, heat transfer, devolatilization and volatile matter burning rates in combination with developed solid carbon burning expressions will give the complete model of peat particle combustion rate.

3.2 OBJECTIVES OF PRESENT RESEARCH

Considering the points mentioned in section 3.1 of this present work the following objectives were set:

i) Design and fabrication of experimental set-up to study combustion phenomena of devolatilised peat particles.

ii) A study of variables affecting combustion of develatilised peat particles. The factors are:

   a) Devolatilised peat particle diameter.

   b) Air flow rate.

   c) Temperature.

   d) Reaction time.

CHAPTER-4 EXPERIMENTAL SET-UP

4. EXPERIMENTAL SET-UP

4.1 GENERAL ARRANGEMENT OF THE EXPERIMENTAL SET-UP

A flow diagram of the experimental set up is shown in Figure 4.1 alongwith a key. The different parts of the reactor and the sample carrier hav e been shown in Figures 4.2 and 4.3. Plates 4.1, 4.2 and 4.3 present the salient features of the experimental set up, the test section and the control pannel. Brief description of the main parts of the experimental set-up is as follows:

4.1.1 REACTOR

The reactor is a stainless steel tube 25.40 mm in diameter and 0.85 metere long. The top end is jacketed and cold compressed air is passed through this jacket for cooling this part of the reactor. There are two inlets to the reactor, one for air and the other for nitrogen. The details of the reactor are shown in Figure 4.2. The reactor except the jacketed part is placed in a tubular furnace during the experimental runs.

4.1.2 THE SAMPLE CARRIER

The sample carrier/container consists of a hemispherical basket made of a highly perforated net of stainless steel suspended from a suspends made of 21 milli metre diameter stainless steel rod as shown in Figure 4.3. The upper end of the reactor is closed by the threaded end of sample carrier. The sample carrier assembly can be strewed into the reactor.

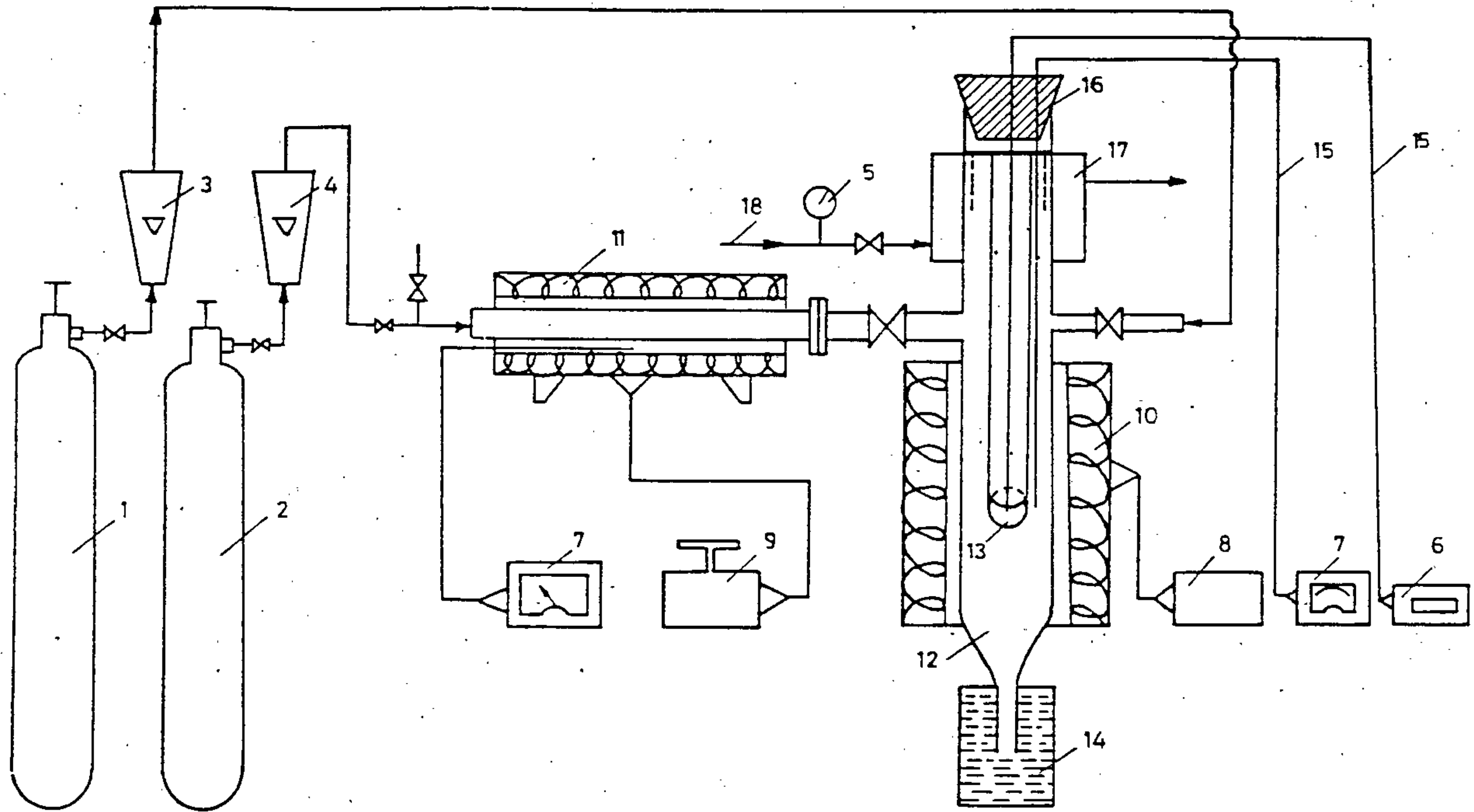
4.1.3 THE PREHEATER

The preheater is 25.40 mm diameter and 0.80 metre long stainless steel tube placed in a tubular electrical furnace. The preheater is connected to reactor and rotameter with the help of standard couplings.

4.2 SAMPLE PREPARATION FOR EXPERIMENTS

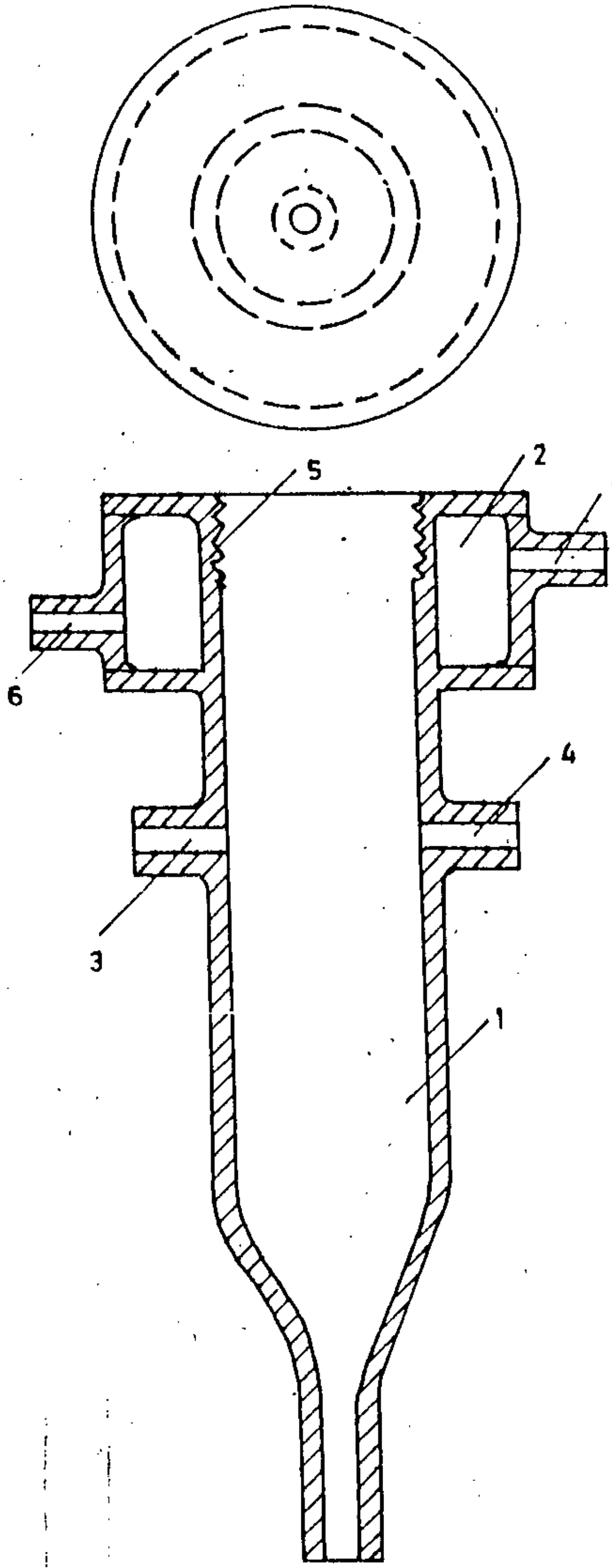
A) Method of drying: Peat spheres of different sizes were prepared from moist peat using equal weight for each size and they were dried in air. After air drying, the spheres were dried in an oven at 100-105°C upto constant weight.

B) Method of devolatilization: Oven dried peat spheres were kept in highly perforated net and purged for half an hour in a stream of cold oxygen-free nitrogen one at a time. Then the temperature of the reactor was raised upto 900°C under a flow of nitrogen and kept for 10 minutes at that temperature. After elapse of 10 minutes the peat sphere was cooled to 100°C in a stream of nitrogen and then the devolatilised peat sphere was removed from the reactor. The retention time of peat spheres at 900°C was selected as 10 minutes because it was experimentally found that the time required for complete devolatilisation of the largest sphere among the peat spheres used in experiments, was 10 minutes.



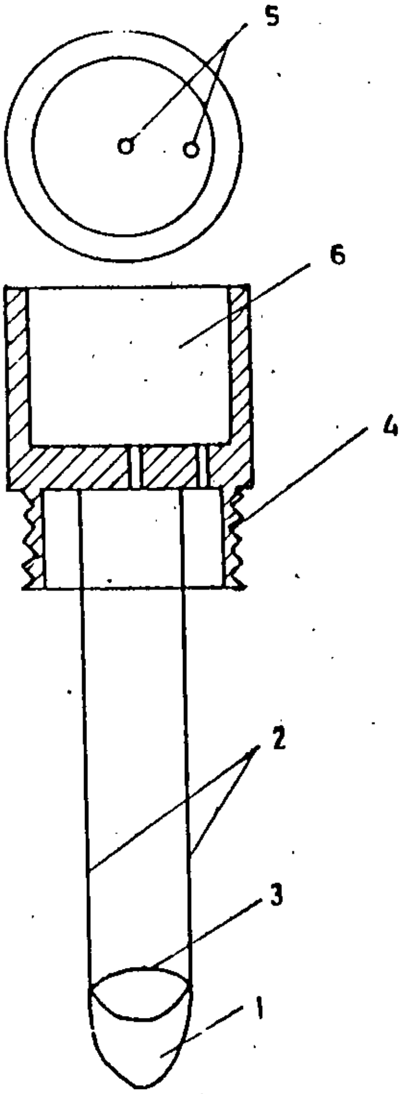
1. DRY NITROGEN CYLINDER.
2. DRY AIR CYLINDER.
3. NITROGEN ROTAMETER.
4. AIR ROTAMETER.
5. PRESSURE GAUGE.
6. TEMPERATURE INDICATOR.
7. MILLI VOLT RECORDER.
8. TEMPERATURE CONTROLLER.
9. VARIAC.
10. REACTOR FURNACE.
11. AIR PREHEATER.
12. REACTOR.
13 SAMPLE CARRIER.
14 WATER POT.
15 THERMOCOUPLES.
16 RUBBER CORK.
17 COOLING AIR JACKET.
18 COOLING AIR LINE.

FIG. 4.1 FLOW DIAGRAM OF EXPERIMENTAL SET-UP.



1. STAINLESS STEEL TUBE.
2. AIR JACKET.
3 AIR INLET TUBE OF REACTOR.
4. NITROGEN INLET TUBE OF REACTOR.
5. THREAD FOR FIXING SAMPLE CARRIER
6. COOLING AIR INLET LINE OF JACKET.
7. COOLING AIR OUTLET LINE.

FIG. 4.2 THE REACTOR ASSEMBLY



1. STAINLESS STEEL NET.
2. STAINLESS STEEL RODS.
3. RING FOR NET HANGING.
4. THREAD FOR FIXING SAMPLE CARRIER.
5. HOLES FOR THERMOCOUPLES.
6. HOLES FOR RUBBER CORK.

FIG. 4.3 THE SAMPLE CARRIER ASSEMBLY.

CHAPTER-5 METHODOLOGY & EXPERIMENTAL DATA

5. METHODOLOGY AND EXPERIMENTAL DATA.

Before carrying out experimental runs with the set-up a few trial runs were made to observe the performance of the experimental set-up. The calibration of the different instruments were checked. The calibration procedure of thermocouple has been presented in Appendix A. The experimental data have been furnished in Tables 5.1 to 5.4. The proximate analysis of peat used in the present study is shown in Table B-1, Appendix-B. Similar analysis carried out by previous investigators(26) have also been presented in Table B-1 for comparison.

5.1 Experimental run

Burning rate study of the devolatilised peat particles were carried out in the stainless steel reactor heated by the tubular furnace. The schematic flow diagram of the experimental set-up has been shown in Figure 4.1 and Plate 4.1. After placing the devolatilised spheres in highly perforated net, the reactor was purged by cold oxygen - free nitrogen for half an hour. Then the reactor was heated upto reaction temperature under slow stream of nitrogen. When steady reaction temperasture attained, the nitrogen inlet valve was closed and air inlet valve of the reactor opened quickly and moisture free preheated air was introduced into the reactor for a definite period of time. At the end of the specified time period, the reaction was not allowed to proceeded any further by interrupting the air flow and the peat particles was cooled in a flow of nitrogen upto room temperature. After being cooled peat particles was taken out of the reactor.

Air and nitrogen flow rates were measured by calibrated rotameters at ambient conditions. Air temperature and surface temperature of peat particle were measured with the help of thermocouples. Temperatures of the reactor furnace and preheator was controlled by temperature controller and variac respectively. Weights of the experimental peat particles were measured in an analytical balance.

TABLE NO. 5.1: EXPERIMENTAL SPECIFIC REACTION RATES OF DEVOLATILISED PEAT PARTICLES AT 60 1/h AIR FLOW RATE.

5.1.1 Particle diameter = 0.42 cm.

Mean Temperature of Peat Particle, T_m (°C)	Initial weight of Peat Particle, W_o (mg)	Final weight of Peat Particle, W_1 (mg)	Specific Reaction Rate of Peat Particle, q (mg/cm^2.s)
461	42.50	38.50	0.123
516	42.00	34.50	0.231
561	43.10	35.70	0.226
620	42.10	29.00	0.404
661	45.00	32.00	0.400
712	42.00	25.40	0.513
736	44.00	28.60	0.473
760	43.20	29.80	0.412
808	43.00	29.10	0.427

5.1.2 Particle diameter = 0.54 cm

Mean Temperature of Peat Particle, T_m (°C)	Initial weight of Peat Particle, W_o (mg)	Final weight of Peat Particle, W_1 (mg)	Specific Reaction Rate of Peat Particle, q (mg/cm^2.s)
459	74.20	68.20	0.110
506	76.40	66.80	0.175
563	72.80	56.30	0.300
604	77.20	59.70	0.318
664	73.40	51.70	0.394
709	78.00	54.10	0.435
735	75.30	50.20	0.455
757	71.50	45.80	0.468
804	74.20	47.30	0.488

5.1.3 Particle diameter = 0.68 cm.

Mean Temperature of Peat Particle, T_m (°C)	Initial weight of Peat Particle, W_o (mg)	Final weight of Peat Particle, W_1 (mg)	Specific Reaction Rate of Peat Particle, q (mg/cm^2.s)
470	185.00	173.90	0.127
528	178.20	160.60	0.202
577	177.40	157.40	0.230
641	179.40	145.60	0.388
680	180.40	143.30	0.426
730	182.00	138.80	0.96
754	183.20	148.80	0.395
769	185.00	146.60	0.441
821	183.40	139.00	0.509

5.1.4 Particle diameter = 0.88 cm.

Mean Temperature of Peat Particle, T_m (°C)	Initial weight of Peat Particle, W_o (mg)	Final weight of Peat Particle, W_1 (mg)	Specific Reaction Rate of Peat Particle, q (mg/cm^2.s)
445	395.30	387.90	0.051
512	400.00	378.00	0.151
568	402.86	377.00	0.177
612	409.00	379.80	0.200
653	407.30	371.20	0.247
703	398.30	352.90	0.311
738	403.30	357.00	0.317
762	397.80	349.60	0.330
809	401.40	342.40	0.404

TABLE NO. 5.2: EXPERIMENTAL SPECIFIC REACTION RATE OF DEVOLATILISED PEAT PARTICLE AT 125 l/h AIR FLOW RATE.

5.2.1 Particle diameter = 0.42 cm.

Mean Temperature of Peat Particle, T_m (°C)	Initial weight of Peat Particle, W_o (mg)	Final weight of Peat Particle, W_1 (mg)	Specific Reaction Rate of Peat Particle, q (mg/cm^2.s)
462	41.30	36.30	0.153
518	43.00	35.40	0.235
567	42.40	31.90	0.325
622	40.80	27.30	0.415
667	42.10	27.40	0.453
717	45.00	28.50	0.509
742	43.10	25.90	0.531
767	44.30	26.70	0.542
817	41.40	23.10	0.563

5.2.2 Particle diameter = 0.54 cm.

Mean Temperature of Peat Particle, T_m (°C)	Initial weight of Peat Particle, W_o (mg)	Final weight of Peat Particle, W_1 (mg)	Specific Reaction Rate of Peat Particle, q (mg/cm^2.s)
459	74.20	66.60	0.139
512	71.50	60.90	0.193
579	75.30	61.00	0.260
613	78.00	62.10	0.289
666	73.40	54.70	0.340
713	77.20	56.80	0.372
739	72.80	51.50	0.388
762	76.40	54.40	0.400
811	74.20	50.60	0.429

5.2.3 Particle diameter = 0.68 cm.

Mean Temperature of Peat Particle, T_m (°C)	Initial weight of Peat Particle, W_o (mg)	Final weight of Peat Particle, W_1 (mg)	Specific Reaction Rate of Peat Particle, q (mg/cm^2.s)
465	183.40	172.20	0.128
520	185.00	169.80	0.175
578	183.20	163.00	0.232
627	182.00	159.30	0.260
673	180.40	154.70	0.295
722	179.40	151.30	0.323
747	177.40	147.40	0.344
766	178.20	146.80	0.360
816	185.00	151.20	0.388

5.2.4 Particle diameter = 0.88 cm

Mean Temperature of Peat Particle, T_m (°C)	Initial weight of Peat Particle, W_o (mg)	Final weight of Peat Particle, W_1 (mg)	Specific Reaction Rate of Peat Particle, q (mg/cm^2.s)
455	401.40	386.00	0.105
516	397.80	375.50	0.153
573	403.30	374.40	0.198
620	398.30	364.00	0.235
663	407.30	369.40	0.260
713	409.60	365.50	0.302
743	402.80	356.40	0.318
764	400.00	352.10	0.328
813	397.30	345.50	0.355

TABLE NO. 5.3 : EXPERIMENTAL SPECIFIC REACTION RATES OF DEVOLATILISED PEAT PARTICLES AT 180 l/h AIR FLOW RATE.

