

Review

# Dust explosions—Cases, causes, consequences, and control

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## Abstract

Dust explosions pose the most serious and widespread of explosion hazards in the process industry alongside vapour cloud explosions (VCE) and boiling liquid expanding vapour explosions (BLEVE). Dust explosions almost always lead to serious financial losses in terms of damage to facilities and down time. They also often cause serious injuries to personnel, and fatalities.

We present the gist of the dust explosion state-of-the-art. Illustrative case studies and past accident analyses reflect the high frequency, geographic spread, and damage potential of dust explosions across the world. The sources and triggers of dust explosions, and the measures with which different factors associated with dust explosions can be quantified are reviewed alongside dust explosion mechanism. The rest of the review is focused on the ways available to prevent dust explosion, and on cushioning the impact of a dust explosion by venting when the accident does take place.

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*Keywords:* Dust explosion; Flammability; Explosion propagation; Vent design; Suppression; Inerting

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## 1. Introduction

### 1.1. Dusts

According to BS 2955: 1958 [1,2], materials with particle size less than 1000  $\mu\text{m}$  (16 BS mesh size) are defined as ‘powders’; when particles have a diameter less than 76  $\mu\text{m}$  (200 BS mesh size), they are referred to as ‘dust’. As per NFPA [3] ‘dust’ is any finely divided solid, 420  $\mu\text{m}$  or less in diameter. Given the nearly six orders-of-magnitude difference between the sizes stipulated by BS 2955 and NFPA 68, it may be safer to follow the somewhat wider view of Palmer [4] which does not exclude from his treatment even particle diameters coarser than 1000  $\mu\text{m}$  [2]. In this write-up we have used the term ‘dust’ for all particulate material, irrespective of the particle size.

More than 70% of dusts processed in industry are combustible [5]. This implies that majority of industrial plants that have dust-processing equipment are susceptible to dust explosions.

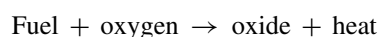
### 1.2. Dust explosion

A dust explosion is initiated by the rapid combustion of flammable particulates suspended in air. Any solid material that can burn in air will do so with a violence and speed that increases with the degree of sub-division of the material [6]. Higher the degree of sub-division (in other words smaller the particle size) more rapid and explosive the burning, till a limiting stage is reached when particles too fine in size tend to lump together. If the ignited dust cloud is unconfined, it would only cause a flash fire. But if the ignited dust cloud is confined, even partially, the heat of combustion may result in rapid development of pressure, with flame propagation across the dust cloud and the evolution of large quantities of heat and reaction products. The furious pace of these events results in an explosion. Besides the particle size, the violence of such an explosion depends on the rate of energy release due to combustion relative to the degree of confinement and heat losses. In exceptional situations a destructive explosion can

occur even in an unconfined dust cloud if the reactions caused by combustion are so fast that pressure builds up in the dust cloud faster than it can be dissipated at the edge of the cloud [7].

The oxygen required for combustion is mostly supplied by air. The condition necessary for a dust explosion is a simultaneous presence of dust cloud of appropriate concentration in air that will support combustion throughout the process and a suitable ignition source. In case of dusts made up of volatile substances, the explosion may occur in three steps which may follow each other in very quick succession—devolatilization (where volatiles are let off by the particle or the particles are vapourized), gas phase mixing of fuel (released by dusts) and oxidant (usually air), and gas phase combustion.

Many combustible dusts if dispersed as a cloud in air and ignited, will allow a flame to propagate through the cloud in a manner similar to (though not identical to) the propagation of flames in premixed fuel–oxidant gases [8]. Such dusts include common foodstuffs like sugar flour, cocoa, synthetic materials such as plastics, chemicals and pharmaceuticals, metals such as aluminum and magnesium, and traditional fuels such as coal and wood. Generally dust explosion involves oxide formation:



But metal dusts can also react with nitrogen or carbon dioxide to generate heat for explosion.

The interdependence of the various parameters which influence the explosion pressure is described by the equation of state for ideal gases:

$$P = \frac{nRT}{V}$$

where  $P$  is the pressure,  $V$  the volume,  $R$  the universal gas constant,  $n$  the number of moles of gas and  $T$  is the temperature. Other factors being equal, the increase of  $T$  due to the heat developed in the burning dust cloud has the deciding influence on the explosion pressure. It follows that higher the heat of combustion of a given dust per mole of  $O_2$  consumed, greater is the likely severity of an explosion. Table 1 presents the heat of combus-

tion of a few commonly encountered dusts. It reveals that metals form highly hazardous dusts from this viewpoint [6,9].

Explosion hazard always exists whenever dusts are produced, stored or processed and where situations can occur when these materials are present as a mixture in air [10,11]. The mixture is deemed ‘explosible’ if combustible dusts are present in such quantities in air that an explosion can occur on ignition.

The differences between the mechanism of combustion of dust clouds and premixed gases has been elaborated in Section 6.

## 2. Illustrative case histories of a few major dust explosions

Even though mention of dust explosions is found in literature since 1785 [6,12], systematic records are available only from the early 20th century. One of the earliest recorded and the most serious of the accidents triggered by dust explosion occurred at Leiden, the Netherlands, on 12 January 1807 [13]. A ship with about 85,000 kg of black powder on board had arrived from Ouderkerk (near Amsterdam) and had moored in the centre of Leiden against all regulations. It is believed that the attempts of the four-member crew to prepare their food on a stove ignited the dust leading to a detonation estimated to be equivalent to 9000 kg of exploding TNT. The explosion killed 151 and wounded about 2000. Houses collapsed up to a distance of 155 m from the ship and within the whole city people were hit by flying debris, glass, and roof tiles. Another catastrophic dust explosion of pre-modern era involved grain dust at the Peavey terminal elevator at Duluth, the USA, in 1916. After the explosion, the cribbed grain bins caught fire, completely destroying the elevator. It was described as one of the ‘worst roaring infernos’ witnessed in a dust explosion [14].

In 1919, a dust explosion occurred in a corn processing plant in Iowa, the USA, killing 43. Five years later, a dust explosion at a similar plant in Illinois (the USA) left 42 dead [5].

In 1949, at the Port Colbourne elevator in Ontario, Canada, an explosion occurred in the steel bins which blew off the entire roof. The ensuing fire also caused significant damage to the head house. A grain dust explosion ripped through the grain elevator no. 4A of the Saskatchewan grain pools on 24 September 1952, killing 6 and injuring 14. The primary explosion in a shipping bin was followed by a secondary explosion involving large quantities of dust, which had been allowed to accumulate in the building. The roof gallery above the bins was also destroyed [14].

Dust explosions occur frequently in feed mills. A corn dust explosion in the Wayne Feeds at Waynesboro, the USA, on 25 May 1955, killed 3 and injured 13. The violence of the explosion caused extensive property damage. Another severe explosion occurred in Kansas City, the USA, in 1958, when the Murray elevator was badly damaged. The head house of steel construction was completely shattered and its installations destroyed by fire [14].

A dust explosion, which was caused during the welding of a spout, excessively worn by the flow of grain, occurred in the Kampffmeyer grain silo at Albern near Vienna, Austria, 4 July 1960. The welding was being performed in the elevator pit, when a spark ignited the dust in the running bucket elevator. The pres-

Table 1  
Heat of combustion of some common dusts [6]

Material	Oxidation products	Heat of combustion (kJ/mol $O_2$ )
Calcium	CaO	1270
Magnesium	MgO	1240
Aluminum	$Al_2O_3$	1100
Silicon	$SiO_2$	830
Chromium	$Cr_2O_3$	750
Zinc	ZnO	700
Iron	$Fe_2O_3$	530
Copper	CuO	300
Sucrose	$CO_2 + H_2O$	470
Starch	$CO_2 + H_2O$	470
Polyethylene	$CO_2 + H_2O$	390
Carbon	$CO_2$	400
Coal	$CO_2 + H_2O$	400
Sulphur	$SO_2$	300

sure wave of the explosion went up through the elevator shaft, ripping the casing of the elevator leg, and continuing up to the roof, causing severe damage to the building and machinery [14].

At the Sun flourmills in London, UK, on 7 August 1965, 4 men died and 37 were injured when a giant blast shattered and set ablaze the mill building and a wheat storage silo of cribbed construction. The explosion is believed to have been triggered by a welding flame being used on a flour bin [14].

A violent dust explosion on 14 December 1970, shattered the grain silos at Kiel-Nordhafen on the Kaiser Wilhelm shipping canal, connecting the North and the Baltic seas. It was the worst accident of its kind in Germany. Six men died and 17 were injured. The damage to plant, building and machinery is estimated at 10 million dollars. At Destrahan near New Orleans, USA, a Bunge Corporation terminal elevator with an 8,000,000 bushel capacity was badly damaged in a massive dust explosion. The entire roof gallery above the storage tanks was blown off. The heat from the explosion and the resultant fire badly damaged the concrete storage bins and the adjacent workhouse [14].