5.3.1 Particle diameter = 0.42 cm.

Mean Temperature of Peat Particle, T_m (°C)	Initial weight of Peat Particle, W_o (mg)	Final weight of Peat Particle, W_1 (mg)	Specific Reaction Rate of Peat Particle, q (mg/cm^2.s)
460	41.00	35.30	0.177
487	42.10	33.70	0.260
551	45.00	34.32	0.329
609	42.30	28.34	0.430
673	43.80	27.20	0.512
707	41.10	25.80	0.471
728	42.00	26.40	0.480
763	43.00	28.00	0.461
816	42.60	26.10	0.509

5.3.2 Particle diameter = 0.54 cm.

Mean Temperature of Peat Particle, T_m (°C)	Initial weight of Peat Particle, W_o (mg)	Final weight of Peat Particle, W_1 (mg)	Specific Reaction Rate of Peat Particle, q (mg/cm^2.s)
463	71.60	61.10	0.191
515	72.30	57.80	0.264
559	74.50	56.40	0.330
619	73.30	51.00	0.405
672	77.10	52.80	0.443
726	73.30	47.90	0.463
752	71.80	46.00	0.470
770	77.40	51.60	0.470
814	73.30	47.20	0.475

5.3.3 Particle diameter = 0.68 cm

Mean Temperature of Peat Particle, T_m (°C)	Initial weight of Peat Particle, W_o (mg)	Final weight of Peat Particle, W_1 (mg)	Specific Reaction Rate of Peat Particle, q (mg/cm^2.s)
465	182.60	166.30	0.187
515	179.30	161.00	0.210
569	184.00	167.30	0.192
621	177.40	142.20	0.404
670	175.80	142.80	0.379
730	178.20	132.40	0.526
748	182.00	146.10	0.412
774	180.20	142.80	0.429
815	181.30	141.80	0.453

5.3.4 Particle diameter = 0.88 cm

Mean Temperature of Peat Particle, T_m (°C)	Initial weight of Peat Particle, W_o (mg)	Final weight of Peat Particle, W_1 (mg)	Specific Reaction Rate of Peat Particle, q (mg/cm^2.s)
458	401.40	383.00	0.126
512	401.50	382.50	0.130
555	407.30	391.40	0.109
613	398.80	370.90	0.191
657	397.30	366.80	0.209
712	403.00	370.90	0.220
728	404.40	369.80	0.237
757	400.20	362.00	0.262
810	408.30	368.60	0.272

TABLE NO. 5.4: DATA FOR CALCULATING BURNING RATES OF PEAT PARTICLES VS. TIME AT 60 l/h AIR FLOW RATE.

5.4.1 Particle diameter = 0.54 cm.

Mean Temperature of Peat Particle, T_m (°C)	Initial weight of Peat Particle, W_o (mg)	Residence Time in Reactor, t (seconds)	Final weight of Peat Particle, W_1 (mg)
713	79.30	30	60.80
713	78.00	40	59.80
716	75.60	60	51.60
717	71.60	80	42.80
716	72.60	100	42.30
717	78.40	120	43.10
717	73.30	140	37.20
717	77.20	150	41.10

5.4.2 Particle diameter = 0.88 cm.

Mean Temperature of Peat Particle, T_m (°C)	Initial weight of Peat Particle, W_o (mg)	Residence Time in Reactor, t (seconds)	Final weight of Peat Particle, W_1 (mg)
710	408.30	30	379.80
710	398.20	45	364.90
715	400.00	60	359.10
715	397.40	75	357.40
709	395.80	90	358.50
715	403.30	120	356.50
709	401.20	180	342.20
720	402.40	540	228.80
720	402.20	590	228.50

CHAPTER-6 RESULTS

6. RESULTS

The results of present investigation have been summarized in this chapter. Mathematics involved in calculations have been explained in sections 6.1 to 6.5. Details of sample calculations have been given in appendix-C. Calculated results have been compiled in Table 6.1.1. to 6.5 and illustrated graphically in Figures 6.1.1 to 6.7.2.

6.1 Specific reaction rate

Smith and Gudmandsen (18) suspended 5 mm diameter spheres of electrode carbon in a furnace tube through which air was blown, preheated to the furnace temperature. The method consisted in changing the gas stream in the tube from nitrogen to air and back again after a short interval. They defined specific reaction rate as rate of carbon removal per unit external geometric surface area:

$$\text{Specific reaction rate, } q = \frac{W_o - W_1}{t \times S} \qquad \ldots \qquad (6.1.1)$$

This expression has been used in calculating the specific reaction rate of this research work because of the fact that experiments were conducted in accordance with the method of smith and Gudmundsen.

Burning rates calculated at different flow rates against temperature have been plotted in rectangular co-ordinates for peat particles of different average sizes. The curves are shown in Figures 6.1.1, 6.1.2 and 6.1.3.

6.2 Diffusion controlled reaction rate

Field, Gill and Morgan (19) developed a model for diffusion controlled reaction rate of Carbon sphere on the assumption that the spherical carbon particle is surrounded by a boundary layer of stagnant gas several particle diameters thick, through which oxygen diffuses from the free stream, reacting at the surface to form products which diffusion controlled reaction model is expressed as follows:

$$q_{diff} = \frac{(24 \Phi D)}{d R'Tm} \cdot P_g \qquad \ldots \qquad (6.2.1)$$

Wen and Chaung (20) developed another expression for diffusion controlled reaction rate for char - 0_2 reaction. Equation 6.2.2 represents the Wens model:

$$q_{diff} = 0.292 \quad \frac{(4.26)}{T_g} \; \frac{(T_g)^{1.75}}{1800} \quad \frac{P_g}{P_t d} \qquad \ldots \qquad (6.2.2)$$

It may be mentioned here that the same results has been obtained using any one of the above mentioned equations.

Many investigators (19, p.329-345) calculated q_{diff} using the following expression to allow for the effect of gas velocity:

$$q_{diff} = \frac{(24 \Phi D)}{d R' Tm} \cdot P_g \cdot f' \qquad \ldots \qquad (6.2.3)$$

In this research work q_{diff} has been calculated using Equation 6.2.3 and the results have been plotted against temperature in Figures 6.2.1, 6.2.2 and 6.2.3. (The actual values of $q_{diff.}$ are 2.5 times lesser than the values showed in Tables and Figures).

6.3 Surface reaction rate co-efficient:

According to Field (19) surface reaction rate co-efficient may be defined as:

$$k_s = \frac{q}{P_s} \quad \ldots \quad (6.3.1)$$

Again the expression, $\frac{P_g - P_s}{P_g} = \frac{q}{q_{diff}}$ ... (6.3.2)

is valid if mass transfer is not wholly dominant and no resistance of ash layer.

The above mentioned two conditions may be considered to be fulfilled upto 30 seconds after starting burning of peat spheres (d = 0.54 cm and d = 0.88 cm).

Ash layer resistance can be considered negligible because of the fact that burning of peat spheres confined mainly at the outer surface during the period of 30 seconds. Experimental results show that q_{diff} is greater than q under studied conditions. So, mass transfer resistance is not wholly dominant.

On the basis of above mentioned points the Equation 6.3.2 may be used for calculating P_s.

In this research work, k_s at 700°C has been calculated using Equations 6.3.1 and 6.3.2. The results have been tabulated in Table 6.5.

6.4 Overall reaction rate co-efficient

According to Field (19), overall reaction rate co-efficient can be determined using the following expression:

$$k_{ov} = \frac{q}{P_g} \quad \ldots \quad (6.4.1)$$

This Equation has been used for calculating overall reaction rate and the results of calculations are presented in Tables 6.1 to 6.3 and have been plotted against temperature in Figures 6.3.1, 6.3.2 and 6.3.3.

6.5 Diffusion controlled reaction rate co-efficient

Diffusion controlled reaction rate co-efficient has been determined using the Equation:

$$k_{diff} = \frac{q_{diff}}{P_g} \quad \ldots \quad (6.5.1)$$

Calculated data have been tabulated in Table 6.1 to 6.4 and graphically presented by Figures 6.4.1, 6.4.2 and 6.4.3. (The actual values of k_{diff} are 2.5 times lesser than the values showed in Tables and Figures).

6.6 Ash film diffusion constant

Ash film diffusion constant, k_{ash} has been calculated using the following Equation:

$$\frac{1}{k_{ov}} = \frac{1}{k_{diff}} + \frac{1}{k_s} + \frac{1}{k_{ash}} \quad \ldots \quad (6.6.1)$$

The result has been tabulated in Table 6.4.1 and 6.4.2. Figures 6.7.1 and 6.7.2 show graphical representation of k_{ash} Vs. time.

TABLE NO. 6.1: BURNING RATES AND RATE CO-EFFICIENTS OF PEAT PARTICLES AT 60 l/h AIR FLOWRATE

6.1.1 Particle diameter = 0.42 cm.

Mean Temperature of Peat Particle T_m (°C)	Specific Reaction rate, q (mg/cm^2.s)	Theoritical Diffusional Rate, q_{diff} (mg/cm^2.s)	Overall Rate co-efficient, k_{ov} (mg/cm^2.s. atm)	Diffusional rate co-efficient, k_{diff} (mg/cm^2.s. atm)
475	0.145	0.740	0.69	3.52
525	0.216	0.773	1.03	3.68
575	0.290	0.805	1.38	3.83
625	0.365	0.840	1.74	4.00
675	0.429	0.872	2.04	4.15
725	0.467	0.907	2.22	4.32
750	0.480	0.924	2.29	4.40
775	0.495	0.939	2.36	4.47
800	0.505	0.958	2.41	4.56

6.1.2 Particle diameter = 0.54 cm.

Mean Temperature of Peat Particle T_m (°C)	Specific Reaction rate, q (mg/cm^2.s)	Theoritical Diffusional Rate, q_{diff} (mg/cm^2.s)	Overall Rate co-efficient, k_{ov} (mg/cm^2.s. atm)	Diffusional rate co-efficient, k_{diff} (mg/cm^2.s. atm)
475	0.135	0.608	0.64	2.90
525	0.200	0.636	0.95	3.03
575	0.275	0.659	1.31	3.14
625	0.345	0.690	1.64	3.29
675	0.405	0.720	1.93	3.43
725	0.450	0.750	2.14	3.57
750	0.465	0.770	2.21	3.67
775	0.476	0.780	2.27	3.77
800	0.488	0.792	2.32	3.77

6.1.3 Particle diameter = 0.68 cm.

Mean Temperature of Peat Particle T_m (°C)	Specific Reaction rate, q (mg/cm^2.s)	Theoritical Diffusional Rate, q_{diff} (mg/cm^2.s)	Overall Rate co-efficient, k_{ov} (mg/cm^2.s. atm)	Diffusional rate co-efficient, k_{diff} (mg/cm^2.s. atm)
475	0121	0.500	0.58	2.38
525	0.190	0.521	0.91	2.48
575	0.265	0.542	1.26	2.58
625	0.334	0.568	1.59	2.71
675	0.390	0.594	1.86	2.83
725	0.435	0.620	2.07	2.95
750	0.451	0.630	2.15	3.00
775	0.465	0.640	2.21	3.05
800	0.472	0.652	2.25	3.11

6.1.4 Particle diameter = 0.88 cm.

Mean Temperature of Peat Particle T_m (°C)	Specific Reaction rate, q (mg/cm^2.s)	Theoritical Diffusional Rate, q_{diff} (mg/cm^2.s)	Overall Rate co-efficient, k_{ov} (mg/cm^2.s. atm)	Diffusional rate co-efficient, k_{diff} (mg/cm^2.s. atm)
475	0.078	0.420	0.37	2.00
575	0.140	0.440	0.67	2.10
575	0.185	0.460	0.88	2.19
625	0.225	0.480	1.07	2.29
675	0.266	0.501	1.27	2.39
725	0.305	0.522	1.45	2.49
750	0.325	0.536	1.55	2.55
775	0.355	0.541	1.69	2.58
800	0.388	0.558	1.85	2.66

TABLE NO. 6.2: BURNING RATES AND RATE CO-EFFICIENTS OF PEAT PARTICLES AT 125 1/h AIR FLOW RATE

6.2.1 Particle diameter = 0.42 cm.

Mean Temperature of Peat Particle T_m (°C)	Specific Reaction rate, q (mg/cm^2.s)	Theoritical Diffusional Rate, q_{diff} (mg/cm^2.s)	Overall Rate co-efficient, k_{ov} (mg/cm^2.s. atm)	Diffusional rate co-efficient, k_{diff} (mg/cm^2.s. atm)
475	0.167	0.874	0.80	4.16
525	0.257	0.948	1.22	4.51
575	0.338	0.990	1.61	4.71
625	0.408	1.027	1.94	4.89
675	0.467	1.045	2.22	4.98
725	0.514	1.110	2.45	5.29
750	0.534	1.130	2.54	5.38
775	0.551	1.149	2.62	5.47
800	0.558	1.172	2.66	5.58

6.2.2 Particle diameter = 0.54 cm.

Mean Temperature of Peat Particle T_m (°C)	Specific Reaction rate, q (mg/cm^2.s)	Theoritical Diffusional Rate, q_{diff} (mg/cm^2.s)	Overall Rate co-efficient, k_{ov} (mg/cm^2.s. atm)	Diffusional rate co-efficient, k_{diff} (mg/cm^2.s. atm)
475	0.155	0.795	0.74	3.79
525	0.210	0.863	1.00	4.11
575	0.252	0.901	1.20	4.29
625	0.300	0.934	1.43	4.45
675	0.345	0.951	1.64	4.53
725	0.385	1.010	1.83	4.81
750	0.400	1.029	1.91	4.90
775	0.410	1.045	1.95	4.98
800	0.420	1.070	2.00	5.10

6.2.3 Particle diameter = 0.68 cm.

Mean Temperature of Peat Particle T_m (°C)	Specific Reaction rate, q (mg/cm^2.s)	Theoritical Diffusional Rate, q_{diff} (mg/cm^2.s)	Overall Rate co-efficient, k_{ov} (mg/cm^2.s. atm)	Diffusional rate co-efficient, k_{diff} (mg/cm^2.s. atm)
475	0.140	0.636	0.67	3.03
525	0.185	0.691	0.88	3.29
575	0.219	0.721	1.04	3.43
625	0.255	0.748	1.21	3.56
675	0.294	0.762	1.40	3.62
725	0.334	0.808	1.59	3.85
750	0.353	0.823	1.68	3.92
775	0.370	0.837	1.76	3.99
800	0.380	0.860	1.81	4.10

6.2.4. Particle diameter = 0.88 cm.

Mean Temperature of Peat Particle T_m (°C)	Specific Reaction rate, q (mg/cm^2.s)	Theoritical Diffusional Rate, q_{diff} (mg/cm^2.s)	Overall Rate co-efficient, k_{ov} (mg/cm^2.s. atm)	Diffusional rate co-efficient, k_{diff} (mg/cm^2.s. atm)
475	0.097	0.523	0.46	2.49
525	0.162	0.568	0.77	2.71
575	0.198	0.593	0.94	2.82
625	0.236	0.615	1.12	2.93
675	0.273	0.626	1.30	2.98
725	0.308	0.665	1.47	3.17
750	0.323	0.677	1.54	3.22
775	0.337	0.688	1.61	3.28
800	0.350	0.718	1.67	3.42

TABLE NO. 6.3: BURNING RATES AND RATE CO-EFFICIENTS OF PEAT PARTICLES AT 180 1/h AIR FLOW RATE

6.3.1 Particle diameter = 0.42 cm.

Mean Temperature of Peat Particle T_m (°C)	Specific Reaction rate, q (mg/cm^2.s)	Theoritical Diffusional Rate, q_{diff} (mg/cm^2.s)	Overall Rate co-efficient, k_{ov} (mg/cm^2.s.atm)	Diffusional rate co-efficient, k_{diff} (mg/cm^2.s.atm)
475	0.225	1.022	1.07	4.87
525	0.305	1.070	1.45	5.10
575	0.380	1.120	1.81	5.33
625	0.430	1.164	2.05	5.54
675	0.459	1.210	2.19	5.76
725	0.475	1.254	2.26	5.97
750	0.480	1.280	2.29	6.10
775	0.481	1.300	2.29	6.19
800	0.483	1.320	2.30	6.29

6.3.2. Particle diameter = 0.54 cm.

Mean Temperature of Peat Particle T_m (°C)	Specific Reaction rate, q (mg/cm^2.s)	Theoritical Diffusional Rate, q_{diff} (mg/cm^2.s)	Overall Rate co-efficient, k_{ov} (mg/cm^2.s.atm)	Diffusional rate co-efficient, k_{diff} (mg/cm^2.s.atm)
475	0.205	0.851	0.98	4.05
525	0.280	0.888	1.33	4.23
575	0.355	0.924	1.69	4.40
625	0.410	0.964	1.95	4.59
675	0.445	1.000	2.12	4.76
725	0.413	1.040	2.21	4.95
750	0.470	1.063	2.24	5.06
775	0.471	1.080	2.24	5.14
800	0.473	1.100	2.25	5.24

6.3.3 Particle diameter = 0.68 cm.

Mean Temperature of Peat Particle T_m (°C)	Specific Reaction rate, q (mg/cm^2.s)	Theoritical Diffusional Rate, q_{diff} (mg/cm^2.s)	Overall Rate co-efficient, k_{ov} (mg/cm^2.s. atm)	Diffusional rate co-efficient, k_{diff} (mg/cm^2.s. atm)
475	0.132	0.732	0.63	3.49
525	0.200	0.767	0.95	3.65
575	0.270	0.800	1.29	3.81
625	0.335	0.831	1.60	3.96
675	0.395	0.870	1.88	4.14
725	0.439	0.962	2.09	4.30
750	0.451	0.924	2.15	4.40
775	0.460	0.940	2.19	4.48
800	0.462	0.960	2.20	4.57

6.3.4 Particle diameter = 0.88 cm.

Mean Temperature of Peat Particle T_m (°C)	Specific Reaction rate, q (mg/cm^2.s)	Theoritical Diffusional Rate, q_{diff} (mg/cm^2.s)	Overall Rate co-efficient, k_{ov} (mg/cm^2.s. atm)	Diffusional rate co-efficient, k_{diff} (mg/cm^2.s. atm)
475	0.082	0.621	0.39	2.96
525	0.120	0.632	0.57	3.01
575	0.150	0.682	0.71	3.25
625	0.178	0.711	0.85	3.39
675	0.205	0.740	0.98	3.52
725	0.233	0.770	1.11	3.67
750	0.246	0.783	1.17	3.73
775	0.255	0.794	1.21	3.79
800	0.261	0.808	1.24	3.85

TABLE NO. 6.4: BURNING RATES AND RATE CO-EFFICIENTS OF PEAT PARTICLES AGAINST TIME FOR 60 l/h AIR FLOWRATE.