In a dust explosion in a silicon powder grinding plant at Bre-manger, Norway, in 1972, five workers lost their lives and four were severely injured. The explosion, which occurred in the milling section of the plant, was extensive, rupturing or buckling most of the process equipment and blowing out practically all the wall panels of the factory building [6,15].

A blast and flame from a primary explosion, which occurred in a 5.2 m<sup>3</sup> batch mixer when fine aluminum flakes, sulphur and some other ingredients were being mixed at a slurry explosive factory in Norway in 1973, then generated and ignited a larger dust cloud. The massive secondary explosion killed half of the 10 workers that were on site at that time, seriously injuring 2 others. A substantial part of the plant was totally demolished [6].

A catastrophic explosion at a large export grain silo plant at Christi, TX, USA, on 11 April 1981, killed 9 persons and injured another 30. The material loss was estimated at \$30 million [16]. The probable cause of ignition was smoldering lumps of sorghum that entered a bucket elevator together with the grain and ignited the dust cloud in the elevator.

Another catastrophic dust explosion which occurred at the Harbin Linen Textile Plant, People's Republic of China on 15 March 1987, killed 58 persons and injured another 177. It destroyed 13,000 m<sup>2</sup> of factory area [17,18]. The ignition was possibly caused by an electrostatic spark in one of the dust collecting units. The explosion then propagated through the other seven dust collecting units, demolishing most of the plant.

Dry BPO (benzoylperoxide) exploded at a BPO manufacturing plant of the Dai-ichi Kasei Kogyo Company, Japan, in August 1990. Nine workers were killed, and 17 were injured in the accident. The storage of an illegally large quantity of dry BPO apparently resulted in significantly more damage than would otherwise have occurred [19], because BPO can deflagrate even without being dispersed in air.

A dust explosion at the Daido Kako Enka Firework manufacturing factory, Moriya, Japan, in June 1992 resulted in the death of 3 persons and injury to 58, including passers by. The presumed cause of this accident was the ignition of powdery mixture of

potassium chlorate and aluminum by friction-induced sparks during a mixing operation [19].

At a textile mill of the Shinko Seishoku Company at Okaharu, Japan, a fire, caused by the ignition of dusty cotton waste as a result of electricity leak, destroyed the factory in December 1994. In the same year, a large dust explosion at the Kanaya Shoe making factory, near Tokyo, killed 5 and injured 22. The cause of the explosion was static electricity generated in the fine rubber waste dust [5].

Blaye, France, witnessed a massive dust explosion in a grain storage facility during August 1997. Two towers which housed bucket elevators and dust collection equipment were totally destroyed; so were a gallery and 28 silos. The concrete debris rained down the control room, killing 11 persons. The most worrisome aspect of this catastrophe is that the whole grain handled by the factory is generally regarded as posing low risk. The explosion properties likewise seemed modest; samples passed a 500  $\mu\text{m}$  sieve; they had  $P_{\text{max}}$  (maximum absolute explosion pressure) values of just 6 bar and low  $K_{\text{St}}$  (maximum rate of pressure rise) values as well [20].

A tantalum dust deflagration occurred in a bag filter dust collecting device in April 1997, which resulted in a fatal accident. One worker was killed, and another seriously wounded. A Mg–Al alloy dust exploded in a bag filter dust collecting device at a manufacturing plant of electronic devices in October 2000. One worker was killed and another injured [21].

On 1 February 1999, explosions in a powerhouse of the Ford Motor Company in Michigan caused the death of 6 workers and injured 14 others. The primary explosion was found to have been caused due to natural gas build-up in a boiler that was being isolated for maintenance. It has been suggested that it was a result of the secondary explosions involving coal dust which wrought the main damage. The powerhouse building and connected facilities were widely damaged. At over \$1 billion, it is one of the most expensive accidents in the history of the USA [22].

An explosion in a mould fabrication station in the Jahn Foundry, Massachusetts, injured 12 employees on 25 February 1999, three of whom died later of the burns [23]. Either the accumulation of natural gas/air mixture in one of the ovens or the ignition of an airborne cloud of combustible resin dust in the hot oven is thought to be the cause of the explosion.

Five workers died from severe burns suffered from an explosion in a rubber recycling plant of Rouse Polymerics International Inc. in Mississippi on 16 May 2002. Sparks leaving an oven exhaust pipe are believed to have caused the initial explosion, which, then, triggered a secondary explosion of the dust accumulated in the building. The rubber dust was of very small size (75–180  $\mu\text{m}$ ), contributing to its very high ignition hazard [24].

On 20 February 2003, an explosion and fire damaged the CTA Acoustics manufacturing plant in Corbin, KY, the USA, fatally injuring seven workers. The explosion was caused by the accidental ignition of resin dust accumulated in a production line that had been partially shut down for cleaning. Apparently a thick cloud of dust dispersed by cleaning activities was ignited by the flames in an oven whose door had been left open. The resulting explosion propagated through the facility, causing secondary

explosions as powdered resin dust was dislodged from surfaces, adding to the airborne fuel loading [25].

An explosion and fire involving polyethylene dust killed 6 workers and injured 38 others at the USA-based West Pharmaceuticals (Kinston, NC) on 29 January 2003. Two firefighters were among those killed in a massive blast of which impact was felt over a large area; the burning debris triggered secondary fires up to 2 miles away. Some initiating event caused dust to become airborne above a suspended ceiling. There it contacted an ignition source leading to the catastrophic event [25].

According to Marmo et al. [26] a series of ‘very serious’ dust explosion accidents occurred in an industrial district in north-

ern Italy over a period of 7 years (1994–2000) causing deaths, several injuries, and great material damage.

Schoeff [27] has noted that on an average more than 10 major dust explosions have been occurring in the agricultural factories in the USA. In 2005, 13 such explosions took place killing 2 persons, injuring 11, and causing property damage worth \$56 million.

An illustrative list of dust explosions is presented in Table 2. Indeed it is commonly admitted [8] that at least one dust explosion occurs in each industrialized country every day!

Surprisingly, whereas copious information is available on dust explosions in the developed countries, there is practically

Table 2  
Illustrative examples of dust explosion incidents (1911–2004)

Date	Location	Material	Plant/building	Dead/injured	Reference
1785	Turin, Italy	Wheat flour	Bakery	2i	[6]
1807	Leiden, The Netherlands	Black powder	Ship	151d/2000i	[13]
1911	Glasgow, UK	<sup>a</sup>	<sup>a</sup>	5d/8i	[5]
1911	Liverpool, UK	<sup>a</sup>	<sup>a</sup>	37d/100i	[5]
1911	Manchester, UK	<sup>a</sup>	<sup>a</sup>	3d/5i	[5]
1913	Manchester, UK	<sup>a</sup>	<sup>a</sup>	3d/5i	[5]
1916	Duluth, MN	Grain	Steel bin	–	[14]
1919	Cedar Rapids, IA	Corn starch	Starch plant	43d	[5]
1924	Peking, IL	Corn starch	Starch plant	42d	[5]
1924	USA	Sulphide dust	<sup>a</sup>	1d/6i	[5]
1924	USA	Sulphide dust	<sup>a</sup>	1d/1i	[5]
1924	USA	Sulphide dust	<sup>a</sup>	2d/1i	[5]
1926	USA	Sulphide dust	<sup>a</sup>	3d/1i	[5]
1930	Liverpool, UK	<sup>a</sup>	<sup>a</sup>	11d/32i	[5]
1944	Kansas City, KS	Grain dust	<sup>a</sup>	<sup>a</sup>	[5]
1949	Port Colbourne, CA	Grain	Steel bin	–	[14]
1952	Bound Brook, NJ	Phenolic resin dust	Hammer mill	5d/21i	[5]
1952	Saskatchewan	Grain dust	Shipping bin	6d/14i	[14]
1955	Waynesboro, GA	Grain dust	Feed plant	3d/13i	[14]
1956	South Chicago	Grain dust	Elevator	–	[14]
1958	Kansas City	Grain dust	Elevator	–	[14]
1960	Canada	Sulphide dust	<sup>a</sup>	2d/–	[5]
1960	Albern, Vienna	Grain dust	<sup>a</sup>	–	[14]
1962	St. Louis, MO	Grain dust	Feed plant	3d/13i	[5]
1964	Paisley, UK	<sup>a</sup>	<sup>a</sup>	2d/34i	[5]
1965	London, UK	Flour	Flour mill	4d/37i	[14]
1969	Sweden	Sulphide dust	<sup>a</sup>	2d/1i	[5]
1970	Kiel, FRG	Grain dust	Grain silo	6d/18i	[5]
1970	Germany	Grain dust	Silos on shipping canal	6d/17i, loss \$10 million	[14]
1970	Norway	Wheat grain dust	Silo	<sup>a</sup>	[6]
1971	New Orleans	Bushel	Elevator	<sup>a</sup>	[14]
1972	Norway	Silicon	Milling section	5d/4i	[6,15]
1973	Norway	Aluminum	Mixing vessel	5d/2i	[6]
1974	Canada	Sulphide dust	Fox mines	<sup>a</sup>	[5]
1974	Preska, South Africa	Sulphide dust	Mines	<sup>a</sup>	[5]
1975	Norway	Fish meal	Fish meal grinding plant	1d/1i	[6]
1976	Norway	Barley/oats dust	Silo	–	[6]
1976	Oslo, Norway	Malted barley dust	Silo	–	[6]
1977	Galvesto, TX	Grain dust	Grain silo	15d	[5]
1977	Westwego, Louisiana	Grain dust	Grain silo	36d/10i	[15]
1979	Lerida, Spain	Grain dust	Grain silo	7d	[5]
1979	Canada	Sulphide dust	Ruttan mines	<sup>a</sup>	[5]
1980	Germany	Coal	Cement factory	–	[6]
1980	Iowa, USA	Corn dust	Bucket elevator	–	[16]
1980	Minnesota, USA	Grain dust	Cross tunnel, bucket elevators	13i	[16]
1980	Naples, Italy	Grain dust	Grain silo	8i	[5]
1980	Ohama, NE, USA	Grain dust	Head house	Loss \$3,300,000	[5]
1980	St. Joseph, MO, USA	Grain dust	Shipping bin	1d/4i, loss \$2,000,000	[16]