6.4.1 Particle diameter = 0.54 cm.

Mean Temperature of Peat Particle, T_m (°C)	Residence Time in reactor, t (s)	Specific Reaction rate, q (mg/cm^2.s)	Diffusional rate co-efficient, k_{diff} (mg/cm^2.s.atm)	Ash Film diffusion constant, K_{ash} (mg/cm^2.s.atm.)
713	30	0.67	3.46	-
713	40	0.50	3.46	9.09
716	60	0.44	3.48	5.95
717	80	0.40	3.49	4.61
716	100	0.33	3.49	3.02
717	120	0.32	3.49	2.87
717	140	0.28	3.49	2.25
717	150	0.26	3.49	2.01

6.4.2 Particle diameter = 0.88 cm.

Mean Temperature of Peat Particle, T_m (°C)	Residence Time in reactor, t (s)	Specific Reaction rate, q (mg/cm^2.s)	Diffusional rate co-efficient, k_{diff} (mg/cm^2.s.atm)	Ash Film diffusion constant, K_{ash} (mg/cm^2.s.atm.)
710	30	0.39	2.13	-
710	45	0.31	2.12	7.41
715	60	0.28	2.14	4.66
715	75	0.22	2.14	2.41
709	90	0.17	2.12	1.44
715	120	0.16	2.14	1.28
709	180	0.14	2.12	1.05
720	540	0.13	2.15	0.93
720	590	0.12	2.15	0.82

TABLE NO. 6.5: SURFACE REACTION RATE CO-EFFICIENT OF PEAT PARTICLES AT 700°C FOR 60 l/h FLOWRATE OF DRY AIR.

Diameter, d (cm)	Experimental surface reaction rate co-efficient, k_s ($mg/cm^2.s.\ atm$)	Theoretical surface reaction rate co-efficient, $K_s=8710\ \exp(-35700/RT_s)$ (k_s) ($mg/cm^2.s.\ atm.$)
0.54	45.9	0.11
0.88	14.5	0.10

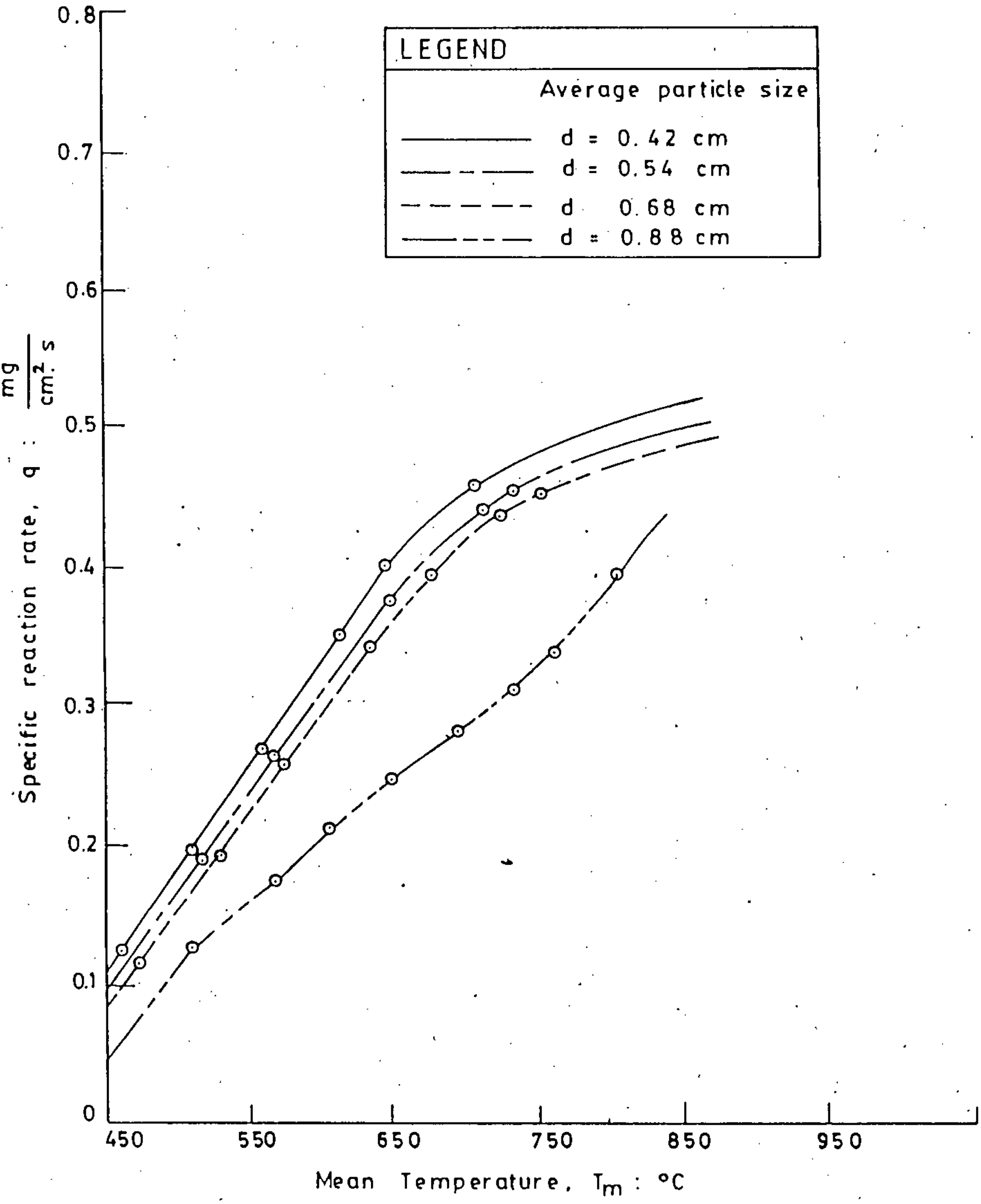


Fig. 6.1.1: Variation of Specific reaction rate with particle temperature at 60 l/h air flowrate.

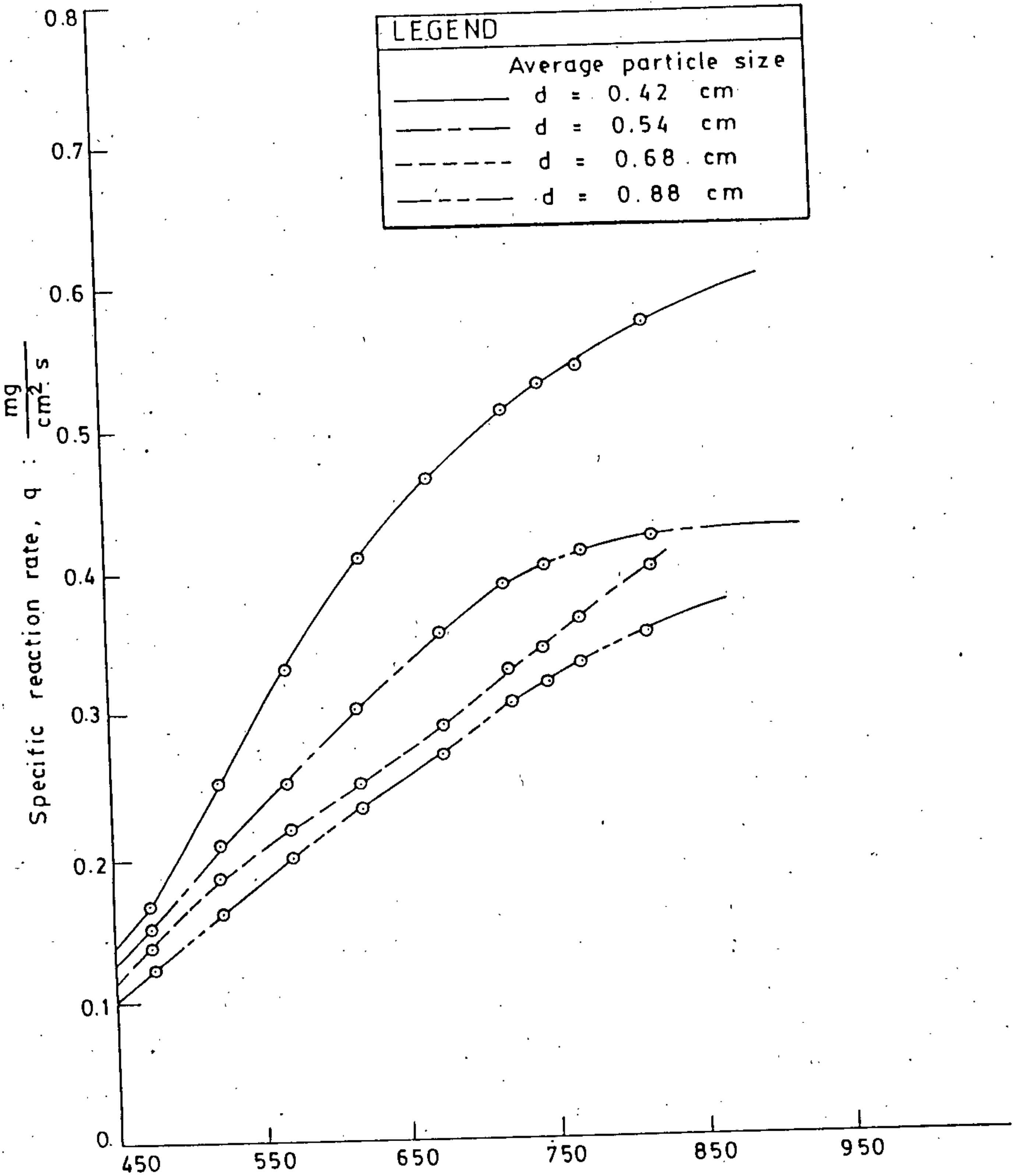


Fig. 6.1.2: Variation of Specific reaction rate with particle temperature at 125 l/h air flowrate.

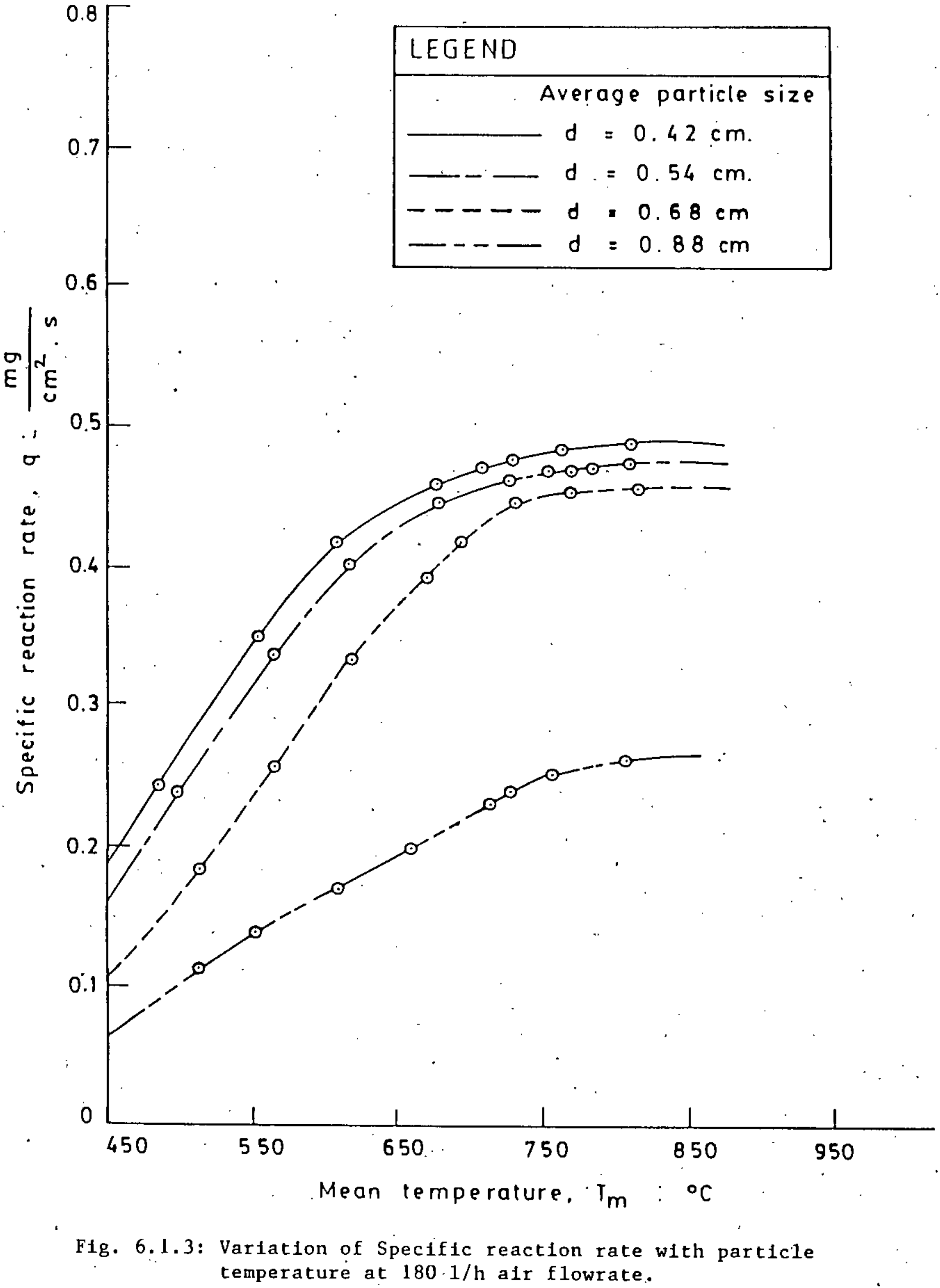


Fig. 6.1.3: Variation of Specific reaction rate with particle temperature at 180 l/h air flowrate.

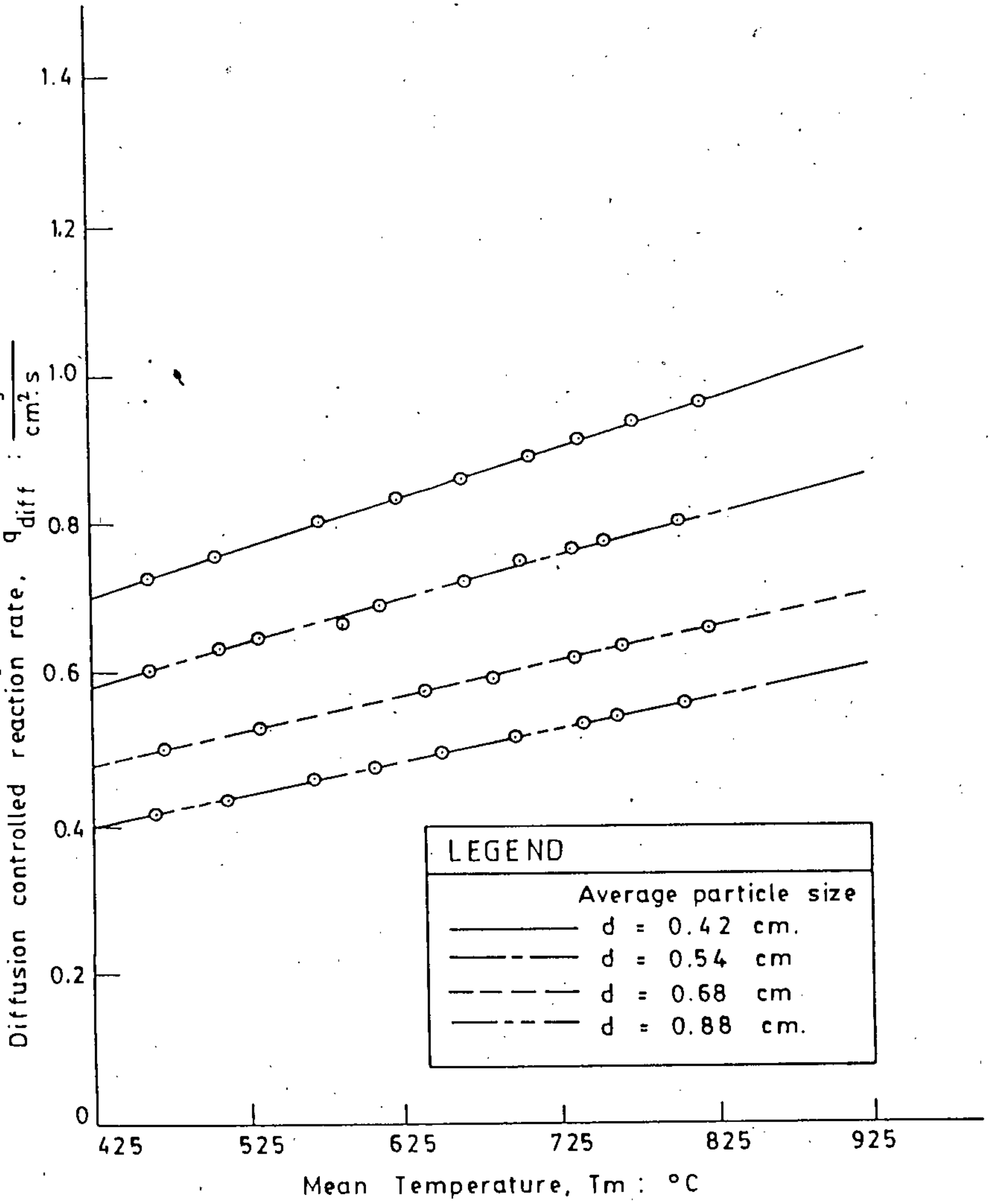


Fig. 6.2.1: Relation between diffusion controlled reaction rate and mean temperature of peat particle at 60 l/h of air flowrate.

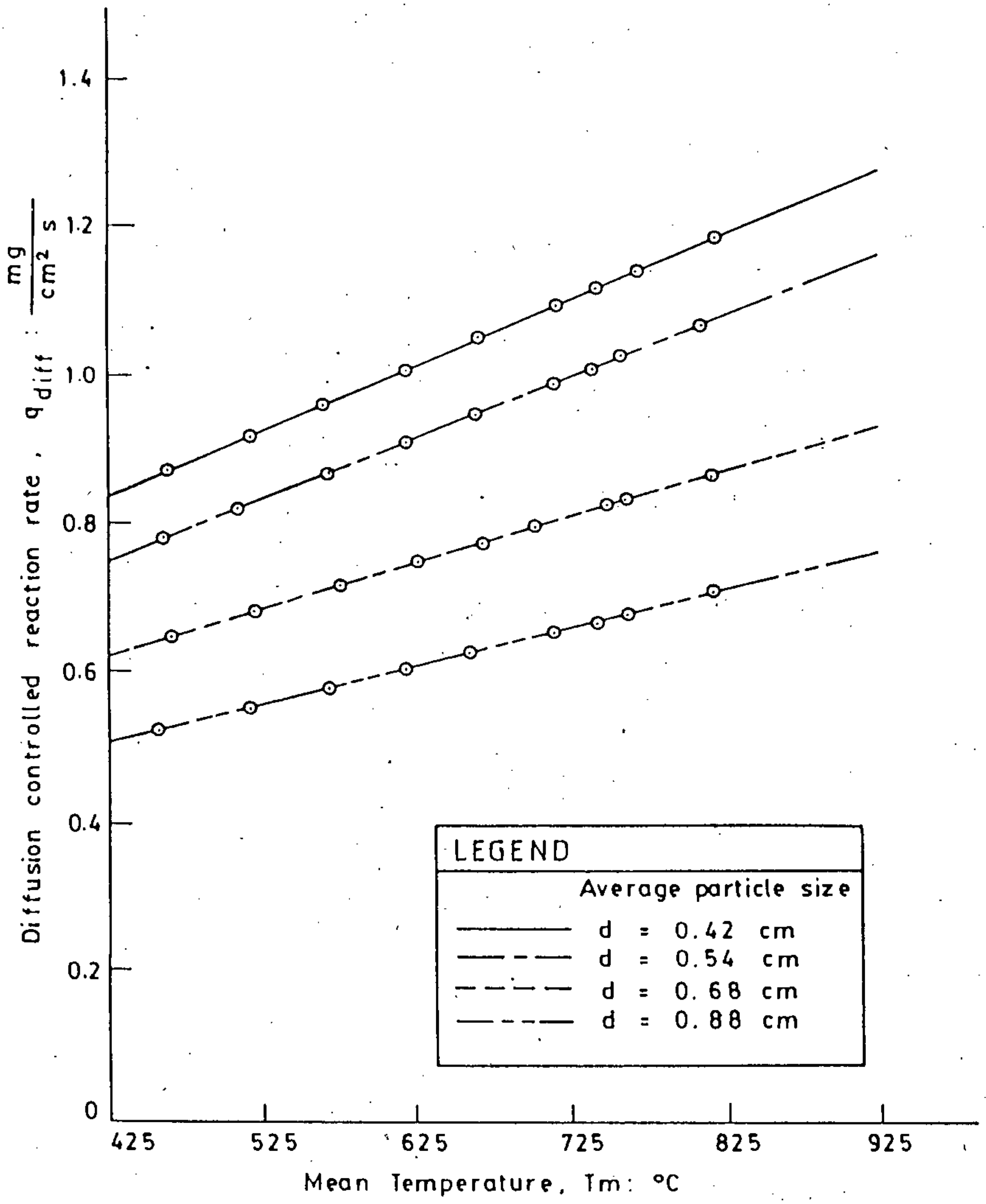


Fig. 6.2.2: Relation between diffusion controlled reaction rate and mean temperature of peat particle at 125 l/h of air flowrate.

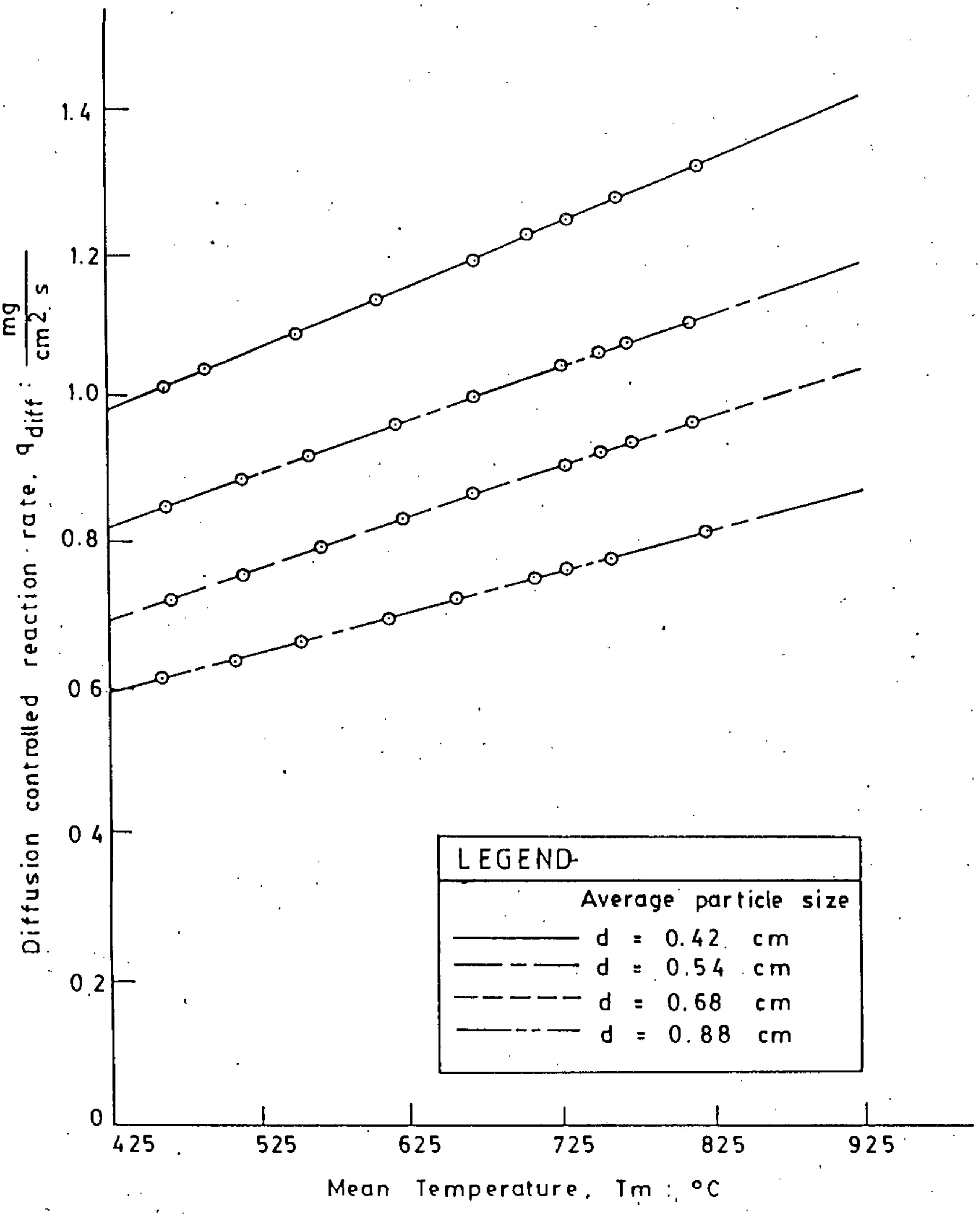


Fig. 6.2.3: Relation between diffusion Controlled reaction rate and mean temperature of peat particles at 180 l/h of air flowrate.

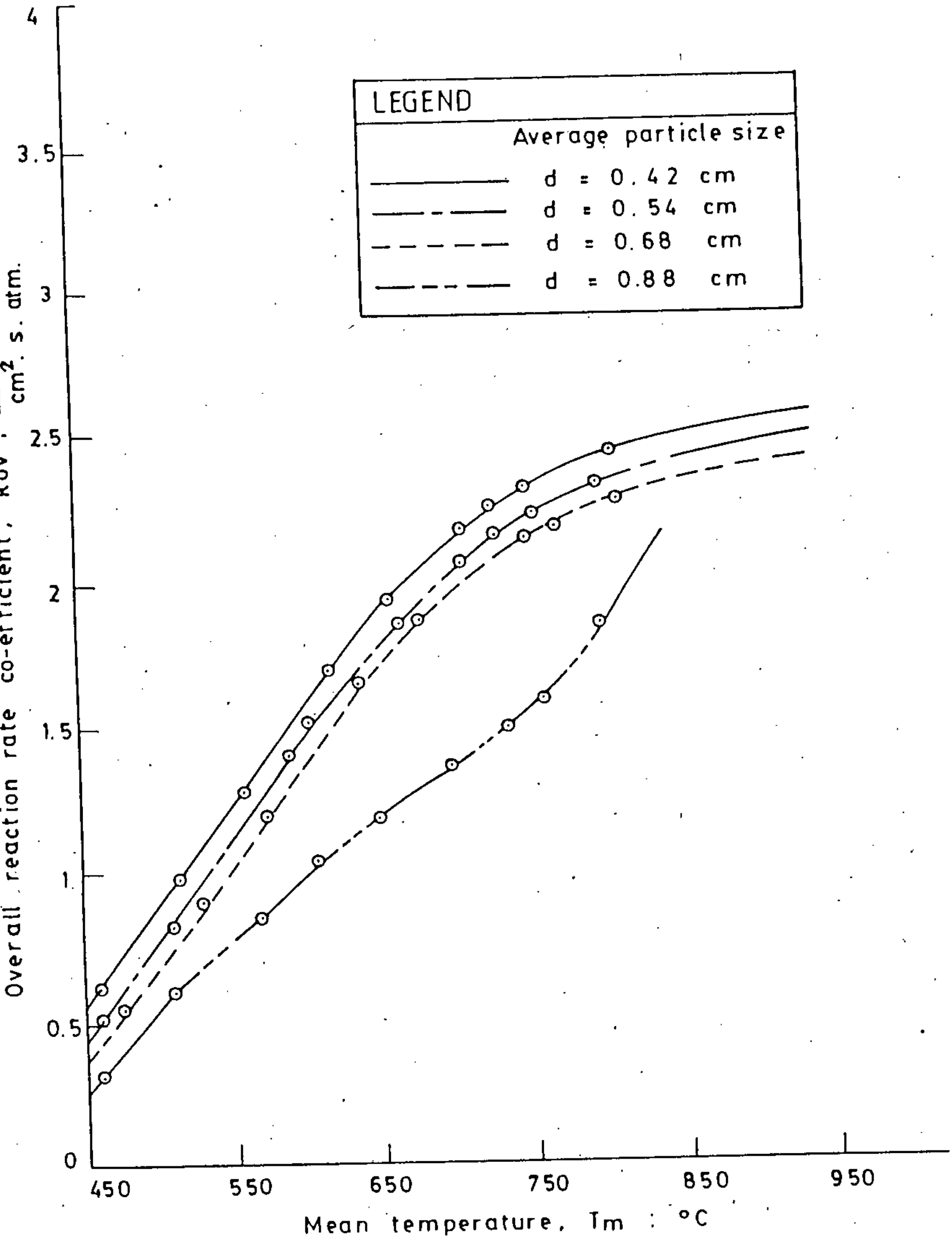


Fig. 6.3.1: Variation of Overall reaction rate co-efficient with temperature at 60 l/h of air flowrate.

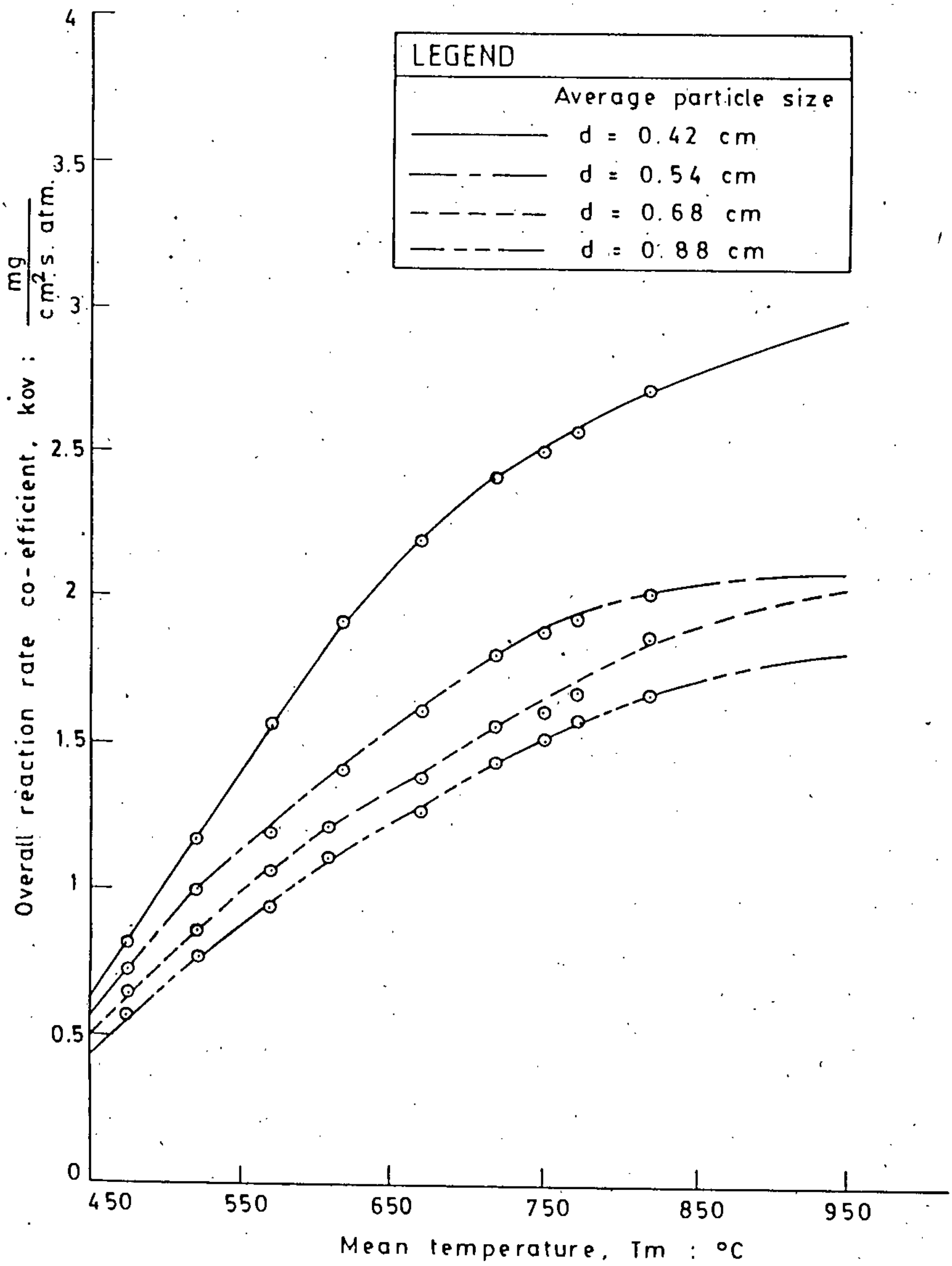


Fig. 6.3.2: Variation of Overall reaction rate co-efficient with temperature at 125 l/h of air flowrate

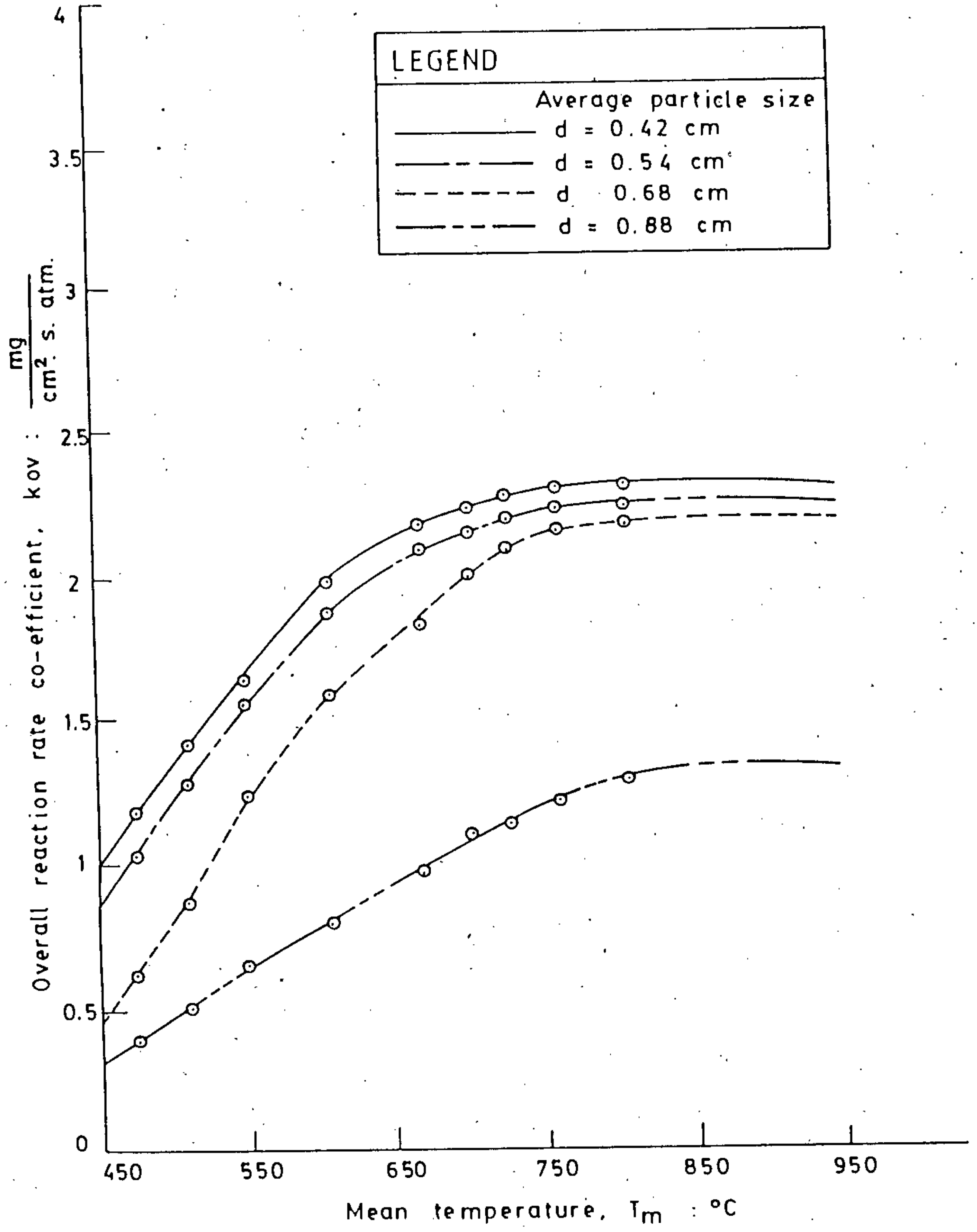


Fig. 6.3.3: Variation of Overall reaction rate Co-efficient with temperature at 180 l/h of air flowrate