Table 2 (Continued)

Date	Location	Material	Plant/building	Dead/injured	Reference
1981	Canada	Sulphide dust	Mattabi mines	<sup>a</sup>	[5,15]
1981	Corpus Christi, TX	Grain dust	Bucket elevator Elevator	9d/30i	[16]
1981	Bellwood, NE, USA	Grain dust	Bucket elevator	Loss \$6,400,000	[5]
1981	Germany	Coal	Coal dust burner plant, cement works	–	[6]
1982	British Columbia, Canada	Coal	Silo	–	[6]
1983	Anglesey, UK	Aluminum	Aluminum powder production	2i	[6]
1984	USA	Coal	Silo	–	[6]
1985	Australia	Sulphide dust	Elura mines	<sup>a</sup>	[15]
1985	Canada	Sulphide dust	Lynn lake	<sup>a</sup>	[5]
1985	Germany	Coal	Silo	1i	[6]
1985	Norway	Rape seed flour pellets	Silo	–	[6]
1986	Canada	Sulphide dust	Brunswick mines	<sup>a</sup>	[5]
1986	Sweden	Sulphide dust	Langsele mines	<sup>a</sup>	[5]
1986	Canada	Sulphide dust	Dumugami mines	<sup>a</sup>	[5]
1986	Australia	Sulphide dust	Woodlawn	<sup>a</sup>	[5]
1987	Canada	Sulphide dust	GECO mines	<sup>a</sup>	[5]
1987	China	Textile dust	Dust collection system	58d/177i	[17,18]
1987	Oslo, Norway	Malted barley dust	Silo	–	[6]
1988	Norway	Wheat grain dust	Silo	–	[6]
1988	Sweden	Coal	Silo	–	[6]
1989	Sweden	Palletized wheat bran	Silo	–	[6]
1990	Japan	Benzoylperoxide	Storage	9d/17i	[19]
1992	Moriya, Japan	Potassium chlorate and aluminum dust	Mixing operation	3d/58i	[19]
1994	Okaharu, Japan	Cotton waste	Textile mill	<sup>a</sup>	[5]
1994	Tokyo, Japan	Rubber waste	Shoe factory	5d/22i	[5]
1997	Japan	Tantalum dust	<sup>a</sup>	1d/1i	[21]
1997	Blaye, France	Grain	Storage	11d	[20]
1999	Michigan	Coal dust (cause for secondary explosion)	Powerhouse	6d/14i	[22]
1999	Massachusetts	Resin	Oven	3d/12i	[23]
2000	Japan	Mg–Al alloy		1d/1i	[5]
2000	Modesto California	Aluminum dust	<sup>a</sup>	<sup>a</sup>	[15]
2002	Mississippi	Rubber	Recycling plant	5d/ <sup>a</sup>	[24]
2003	Kentucky	Resin	Production line	7d	[25]
2003	Kinston, NC	Polyethylene	Pharmaceutical plant	6d/38i	[25]
2004	Avon, OH	Lacquer dust	<sup>a</sup>	<sup>a</sup>	[15]

<sup>a</sup> Details not available.

no material in printed or electronic form on dust explosions in the developing countries. This is surprisingly so even in case of a technologically advanced country like India from where, otherwise, significant original contributions have come to loss prevention R&D [28].