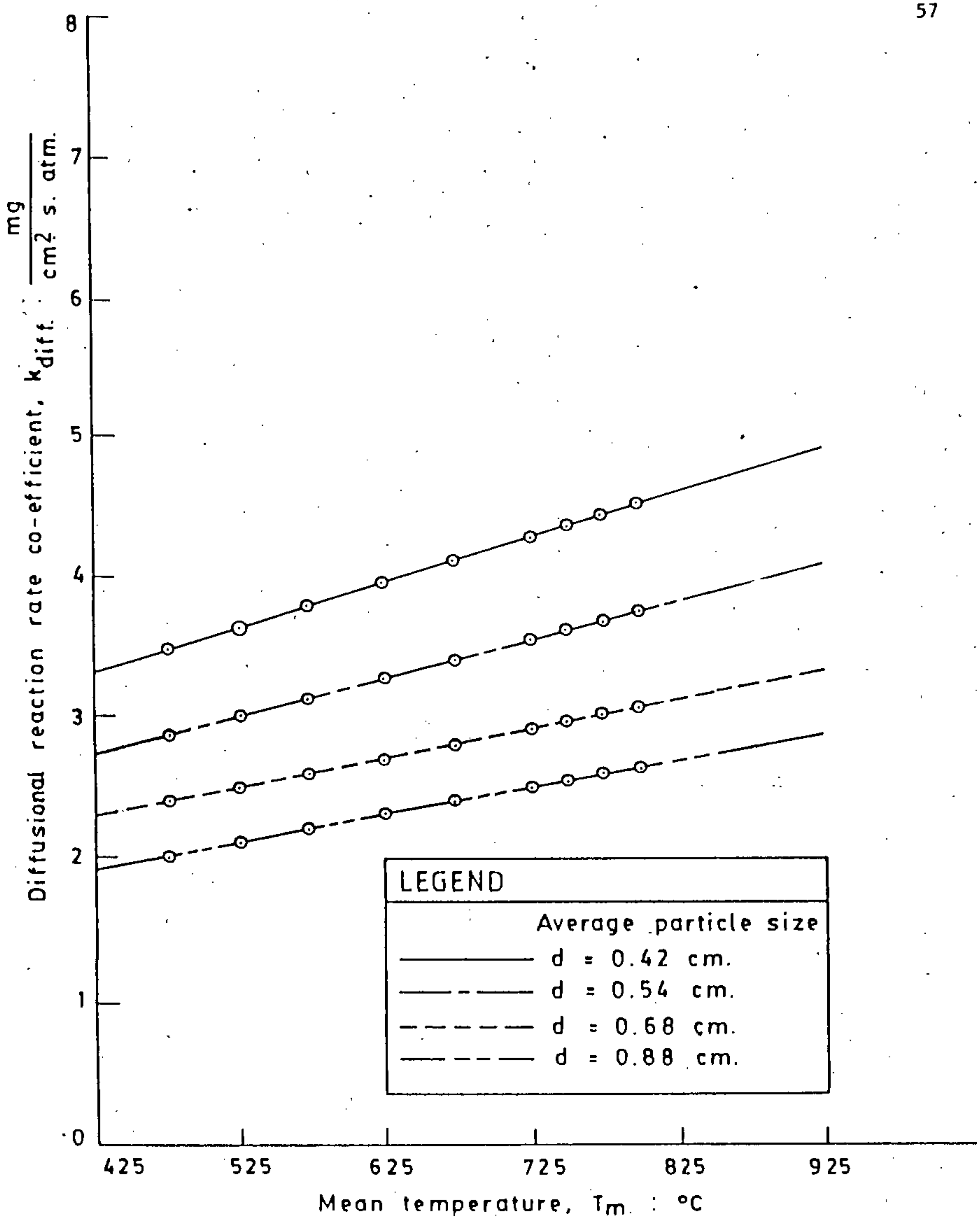


Fig. 6.4.1: Relation between diffusional reaction rate co-efficient and mean temperature of peat particle at 60 l/h of air flowrate.

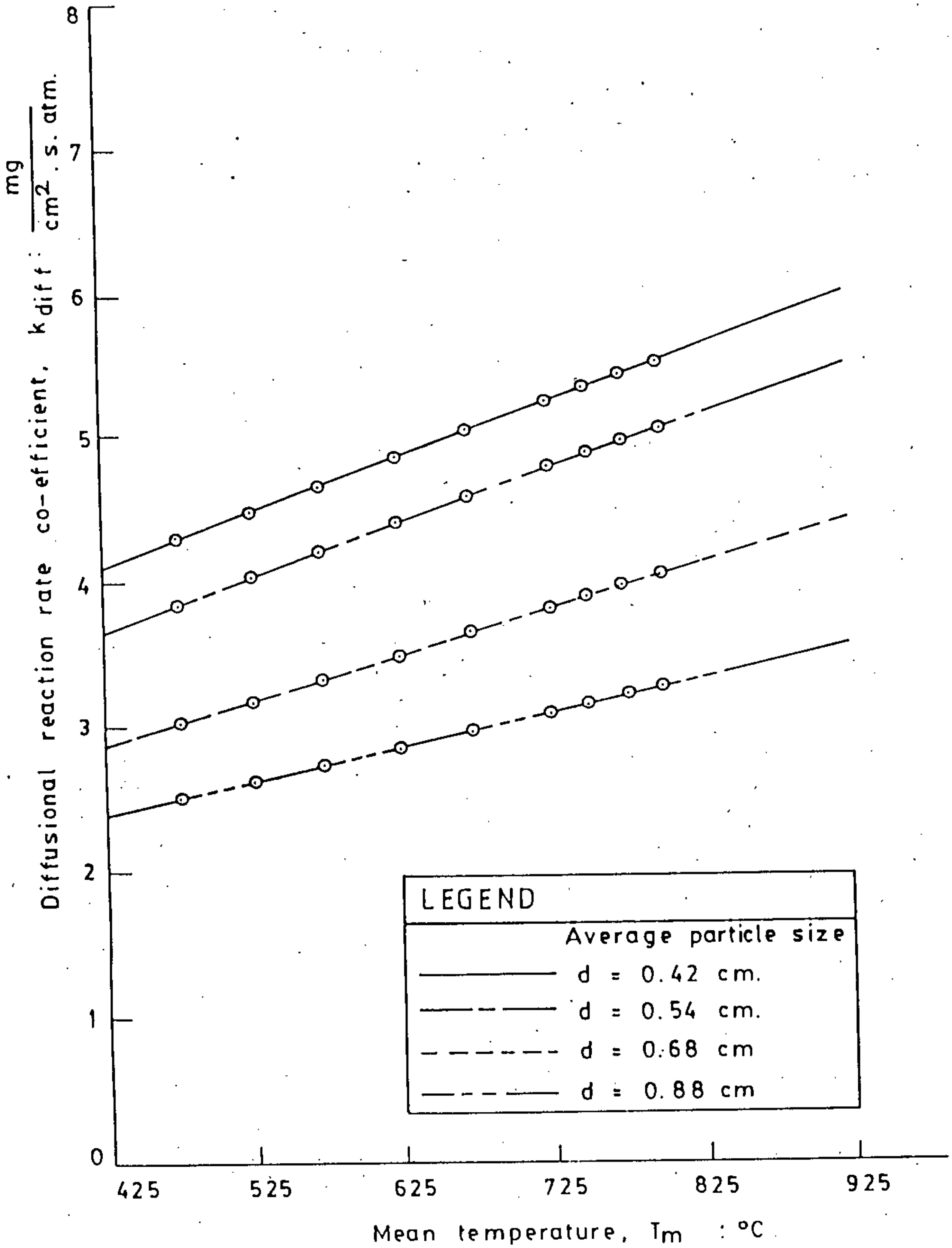


Fig. 6.4.2: Relation between diffusional reaction rate Co-efficient and mean temperature of peat particle at 125 l/h of air flowrate.

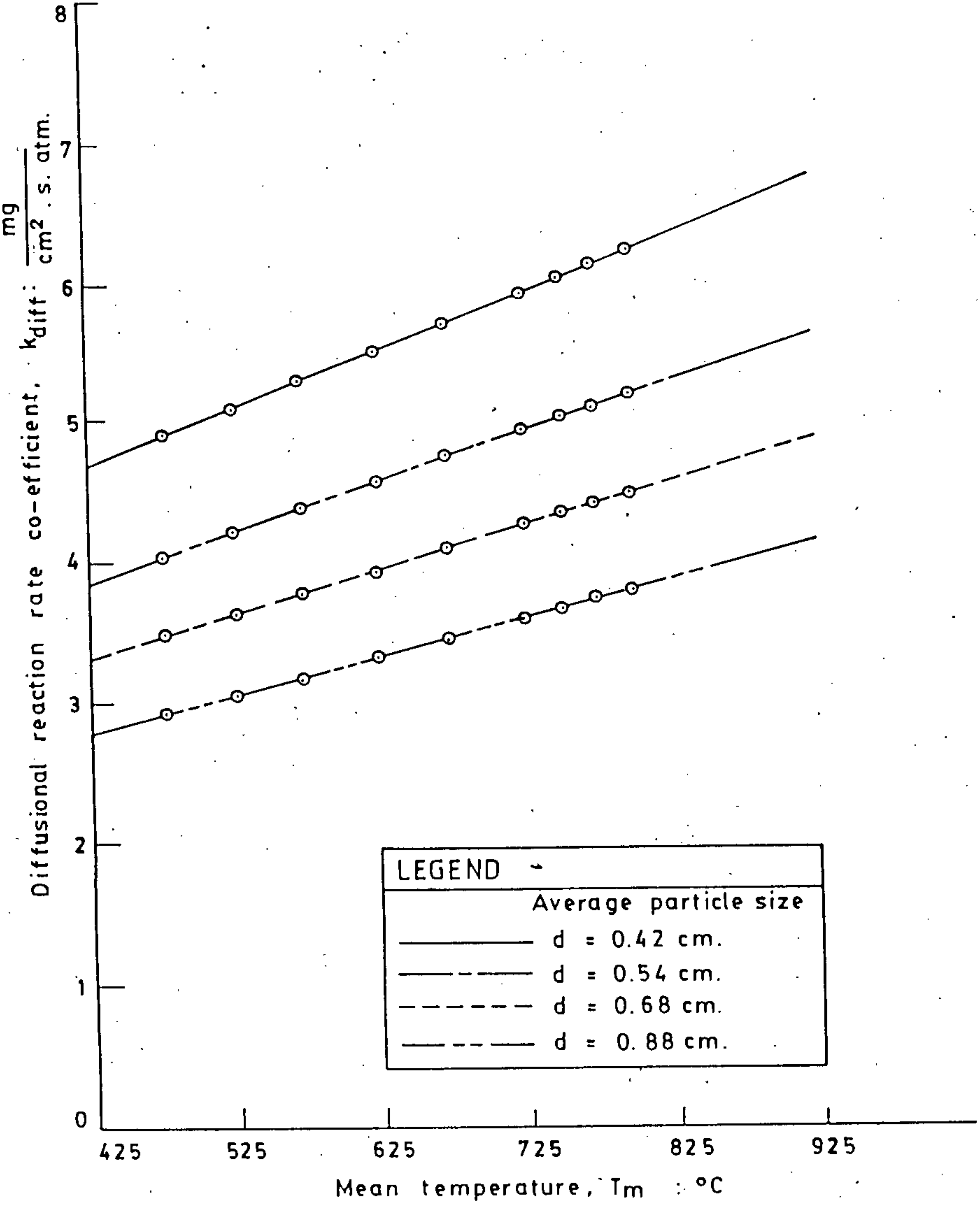


Fig. 6.4.3: Relation between diffusional reaction rate co-efficient and mean temperature of peat particle at 180 l/h of air flowrate.

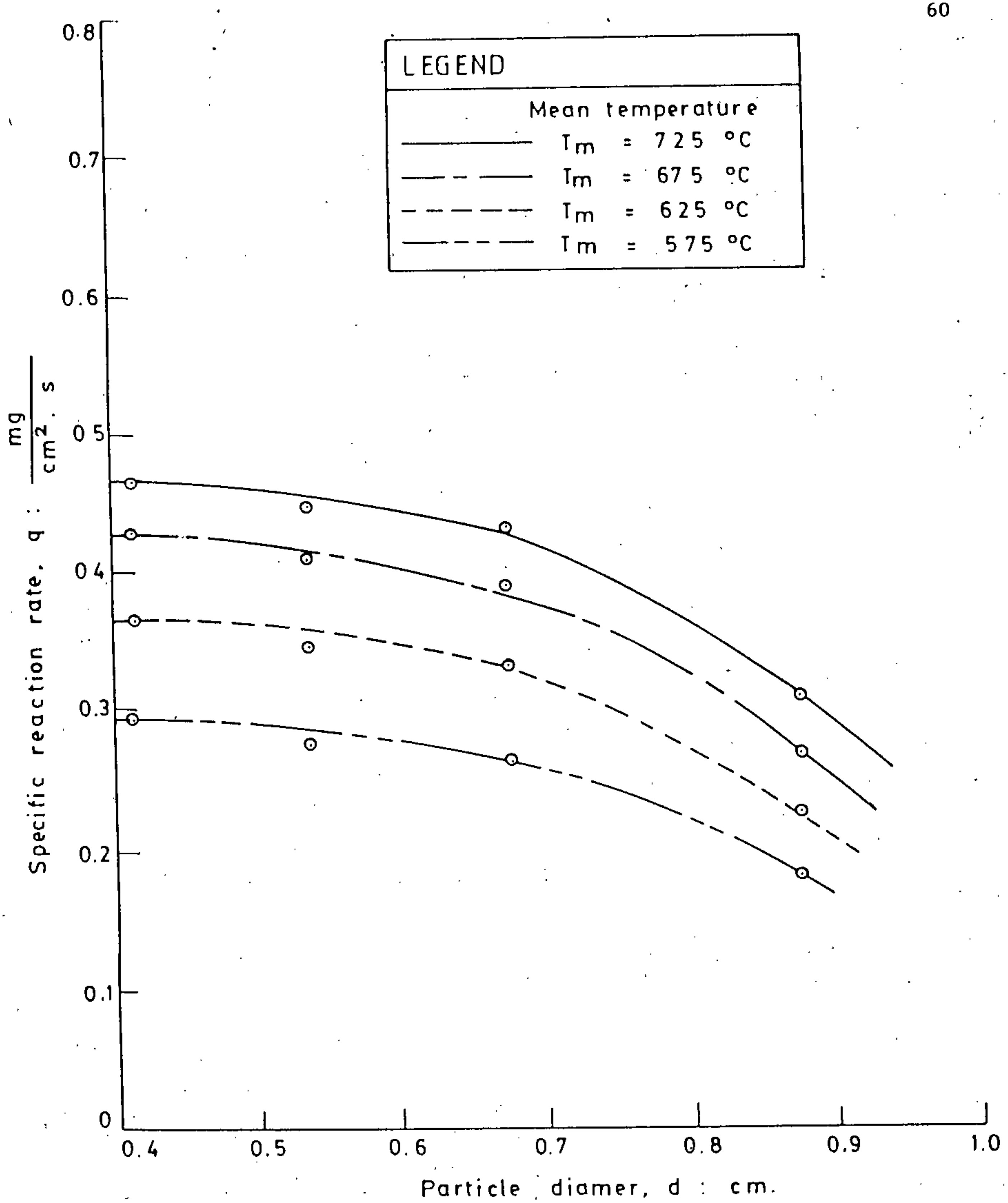


Fig. 6.5.1: Relation between Specific reaction rate and peat particle diameter at 60 l/h of air flowrate.

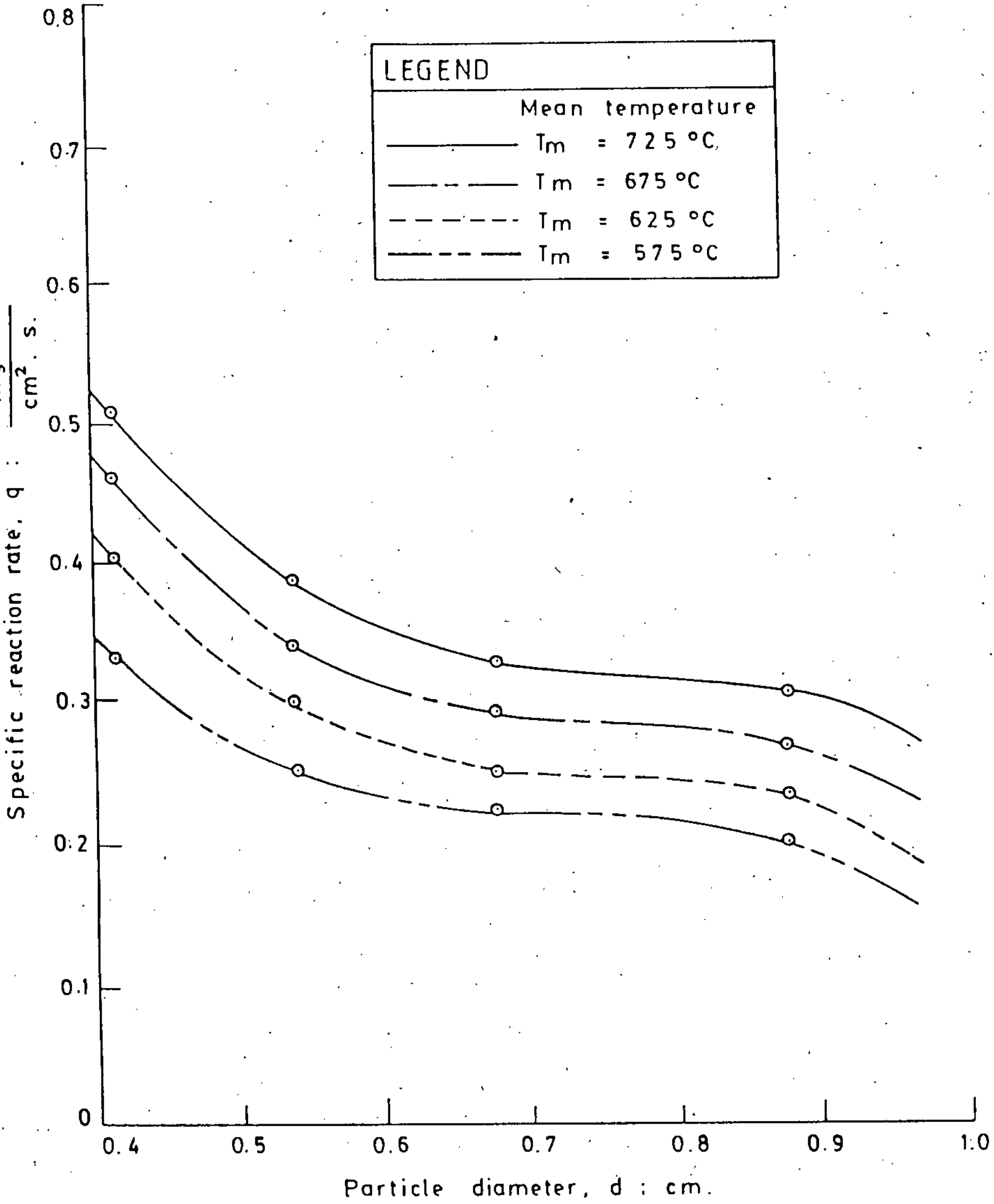


Fig. 6.5.2: Relation between Specific reaction rate and peat particle diameter at 125 l/h of air flowrate.

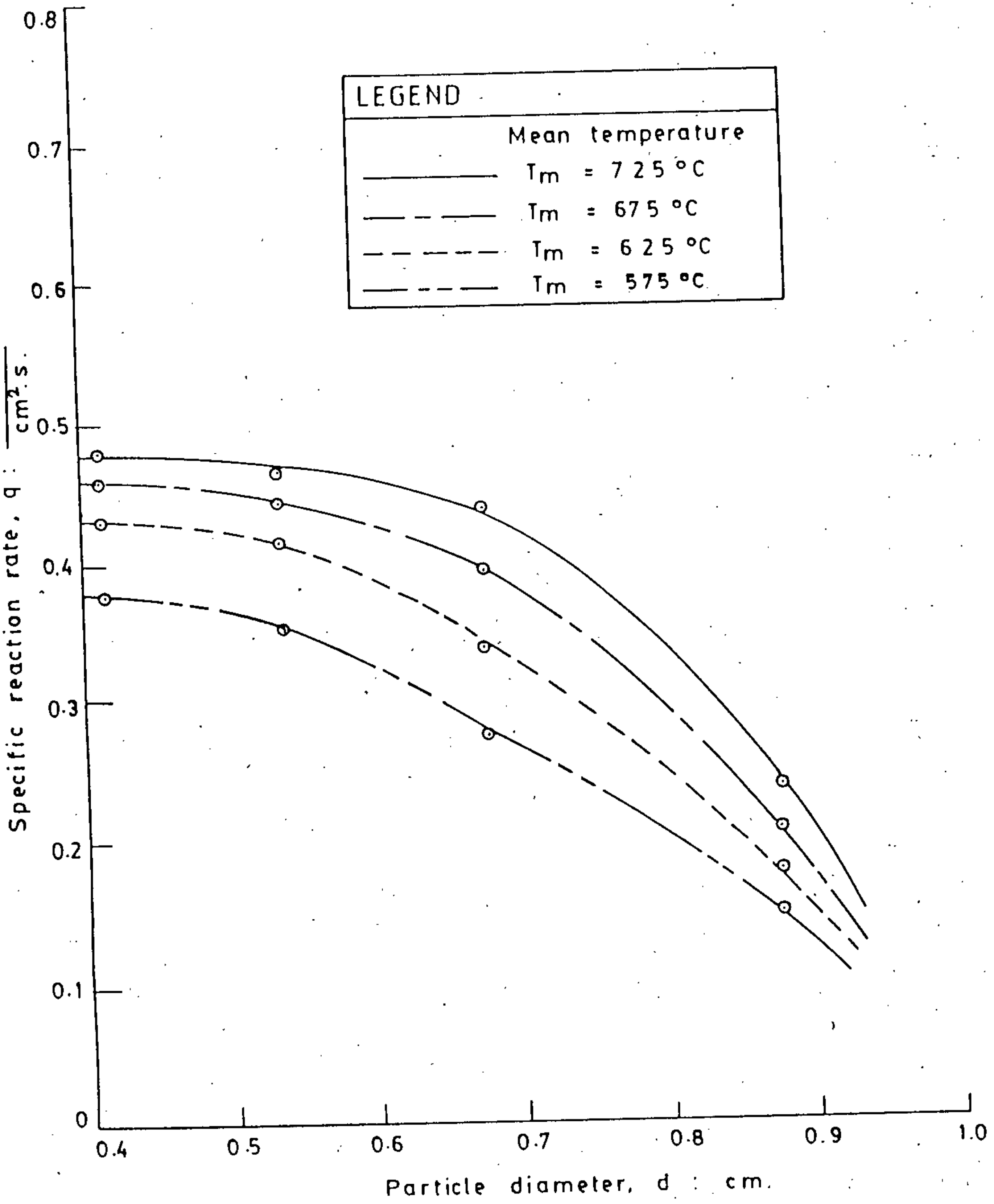


Fig. 6.5.3: Relation between Specific reaction rate and peat particle diameter at 180 l/h of air flowrate.

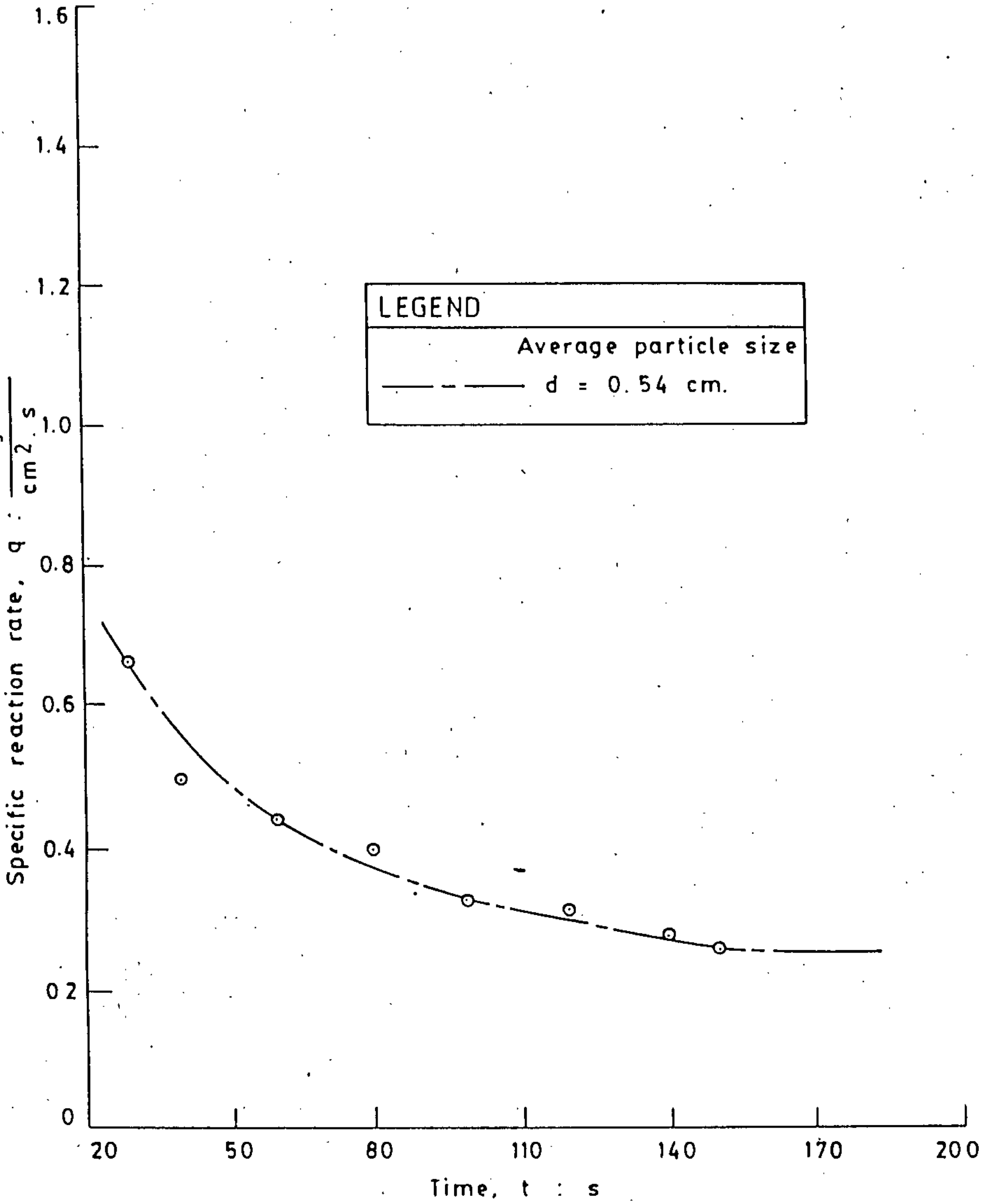


Fig. 6.6.1: Relation between Specific reaction rate and residence time of peat particle at Reactor temperature 700°C & 60 l/h of air flowrate.

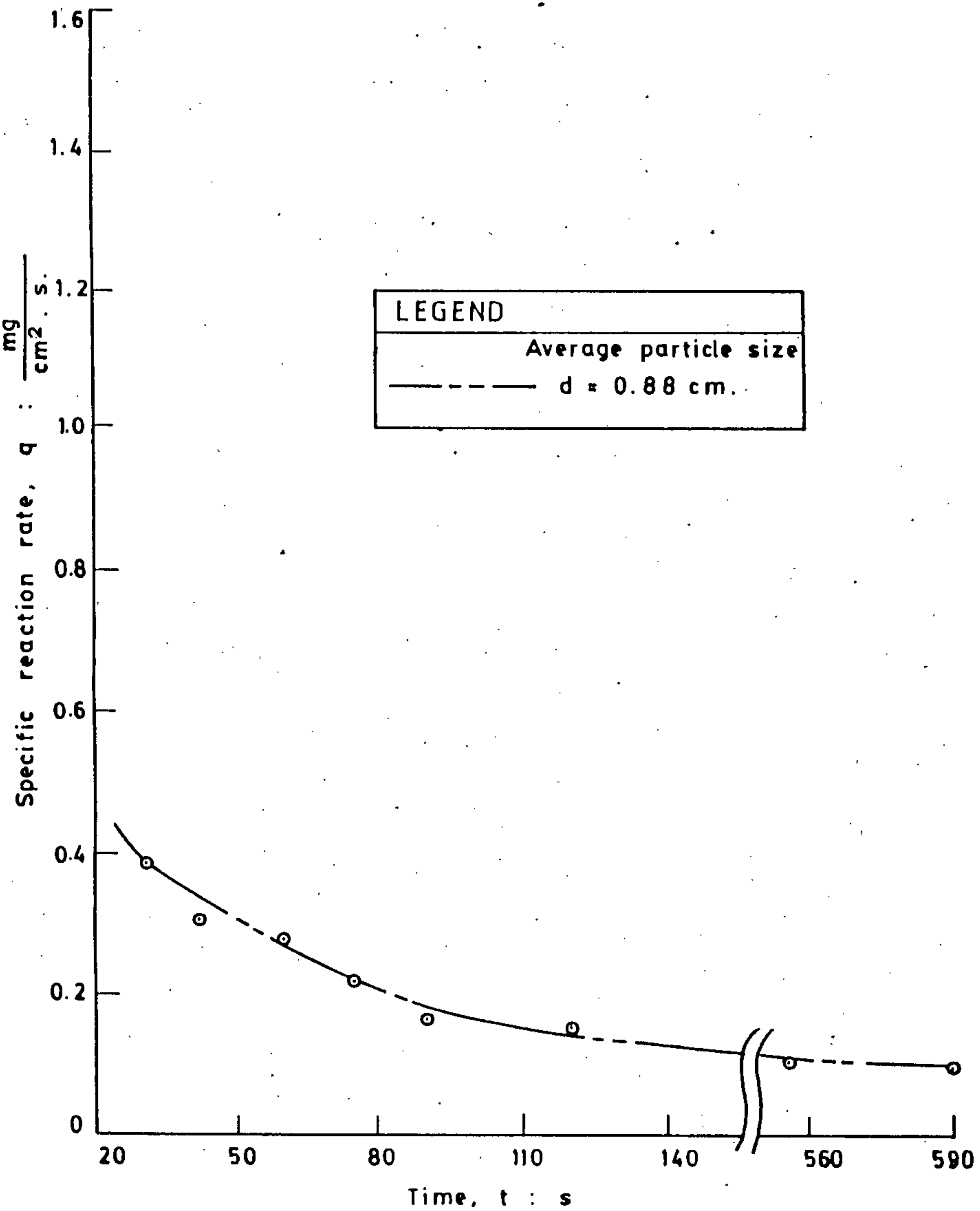


Fig. 6.6.2: Relation between Specific reaction rate and residence time of peat particle at Reactor temperature 700°C & 60 l/h of air flowrate.

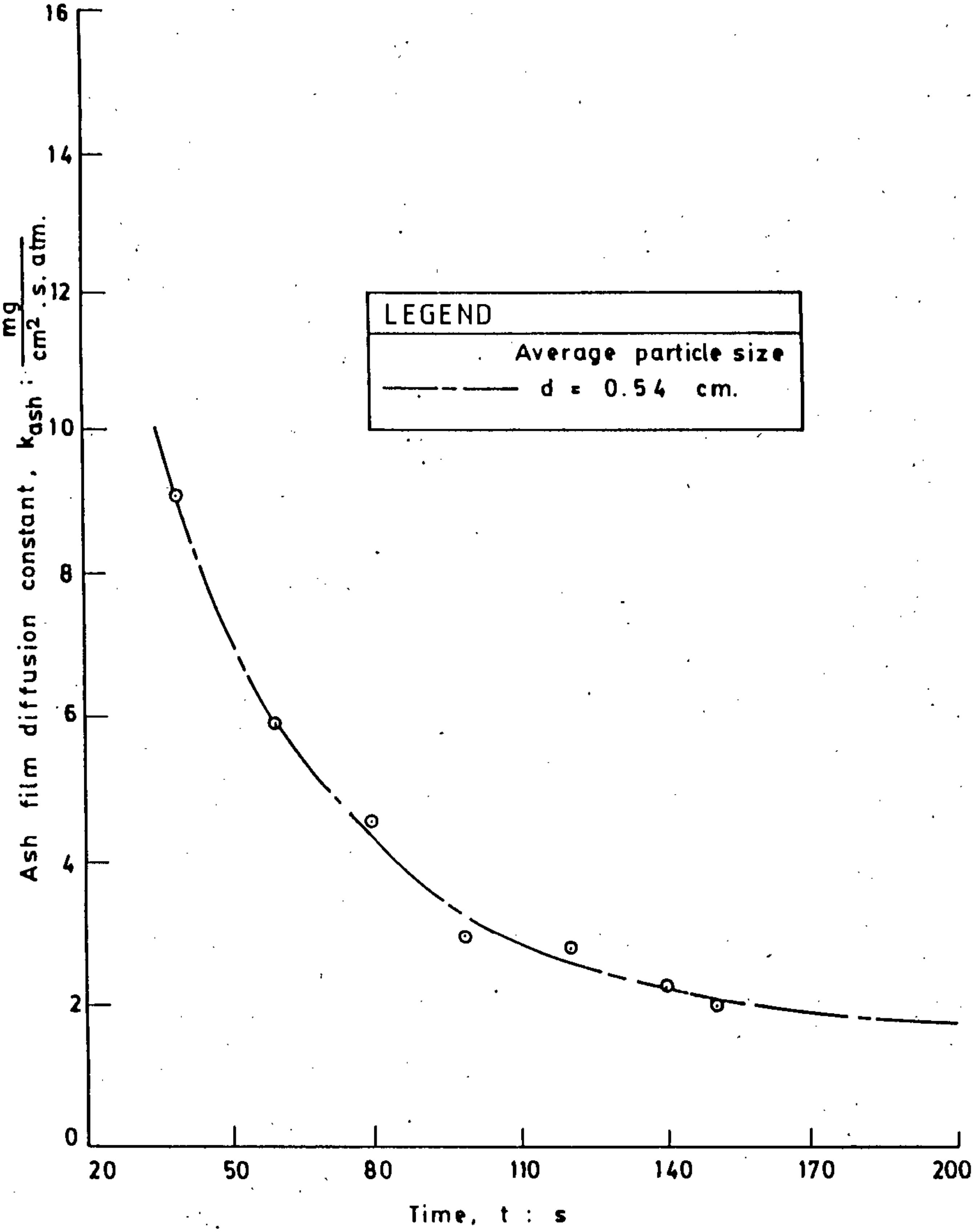


Fig. 6.7.1: Relation between Ash film diffusion constant and residence time of peat particles at Reactor temperature 700°C & 60.1/h of air flowrate.

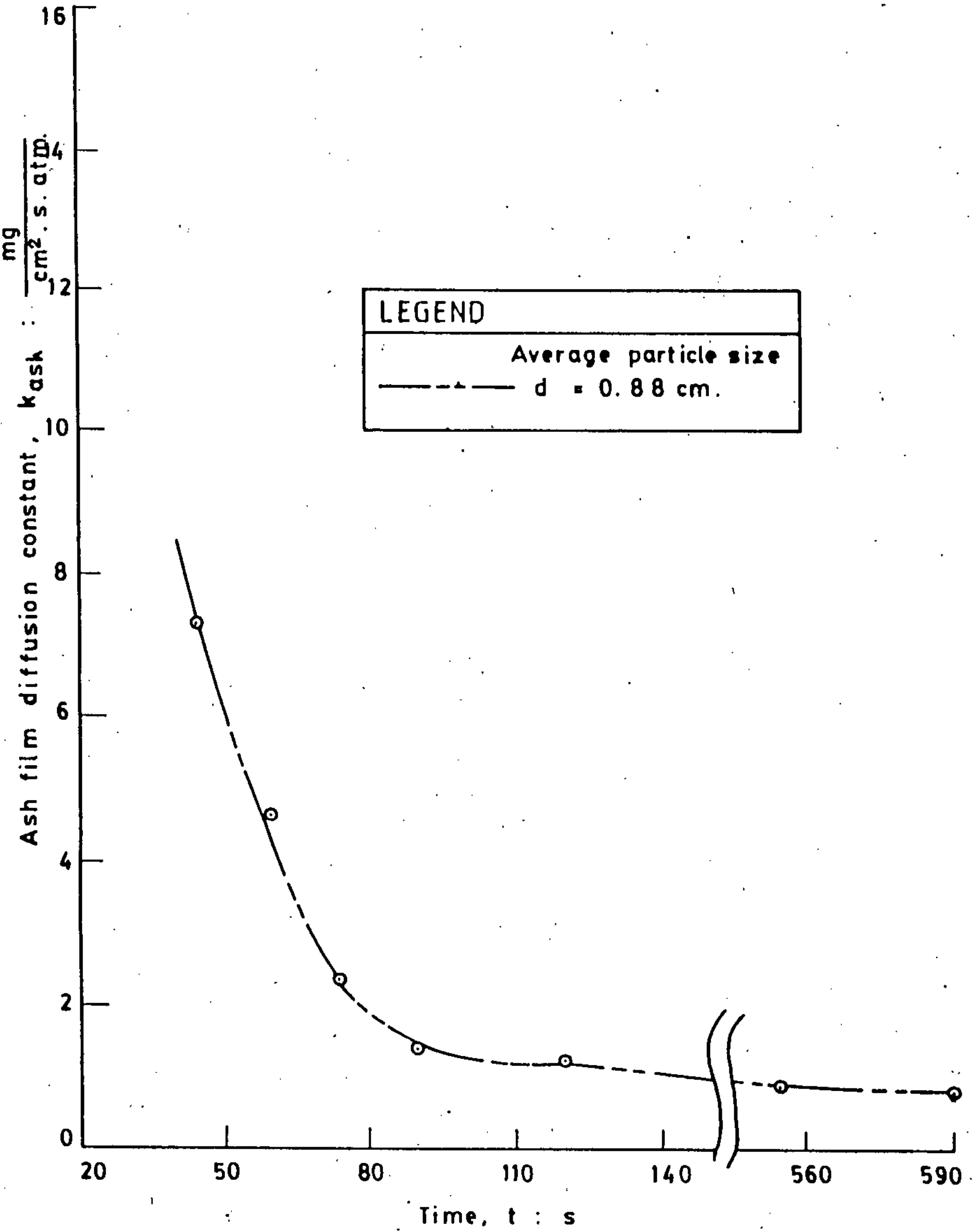


Fig. 6.7.2: Relation between Ash film diffusion constant and residence time of peat particle at Reactor temperature 700°C & 60 l/h of air flowrate.