Indeed when the subject of accidents in chemical process industries is discussed in the third world, the focus is mainly on toxic releases such as the one that occurred in Bhopal in 1984 [29,30] or on fires caused by vessel bursts such as the ones that occurred in 1996 at the HPCL refinery in Vishakapatnam [31] and in 2004 at Libya. The terms vapour cloud explosion (VCE), boiling liquid expanding vapour explosion (BLEVE), pool fire, flash fire, fireball are heard most of the time during discussions on process industry accidents [32–36]. In comparison, the phenomena of dust explosion is much less deliberated upon or cognized in the third world. This is highly worrisome because some of the most damaging accidents in chemical and agro processing industries have been caused not by flammable liquids or gases but by ‘dusts’. As would be reflected from the case studies

reported later in this paper, multiple fatalities and catastrophic damage to property is a common feature of accidents involving dust explosions.

Even though, as stated earlier, India is among the most advanced countries in terms of technological competence, information pertaining to dust explosions occurring here is almost non-existent [37] solely because in most accidents that occur in India, the broad term ‘explosion’ is used and recorded while the type of explosion generally goes unreported. Also, in public perception, explosions are what occur in pressurized vessels containing gases or liquids, or by the operation of explosives. Dust is seldom perceived as a cause for explosions. This appears to be typical of the third world. This situation is quite different from the one existing in developed countries where dust explosions are recognized as very major industrial hazards that can match or exceed the ferocity of well-known industrial disasters like the one that occurred in Flixborough [38], and meticulous attention is paid towards their analysis, prevention, and control.

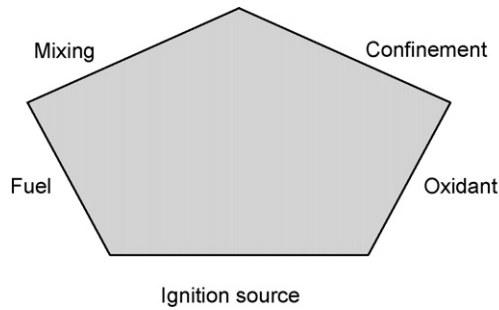


Fig. 1. The dust explosion pentagon [39].

### 3. The dust explosion pentagon

While fire is caused when three factors – fuel, oxidant, and ignition – come together to make what has been called ‘the fire triangle’, a dust explosion demands two more factors: mixing (of dust and air), and confinement (of the dust cloud). The ‘dust explosion pentagon’ [39] is formed when these five factors occur together (Fig. 1):

- (i) presence of combustible dust in a finely divided form;
- (ii) availability of oxidant;
- (iii) presence of an ignition source;
- (iv) some degree of confinement;
- (v) state of mixed reactants.

A point to be noted here is that even partial confinement of an ignited dust cloud is sufficient to cause a highly damaging explosion. In this sense, too, dust clouds behave in a manner similar to clouds of flammable gases [8].

### 4. Sources and triggers of dust explosions

The industries prone to dust explosions are:

- (i) wood processing and storage;
- (ii) grain elevators, bins and silos;
- (iii) flour and feed mills;
- (iv) manufacture and storage of metals such as aluminum and magnesium;
- (v) chemical production;
- (vi) plastic production;
- (vii) starch or candy production;
- (viii) spice sugar and cocoa production and storage;
- (ix) coal handling or processing area;
- (x) pharmaceutical plants.

In addition, dust collection bins or bags, shelves, nooks, cranies, inside of equipment, and above the false ceilings, in all the aforementioned facilities, are prone to dust explosion.

In recent years explosion hazard from polyurethane dust generated in the course of recycling depreciated electronic appliances has increased owing to the increasingly large volumes being recycled [40,41].

### 4.1. Operations involving dusts

The operation in which dusts are generated or handled, comprise of:

- (i) size reduction;
- (ii) conveying—manual or mechanical;
- (iii) pneumatic separation;
- (iv) settling chambers;
- (v) cyclones;
- (vi) filters;
- (vii) scrubbers;
- (viii) electrostatic precipitator;
- (ix) driers;
- (x) tray driers;
- (xi) rotary driers;
- (xii) fluidized bed driers;
- (xiii) pneumatic driers;
- (xiv) spray driers;
- (xv) screening and classifying operations;
- (xvi) mixing and blending operations;
- (xvii) storage;
- (xviii) packing;
- (xix) dust fired heaters.

Charge chutes are often used to feed powder into process vessels by manually tipping the powder from sacks or bags into the charge chute. In the event of a dust explosion occurring in the vessel during the filling operation, a flame could propagate up the charge chute, impinge on the operator, and cause serious injury [42].