CHAPTER-7 DISCUSSION

7. DISCUSSION

7.0 Introduction

During the research work a number of experiments was conducted to study the burning phenomena of devolatilised peat particles. The variables studied were: reaction temperature (400°C - 800°C), air flowrate (60 l/h - 180 l/h), particle diameter (d = 0.42 cm - 0.88 cm) and time. The detail description of the experimental set-up and method of experiments have been described in chapters 4 and 5 respectively. The bone - dried peat spheres were devolatilised at 900°C in a slow stream of nitrogen before each experiment and the devolatilised peat particles were used for experiments. Experimental specific reaction rates calculated by equation 6.1.1, are presented in Tables 5.1 - 5.3. Computer programme for data processing is given in Appendix-F. Tables 6.1.1 to 6.3.3 and Figures 6.1.1 to 6.5.3 present the data calculated on the basis of the computer output of specific reaction rate.

7.1 Accuracy of measurements

The main difficulties encountered during experimentation was quick opening and closing of svalves to avoid error in recording reaction time. Adequate precautions were taken to minimize errors in the experimental measurements nevertheless a considerable error was involved. About 20% of the observations for determining burning rates of peat were repeated, the results were found to be reproducible within ± 5 percent.

7.2 Effect of surface temperature, particle size and flowrate on burning of devolatilised Peat spheres:

Experimental specific reaction rate against temperature has been plotted in Figure 6.1.1 - 6.1.3. Comparing these figures it can be observed that 'q' is directly proportional to temperature and inversely proportional to diameter. The increase of reaction rate with increasing temperature were also observed by Golovina and Kaustovich (21). The inverse relationship of particle diameter and reaction rate was also observed by Farooq in a static system (9). The curves of theoretical bulk diffusional control rate q_{diff} vs. temperature at different flowrates are also presented in Figure 6.2.1 - 6.2.3. q_{diff} increases with the increase in temperatuere and decreases with increase in diameter. But the curves of specific reaction rate and bulk diffusional controlled rate are not of same nature.

The rate of burning of devolatilised peat spheres may be considered to be bulk diffusional controlled if the values of experimental specific reaction rate are closer to bulk diffusional controlled reaction rate (19). It is shown in Tables 6.1 to 6.3 that at all the flowrates and diameters, q and q_{diff} differ substantially from each other. Considering the above mentioned points, it may be assumed that the burning rate of devolatilised peat particles is not bulk diffusion controlled.

On the otherhand in kinetically controlled regime, the burning rate depends upon the temperature exponentially and is independent of particle size and flow rate (19). Experimental specific reaction rate q increases with temperature but exponential increase of q with temperature has not been observed. Besides this, q depends on flowrate at all particle size. So the combustion of devolatilized peat does not seems to be kinetically controlled.

Considering all the points discussed above, it may be concluded that the combustion reaction of devolatilised peat is dependent on the mass transfer to the reacting surface and the kinetics of the reaction. But at the last stage of burning, K_{ash} reaches a limiting point (Table 6.4). The effect of flowrate of air on the reaction rate has been found to be dependent on the level of temperature and particle size. For the smallest particle (0.42 cm), the reaction rate increase with temperature upto 650°C for all flow rates. For large particles, the effect of air flow rates on the rate of reaction do not follow any pattern. Similar phenomena were observed by Tu, Davis and Hottel (22) in the case of coal-char burning when both the diffusion and reaction kinetics affect the combustion process. K_{ash} and K_s are also inversely proportional to diameter (Table 6.4, 6.5). Therefore K_{ov} decreases with the increase in particle size. These facts may be the reason for the observed inverse relationship of specific reaction rate q with diameter.

7.3 Effect of residence time (Reaction time) on burning of devolatalized peat spheres.

Peat spheres of different sizes were burned completely at 700°C and 60 l/h air flowrate. Specific reaction rate calculated for different sizes of peat have been plotted against time in Figures 6.6.1 and 6.6.2. K_{ash} against time also plotted in Figures 6.7.1 and 6.7.2.

The overall reaction rate co-efficient depends on the following factors (i) the bulk diffusion co-efficient (K_{diff}), (ii) the surface reaction rate constant (K_s) and (iii) the ash film diffusion constant (K_{ash}). However, at constant temperature K_{diff} and K_s will be approximately constant and the reaction rate will depend mainly on K_{ash}. It can be seen from Figures 6.6.1 and 6.6.2 that the reaction rate decreases with time, initially rapidly and then slowly. This can be explained by the fact that the ash film resistance grows rapidly at first

& then slows down with time. The reaction rate thus slows down with time & fall exponentially. If the Figures are compared for the two particle diameters it can be seen that the reaction rate falls more rapidly for the smaller particle. This may be related to the lower bulk diffusion resistance for the small particle.

7.4 Surface reaction rate co-efficient:

A number of investigators (21-25) conducted a number of experiments to develop an expression for surface reaction rate co-efficient for coal-char burning. All the results of their experiments can be represented approximately by:

$$K_s = 8710 \exp(-35700/R'T_s) \quad .. \qquad (7.4.1)$$

Table 6.5 shows experimental and theoretical K_s value using the Equation 7.4.1 at 700°C. It is noticed that experimental values differ significantly from the theoretical values. Therefore the expressions of K_s for coal-char burning is not applicable for devolatilized peat particle, burning.

Literature review on K_s reveals that K_s depends on many factors such as carbon type, temperature, oxygen partial pressure, particle size, surface roughness, porosity of carbon etc. However, results of investigators in this area show that K_s depends largely on carbon type and surface temperature only. As devolatilized peat particles and coal-char are structurally different, it can be assumed that theoretical K_s values calculated using Equation 7.4.1 may not be equal to experimental values. Experimental results of the present study have proved this assumption.

CHAPTER-8 CONCLUSIONS & SUGGESTION

8. CONCLUSIONS & SUGGESTIONS

8.1 Conclusions:

The present investigation on burning characteristics of devolatilised suspended peat particles in flow of dry air has led to the following conclusions:

i) Specific reaction rate of peat particles is a function of temperature, particle size and air flowrate.

ii) It has been observed that combustion reaction is not diffusion controlled under the investigated conditions.

iii) Specific reaction rate decreases exponentially with time. Initially it decreases rapidly & then slows down.

iv) Ash film diffusion constant (k_{ash}) is inversely proportional to diameter and decreases exponentially with time.

v) Peat ash retains its size and structuere after burning to completion in a stream of dry air. Since peat ash retains size and structure, diffusion of oxygen into the solid becomes an added resistance in the process of devolatilised peat particle burning.

vi) Expression of K_s for Coal-Char burning is not applicable for burning of devolatilised peat particle.

8.2 Suggestions for future work

The following suggestions can be put forward for consideration by future investigators:

i) In the present investigation, burning rate study in a stream of dry air was carried out using peat samples from one location (Kola-Mouza) only. Similar study may be carried out using peat samples from other locations in Bangladesh.

ii) Expressions for K_s, K_{ash}, K_{ov}, q and q_{diff} may be developed by carrying out further experiments at different temperature, flow rate and particle sizes.

iii) Study may be carried out with peat particles of other shapes & figures.

iv) Studies at higher temperature can be conducted by using high duty furnace.

CHAPTER-9 NOMENCLATURE

9. NOMENCLATURE

Symboles and abbreviations used in this thesis are defined in this chapter. The units have been mentioned whereever applicable.

Symbols	Significance	Unit
C_s	Specific heat of solid	Cal/gm. °K
d	Particle diameter	cm.
D	Diffusional co-efficient	cm^2/s.
f	Fuel, Oxygen ratio	(-)
f'	Multiplying factor for flow system	(-)
k_s'	Thermal conductivity of solid	Cal/cm. sec. °k
k_s	Surface reaction rate of co-efficient	mg/cm^2.sec. atm.
k_{ov}	Overall reaction rate co-efficient	mg/cm^2.sec.atm.
k_{diff}	Diffusional reaction rate co-efficient	mg/cm^2.sec.atm.
k_{ash}	Ash diffusion constant	mg/cm^2.sec.atm.
P	Partial pressure	atm.
P_g	Partial pressure of Oxygen at free stream	atm.
P_s	Partial pressure of Oxygen at surface	atm.
q	Specific reaction rate	mg/cm^2.s.
q_{diff}	Diffusional Reaction rate	mg/cm^2.s.
r	radial distance	cm.
R	Radius of particle	cm.
R'	Universal Gas constant	atm.cm^3/mol.°k.
S	Surface area of particle	cm^2.
t	Time	seconds
T_m	Mean temperature	°C
T_o	Initial Temperature	°k

T_α	Increased surface temperature	°k
T_p	Pyrolysis Temperature	°k
T_s	Surface Temperature	°k
W_o	Initial weight	mg
W_1	Final weight	mg
Y_o	Mass fraction of Oxygen	-
Y_{co_2}	Mass fraction of CO_2	-
Y_{co}	Mass fraction of CO	-
ρ_s	Density of solid	gm/cm^3
ρ_c	Density of carbon	gm/cm^3
α_s	Thermal diffusivity of solid	cm^2/sec.
ϕ	Mechanism factor	(-)
μ	Viscosity	m.s/kg.

Abbreviation	Significance
M.W.	Mega watt
N.T.P.	Normal temperature & pressure
l	Litre
h	Hour
mv	Milli volt
k.j.	Killo Joule
G.J.	Gega Joule
P.J.	Peta Joule

CHAPTER-10 REFERENCES

10. REFERENCES

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APPENDICES

APPENDIX-A

A. CALIBRATION OF THERMOCOUPLE

The Chromel-Alumel thermocouple used in the experiments was calibrated against a standard Pt-Pt-Rh thermocouple connected with a direct temperature indicator in a slow heating muffle furnace keeping the tips of the two thermocouples together and taking the readings from digital display meter and millivolt meter. The resulting calibration curve is shown in Figure A-1. It follows a straight line equation of the following form within the range mentioned:

$$T = \frac{mv-6}{0.035} \qquad 30 < mv < 47.5 \qquad \ldots \qquad (A.1)$$

where T = temperature in 0_C.

mv = millivolt meter reading, millivolts.

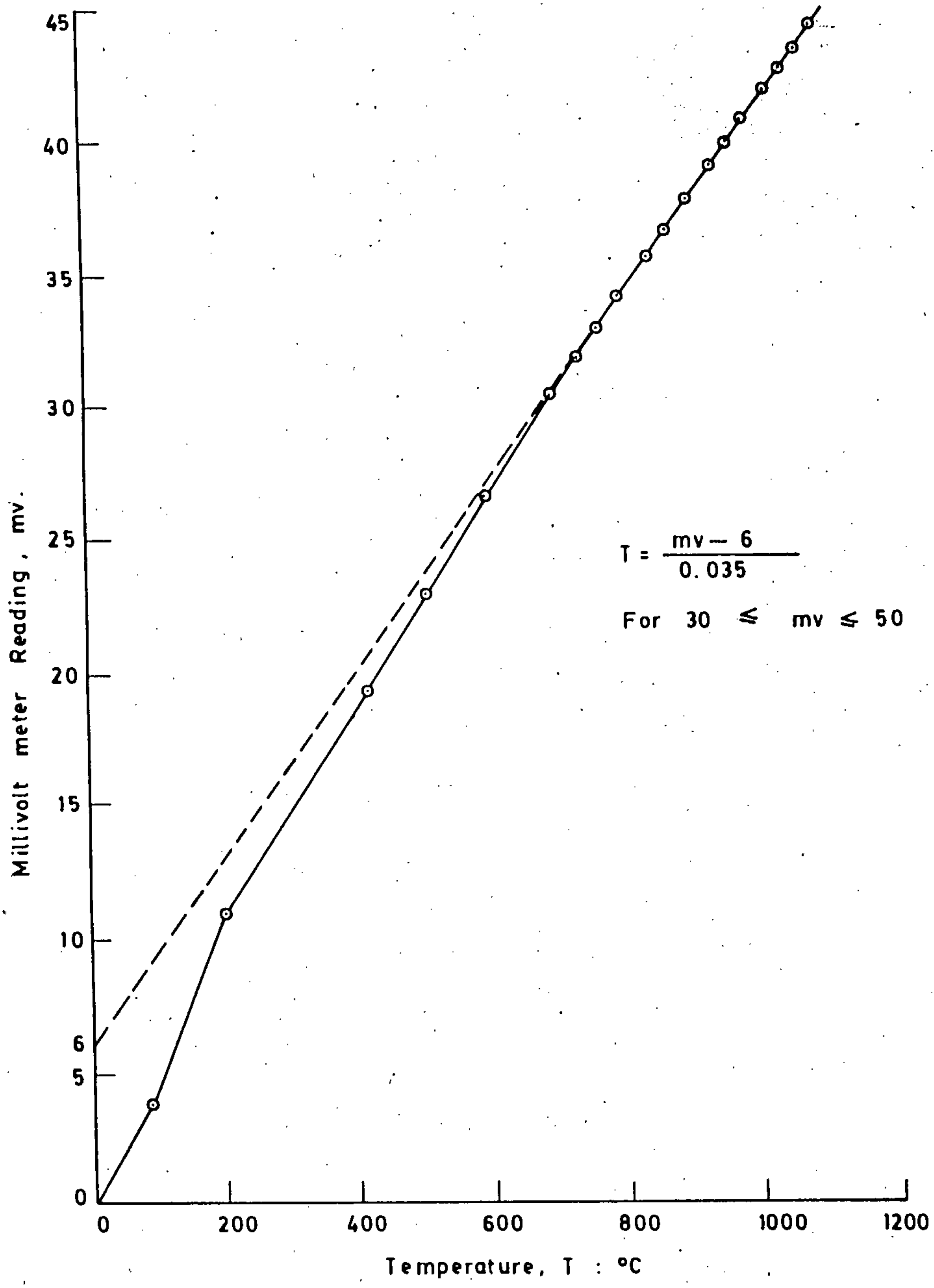

Fig. A-1: Calibration Curve for Thermocouple

APPENDIX-B

B. PROXIMATE ANALYSIS OF PEAT:

Proximate analysis of peat was done following the procedure mentioned in ASTM standard D271-68. Summary of the results are given in Table B-1.

B.1 Moisture content

	Weight in grams		
	Sample-1	Sample-2	Sample-3
Weight of empty crucible	16.60	13.00	10.60
Weight of Crucible + Sample	17.60	14.00	11.60
Weight of Crucible + Sample after drying in oven.	17.51	13.91	11.50
Moisture content	9%	9%	10%

Average moisture content : 9.33%

B.2 Ash content

	Weight in grams		
	Sample-1	Sample-2	Sample-3
Weight of empty crucible	16.60	13.00	10.60
Weight of crucible + Sample	17.60	14.00	11.60
Weight of crucible + Sample after drying in oven	17.51	13.91	11.50
Weight of crucible + Sample after heating in furnace	16.87	13.27	10.87
Ash content (Dry basis)	28.84%	28.77%	28.79%
Ash content (Wet basis)	26.91%	26.84%	26.86%

Average Ash Content (Dry basis) = 28.80%

Average Ash Content (Wet basis) = 26.87%

B.3 Volatile Matter:

Average moisture content of the sample : 9.33%

(detail in section B-1.).

	Weight in grams	
	Sample-1	Sample-2
Weight of empty crucible	16.61	16.61
Weight of Crucible + Sample	17.61	17.61
Weight of crucible + Sample + Lid	28.79	28.78
Weight of crucibles + Sample + Lid after heating in furnace.	28.27	28.26
Volatile matter (Dry basis)	45.81%	45.71%
Volatile Matter (Wet basis)	42.74%	42.65%.

Average volatile combustible matter (Dry basis) = 45.76%

Average volatile combustible matter (wet basis) = 42.70%

B.4 Fixed carbon content:

Percent fixed carbon (dry basis)

= 100 - ash percent - volatile matter percent

= 100 - 28.80% - 45.76%

= 25.44%

Percent fixed carbon (wet basis)

= 100 - ash percent - volatile matter percent - moisture percent.

= 100 - 26.87 - 42.70% - 9.33%

= 21.10%

B.5 Heating value of peat has been determined using Bomb Calorimeter

The value obtained is higher heating value. The value has been reported in TABLE B-1 after correcting for radiation loss.

TABLE B-1: PROXIMATE ANALYSIS OF PEAT

		Analysis of Present investigation	Analysis of Previous investigation(26)	
Moisture content	-	9.33%	-	
Ash (Dry basis)	-	28.80%	Ash (Dry basis)	26%
Ash (Wet basis)	-	26.87%	-	
Volatile matter (Dry basis)	-	45.76%	Volatile matter (Dry basis) Fixed Carbon(Dry basis)	74%
Volatile matter (Wet basis)	-	42.70%		
Fixed carbon (Dry basis)	-	25.44%		
Fixed carbon (Wet basis)	-	21.10%		
Heating value (Higher)	-	14524 kJ/kg	Higher heating value	14570 kJ/kg

APPENDIX-C

C. SAMPLE CALCULATION

C.1 Calculation of specific reaction rate

(Reference TABLE : 5.2.3)

Observation No. 9

Air flow rate, $Q = 125$ l/h

Mean temperature, $T_m = (800 + 832)/2 = 816$°C.

Reaction time, $t = 60$ seconds

Particle diameter $d = 0.68$ cm

Initial weight, $W_o = 185.00$ mg.

Final weight, $W_1 = 151.20$ mg.

Specific reaction rate, $q = \dfrac{W_o - W_1}{4\pi(\frac{d}{2})^2 . t}$

(Ref. Equation 6.1.1, Chapter-6)

$$q = \frac{185.00 - 151.20}{4\pi(\frac{0.68}{2})^2 \; . \; 60}$$

$$= 0{\cdot}388 \text{ mg/cm}^2\text{.s}$$

C.2 Calculation of diffusional reaction rate and rate co-efficients:

(Reference TABLE: 6.2.3).

Observation No. - 8

Air flow rate, $Q = 125$ l/h $= 3.48 \times 10^{-5}$ m^3/s.

Mean temperature, $T_m = 775$°C $= 1048$°K

Reaction time, $T = 60$ seconds

Particle diameter, $d = 0.68$ cm.