### 4.2. Classification of dusts

A dust layer is deemed ‘combustible’ if it can be ignited with a foreign source and the local fire thus generated propagates sufficiently after the outside source is taken away [43].

All explosible dusts ought to be combustible, but not all combustible dusts are easily explosible [5]. For example, anthracite and graphite are not easily explosible, although they have high heats of combustion. As of now sufficiently rigorous theory is not available to forecast explosibility of combustible dusts and recourse has been taken to experiments. If the composition of the dust is known, one may check whether it is explosible by consulting the list of experimentally tested dusts, published by HM Factory Inspectorate of the Department of Employment, UK. According to this classification [20], dusts, which propagated a flame when ignited have been classified under Group A. The dusts, which did not propagate a flame have been classified under Group B. This classification is applicable to dusts which are at or near the atmospheric temperature (25 °C) at the time of ignition. At higher temperature some of the Group B dusts can become explosible. Dusts which are ignitable but not explosible can become explosive if admixed with fuel dust; for example the ignitable but non-explosible fly ash becomes explosible when spiked with pulverized coal or petroleum coke [44]. This occurs due to increased volatile matter provided by fuel dusts.

Another measure of the ignitability of a dust layer and intensity of burning of a dust layer is the Combustion Class [45,46]. This classification is based on the behavior of a defined heap when subjected to a gas flame or hot platinum wire:

- (i) CC1: no ignition; no self-sustained combustion.
- (ii) CC2: short ignition and quick extinguishing; local combustion of short duration.
- (iii) CC3: local burning or glowing without spreading; local sustained combustion but no propagation.
- (iv) CC4: spreading of a glowing fire; propagation smoldering combustion.
- (v) CC5: spreading of an open fire; propagating open flame.
- (vi) CC6: explosible burning; explosive combustion.

A third categorization of dusts is based on ‘ $K_{St}$  value’; the term represents the maximum rate of pressure rise in 1 m<sup>3</sup> vessel when a dust is ignited; in other words the ‘dust explosion violence’ [47]. The  $K_{St}$  concept was introduced by Bartknecht [48,49] who reported that the so-called cube root law:

$$\left(\frac{dP}{dt}\right)_{\max} V^{1/3} = \text{constant} \equiv K_{St},$$

seemed to hold for numerous dusts in vessel of volumes from 0.04 m<sup>3</sup> and upward. The  $K_{St}$  value (bar m/s), being numerically identified with the  $(dP/dt)_{\max}$  (bar/s) in the 1 m<sup>3</sup> standard International Standards Organization (ISO), test [50] was denoted ‘a specific dust constant’.

The abbreviation ‘St’ has its origin in the German word *staub*, meaning *dust*.

The explosibility is ranked as under:

$K_{St}$	= Group St0: Non-explosible	
$0 < K_{St} < 200$	= Group St1 weak	Increasing Explosibility ↓
$200 < K_{St} < 300$	= Group St2 strong	
$300 < K_{St}$	= Group St3 very strong	

But it must be emphasized that the ‘cube root law’ is valid only in geometrically similar vessels, if the flame thickness is negligible compared to the vessel radius, and if the burning velocity as a function of pressure and temperature is identical in all volumes [6]. Hence,  $K_{St}$  is bound to be an arbitrary measure of dust explosion violence as the state of turbulence to which it refers is arbitrary. This fact has sometimes been neglected when discussing  $K_{St}$  in relation to industrial practice and is therefore understood here. Indeed  $K_{St}$  values of a given material, determined in different apparatus, may differ by several orders-of-magnitude, even by factors more than 20 [6].

It follows that when using  $K_{St}$  values to size vent areas and for other purposes according to various codes, it is absolutely essential to use only data obtained from the standard test method specified for determining  $K_{St}$ . Normally this is the method of the ISO [50] or a smaller-scale method calibrated against the ISO method. In addition, it is necessary to appreciate the relative and arbitrary nature even of these  $K_{St}$  values.

The Bureau of Mines has developed an index of explosibility which ranks dusts relative to Pittsburgh coal. The index of

explosibility IE is the product of the explosion severity ES and the ignition sensitivity IS:

$$IE = IS \times ES$$

$$IS = \frac{(\text{MIT} \times \text{MIE} \times \text{MEC})_{\text{Pc}}}{(\text{MIT} \times \text{MIE} \times \text{MEC})_{\text{sample}}}$$

$$ES = \frac{(\text{MEP} \times \text{MRPR})_{\text{Pc}}}{(\text{MEP} \times \text{MRPR})_{\text{sample}}}$$

where MEC is the minimum explosive concentration, MEP the maximum explosion pressure, MIE the minimum ignition energy, MIT the minimum ignition temperature, and MRPR is the maximum rate of pressure rise; the subscripts Pc and *sample* denote Pittsburgh coal and sample. This index of explosibility is a relative one, and is to this extent less dependent on the apparatus used, but its determination requires the conduct of the full range of tests [2,12].