Cross sectional area of the reactor = 5.067 X 10^{-4} m^2

Air density at N.T.P, = 1.205 kg/m^3

Air viscosity, μ = 18 X 10^{-6} m.s./kg

Mass flow rate, G $= \dfrac{3.48 \times 10^{-5} \times 1.205}{5.067 \times 10^{-4}}$

$= 8.27 \times 10^{-2}$ kg/s.m^2

Reynolds Number, $N_{Re} = \dfrac{G.d}{\mu}$

$= \dfrac{8.27 \times 10^{-2} \times 0.0068}{18 \times 10^{-6}}$

$= 31.24$

At Reynolds number 31.24, multiplying factor for flow system f' = 5.55 (Ref. Appendix-E, Figure E-1).

Oxygen partial pressure, P_g = 0.21 atm.

Diffusional co-efficient at 1048°K, D = 1.75 cm^2/s

(Ref. Appendix-D, Figure D-1).

Universal gas constant, R' = 82.06 atm. cm^3/mol. °K

Mechanism factor, ϕ = 1

(Ref. Section 2.3.2, Chapter-2).

Diffusional reaction rate (Ref. Equation 6.2.3, Chapter-6) $q_{diff} = \dfrac{24 \phi D P_g}{d R' T_m} \times f'$

$= \dfrac{24 \times 1 \times 1.75 \times 0.21}{0.68 \times 82.06 \times 1048} \times 5.55$

$= 8.37 \times 10^{-4}$ gm/cm^2.s

$= 0.837$ mg/cm^2.s

Diffusional reaction rate co-efficient, $K_{diff} = \frac{q_{diff}}{P_g}$
(Ref. Equation- 6.5.1, Chapter-6)

$$= \frac{0.837}{0.21}$$

$$= 3.99 \text{ mg/cm}^2\text{.s.atm.}$$

Overall reaction rate co-efficient, $K_{ov} = \frac{q}{P_g}$
(Ref. Equation 6.4.1, Chapter-6)

$$= \frac{0.370}{0.21}$$

$$= 1.76 \text{ mg/cm}^2\text{.s.atm.}$$

C.3 Calculation of surface reaction rate constant:

(Reference TABLE 6.4.1)

Observation No. 1

Air flow rate, $Q = 60$ l/h

Reaction time, $t = 30$ seconds

Particle diameter, $d = 0.54$ cm

Mean temperature $T_m = 713$°C

Oxygen partial pressure $P_g = 0.21$ atm.

Specific reaction rate, $q = 0.675$ mg/cm^2.s.

Diffusional reaction rate constant, $k_{diff} = 3.46$ mg/cm^2.s.atm.

Diffusional reaction rate, $q_{diff} = 0.726$ mg/cm^2.s.

$$\frac{q}{q_{diff}} = \frac{P_g - P_s}{P_g} \quad \text{(Ref. Equation 6.3.2, Chapter-6)}$$

$$\frac{0.675}{0.726} = \frac{0.21 - P_s}{0.21}$$

$P_s = 0.01475$ atm.

$K_s = \frac{q}{P_s}$ (Ref. Equation 6.3.1, Chapter - 6)

$= \frac{0.675}{0.01475}$

$= 45.9$ mg/cm^2.s.atm.

Surface reaction rate constant, $K_s = \frac{q}{P_s}$

$= 45.9$ mg/cm^2.s.atm.

C.4 Calculation of ash film diffusion constant.

(Reference TABLE 6.4.1)

Observation No. 2

Specific reaction rate, $q = 0.67$ mg/cm^2.s.

Overall reaction rate co-efficient, $k_{ov} = 2.38$ mg/cm^2.s.atm.

Surface reaction rate constant, $k_s = 45.9$ mg/cm^2.s.atm.

Diffusional reaction rate constant, $k_{diff} = 3.46$ mg/cm^2.s.atm.

$$\frac{1}{k_{ov}} = \frac{1}{k_s} + \frac{1}{k_{diff}} + \frac{1}{k_{ash}}$$

(Reference Equation 6.6.1, Chapter-6)

$$\frac{1}{2.38} = \frac{1}{45.9} + \frac{1}{3.46} + \frac{1}{K_{ash}}$$

$$\frac{1}{K_{ash}} = \frac{1}{2.38} - \frac{1}{3.46} - \frac{1}{45.9} = 0.11$$

Ash film diffusion constant, $K_{ash} = 9.09$ mg/cm^2.s.atm.

APPENDIX-D

D. VALUES OF THE BINARY DIFFUSION CO-EFFICIENT

The binary diffusion co-efficient for a pair of gases is independent of the properties of the gases, is inversely proportional to the total pressure, and increases with increase of temperature. The method of Hirschfelder, Curtin & Bird (1954), which is based on a theory of ineraction of molecules assuming a lennard-Joner potential, was found to give agreement usually better than 10%. This method has been employed here to calculate the binary diffusion co-efficients.

TABLE D.A. Binary diffusion co-efficients at 1 atm pressure for O_2- N_2, calculated by the method of Hirschfelder, Curtin & Bird.

Temperature (°K)	Diffusion co-efficient (cm^2/s)
300	0.207
400	0.347
500	0.503
600	0.685
700	0.888
800	1.11
900	1.35
1000	1.61
1100	1.88
1200	2.17
1300	2.47
1400	2.79
1500	3.13

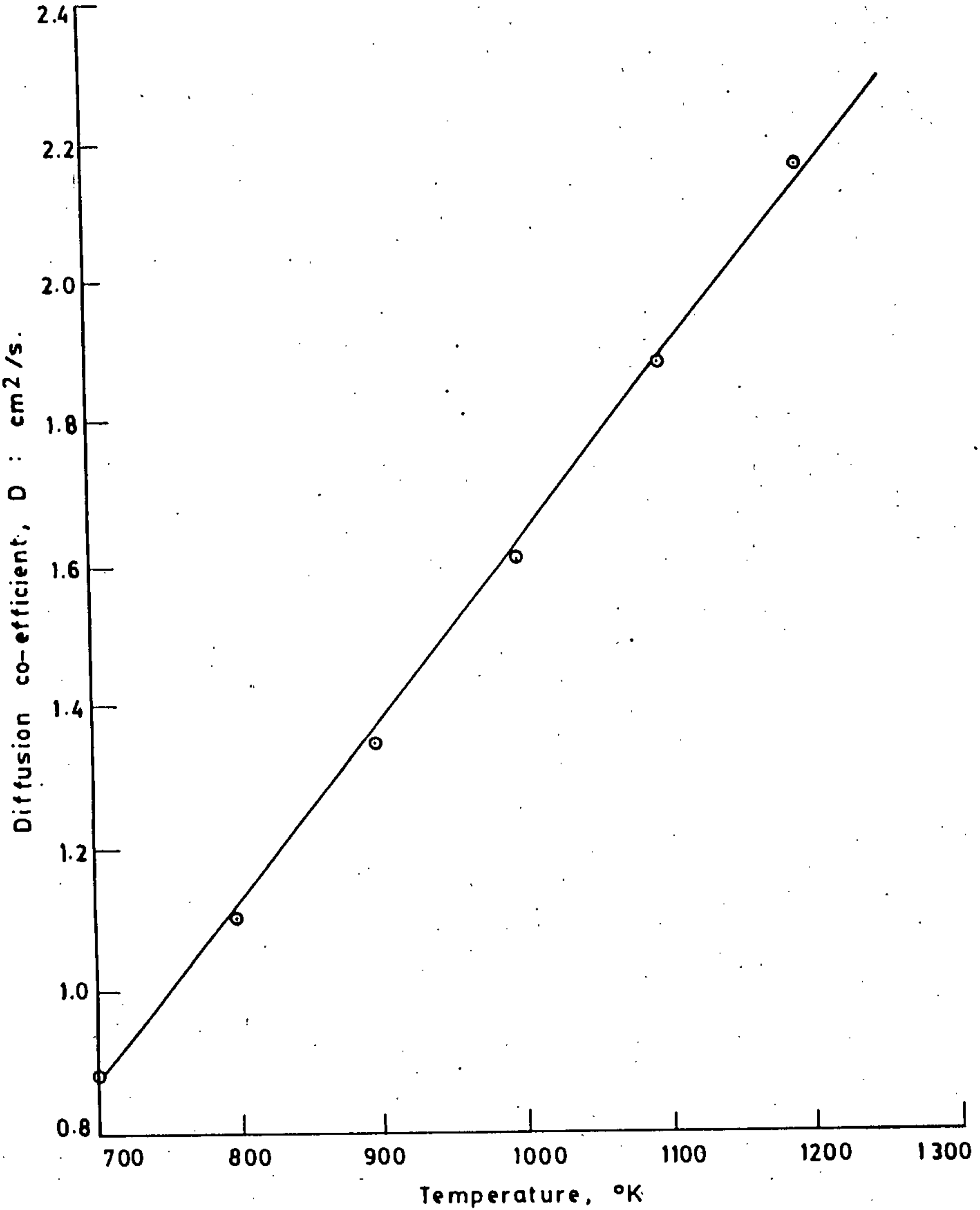

Fig. D-1: Binary Diffusion Co-efficients at 1 atm pressure for O_2-N_2, calculated by the method of Hirschfelder, Curtin & Bird (1954).

APPENDIX-E

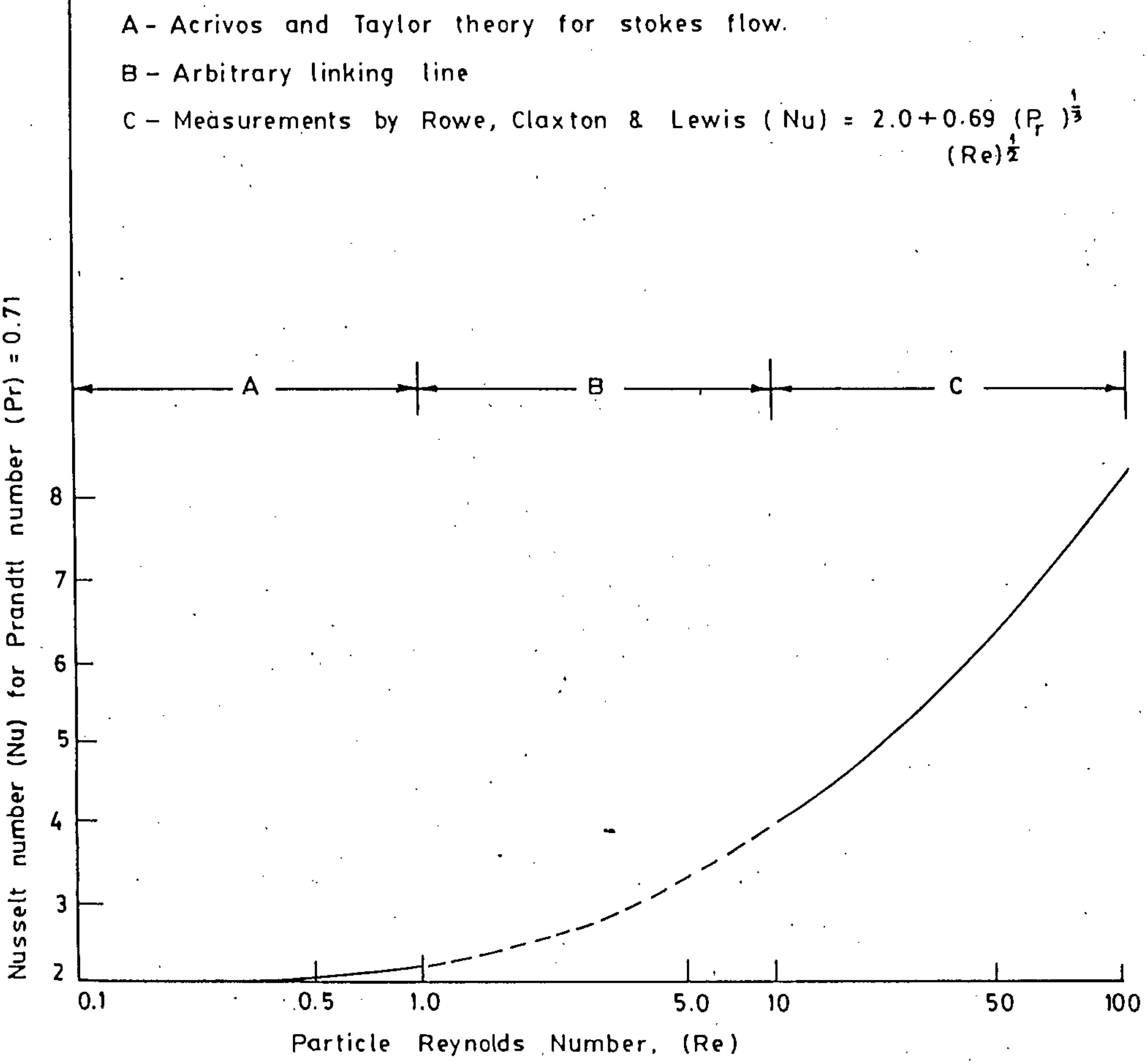

Fig. E-1: Variation of Nusselt (Sherwood) number with particle Reynolds number for constant Prandtl (Schmidt) number of 0.71.

APPENDIX-F

```
FILE: RUN      FORTRAN  A1   BUET COMPUTER CENTRE, DHAKA              VM/SP

C       TO FIND A POLYNOMIAL BY LEAST SQUARE METHOD.
        DIMENSION X(10),Y(10),C(50,50),SUMX(100),SUMY(100),A(50),YY(50)
        NN=9
        OPEN(UNIT=10,STATUS='OLD',FILE='INPUT',ERR=707,FORM='FORMATTED',
       .BLANK='NULL',ACCESS='SEQUENTIAL')
        READ(10,2)(X(I),Y(I),I=1,NN)
  2     FORMAT(2F10.7)
        CLOSE(UNIT=10,STATUS='KEEP')
        NLIMIT=4
        DO 555 NP=1,NLIMIT
        CALL SUMMX(X,Y,NP,NN,SUMX)
        CALL SUMMY(X,Y,NP,NN,SUMY)
        CALL COMAT(SUMX,SUMY,NP,NN,C)
C       NP IS THE HIGHEST POWER OF X IN THE POLYNOMIAL
C       SOLUTION OF LINEAR EQUATIONS BY GAUSS-JORDAN METHOD.
        M=NP+1
        N=M+1
        K=1
        L=1
C       NORMALIZE PIVOT ROW
 15     G=C(K,L)
        DO 10 J=1,N
 10     C(K,J)=C(K,J)/G
C       ZERO THE PIVOT COLUMN
        I=1
 100    B=C(I,K)
        DO 20 J=1,N
        IF(I.EQ.K) GOTO 20
        C(I,J)=C(I,J)-C(K,J)*B
 20     CONTINUE
        I=I+1
        IF(I.GT.M) GOTO 101
        GOTO 100
 101    K=K+1
        L=L+1
        IF(K.GT.M) GOTO 201
        GOTO 15
 201    DO 40 I=1,M
 40     A(I)=C(I,N)
        OPEN(UNIT=3,STATUS='NEW',FORM='FORMATTED',FILE='OUTPUT')
        WRITE(3,55)
 55     FORMAT(9X,'REQUIRED SOLUTION IN SEQUENCE A1,A2,------'//)
        WRITE(3,67)(A(I),I=1,M)
 67     FORMAT(21X,E 15.4/)
        CALL CALYY(A,M,NN,X,YY)
        CALL ERROR(Y,YY,NN,E)
*       WRITE(6,68)
*68     FORMAT(10X,'X',20X,'YSUPPLIED',25X,'YCALCULATED'//)
*       WRITE(6,69)(X(I),Y(I),YY(I),I=1,NN)
*69     FORMAT(9X,F10.5,21X,F10.5,25X,F10.5//)
        WRITE(3,79)E
 79     FORMAT(10X,'SUM OF SQUARES OF ERRORS=',F20.5//)
 555    CONTINUE
 707    WRITE(6,*)'ERROR IN READING DATA FILE'
        STOP
```

```
      END
      SUBROUTINE SUMMX(X,Y,NP,NN,SUMX)
      DIMENSION X(10),Y(10),SUMX(100)
      KK=2*NP
      DO 11 K=1,KK
      SUMX(K)=0.0
      DO 11 I=1,NN
 11   SUMX(K)=SUMX(K)+X(I)**K
      RETURN
      END
      SUBROUTINE SUMMY(X,Y,NP,NN,SUMY)
      DIMENSION X(10),Y(10),SUMY(100)
      K=1
      SUMY(K)=0.0
      DO 14 I=1,NN
 14   SUMY(K)=SUMY(K)+Y(I)
      DO 16 K=1,NP
      SUMY(K+1)=0.0
      DO 16 I=1,NN
 16   SUMY(K+1)=SUMY(K+1)+Y(I)*(X(I)**K)
      RETURN
      END
      SUBROUTINE COMAT(SUMX,SUMY.NP,NN,C)
      DIMENSION SUMX(100),SUMY(100),C(50,50)
      C(1,1)=NN
      L=NP+1
      DO 21 J=2,L
 21   C(1,J)=SUMX(J-1)
      DO 31 I=2,L
      DO 31 J=1,L
 31   C(I,J)=SUMX(I-1+J-1)
      J=L+1
      DO 41 I=1,L
 41   C(I,J)=SUMY(I)
      RETURN
      END
      SUBROUTINE CALYY(A,M,NN,X,YY)
      DIMENSION A(M),YY(NN),X(NN)
      DO 222 I=1,NN
      YY(I)=A(1)
      DO 222 J=2,M
      K=J-1
222   YY(I)=YY(I)+A(J)*(X(I)**K)
      RETURN
      END
      SUBROUTINE ERROR(Y,YY,NN,E)
      DIMENSION Y(NN),YY(NN)
      E=0.0
      DO 333 I=1,NN
333   E=(ABS(YY(I)-Y(I)))**2+E
      RETURN
      END
```

APPENDIX - G

G. PRELIMINARY INVESTIGATIONS ON PEAT SPHERE BURNING

Results of preliminary investigations on the burning of peat spheres in a stream of air at 700°C and 60 l/h is presented in Table G-1. The rate of combustion of solid fuels depends on heat transfer and mass transfer rates, rate of devolatilization, burning rate of volatile matters and of fixed carbon. Comparing Tables G-1 and 5.4.2, it can be concluded that the rate of fixed carbon burning is the limiting stage among the above mentioned factors. For example, data presented in these two tables show that the total time required for heat transfer, devolatilization and volatile matter burning is approximately 1(one) minute whereas about 92% of fixed carbon burns during 10 (ten) minutes in case of 0.88 cm devolatilized peat particle.

TABLE NO. G-1 BURNING OF PEAT SPHERES AT 700°C AND 60 l/h AIR FLOWRATE

Diameter of Peat Spheres cm.	Duration of experiment 's'	Weight loss during reaction $(W_o - W_1)$ 'mg'	Volatiles in Peat Spheres, 'mg'	Comments
1.50	60	766	886	Incomplete burning of volatiles, no fixed carbon burning occurs.
1.30	60	522	555	Incomplete burning of volatiles, no fixed carbon burning occurs.
0.98	60	245	247	Incomplete burning of volatiles.
0.82	60	158	148	Complete burning of volatiles and burning of fixed carbon to some extent.