#### 4.3. Tests to determine the explosibility of dusts

As all the initiatives on the understanding, prevention, and control of dust explosions revolve round ‘dust explosibility’, ‘minimum explosible dust concentration’, ‘minimum ignition energy’, and ‘minimum ignition temperature’, it may be relevant to dwell upon how these parameters are measured and what are the uncertainties involved in the measurements.

The two apparatus most often used for dust explosibility testing have been the ‘Hartmann vertical tube’ and the ‘20 l sphere’. Of these the Hartmann tube was the first to be commonly used and a great deal of data exists which has been generated in the pre-1980 era by this apparatus [2,12]. A Hartmann apparatus consists of a 1.2 l vertical tube in which dust is dispersed by an air blast. A hot wire or a spark igniter serves as ignition source (Fig. 2). Flame propagation is observed as a function of dust particle size, dust concentration, ignition energy, temperature, etc.

Even as the Hartmann vertical tube and its variants – the horizontal tube, and the inflammatory apparatus – have been extensively utilized in the past, it has been increasingly realized that the Hartmann tube is not apt to give uniform conditions for dust dispersion and turbulence. Further, it is subject to wall effects; after the flame goes through initial spherical expansion, it travels as two fronts up and down the tube. These conditions give a lower rate of combustion and of pressure rise than the actual; consequently the strength of the pressure rise one records with the Hartmann bomb is less than one gets from more advanced apparatus. The Hartmann tube may also yield false negatives for dusts that are difficult to ignite with a spark but are ignitable by stronger ignition sources [7]. Choi et al. [203,204] have proposed a new apparatus, as a modification of the Hartmann tube, in which an ultrasonic vibrator, a sieve, and a specially designed dust hopper are used to generate dust cloud and to ignite it. But the apparatus is yet to be rigorously tested.

These problems have been largely overcome by the use of sufficiently large spherical test vessels. In such vessels a dust cloud is simulated better than in tubular apparatus. The two principal vessels adopted are the nearly spherical ‘20 l sphere’



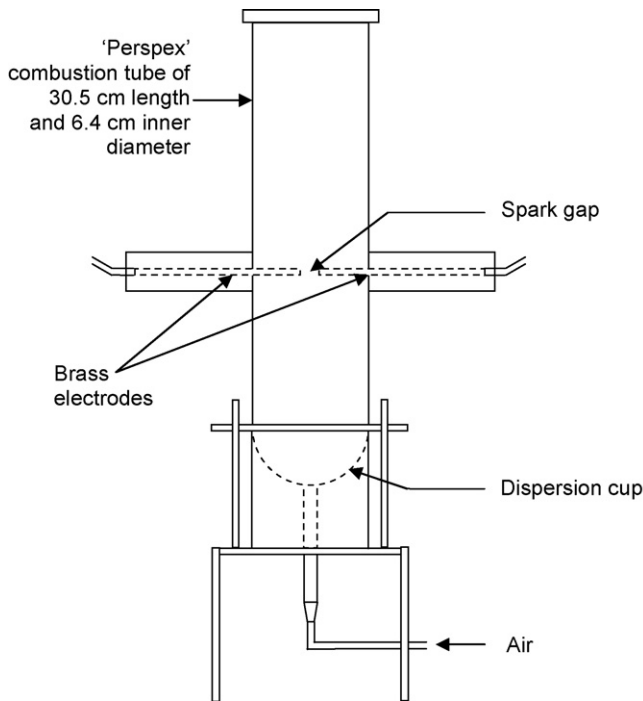


Fig. 2. Hartmann vertical tube apparatus.

introduced by Siwek [51,52] and the standard closed 1 m<sup>3</sup> ISO vessel [6,50]. It has been shown that the former is close to the critical size below which vessel size begins to seriously influence the explosibility measurements and above which such influence is less pronounced. A large number of studies continue to be done on the comparison of results obtained with 20 l and 1 m<sup>3</sup> vessels [7,53–58] and the 20 l sphere is being used increasingly as a standard with or without minor modifications in the apparatus introduced by Siwek [59–62]. In the spherical vessels the ignition source is located in the centre of the sphere and the dust is injected from a separate container (Fig. 3). As with the modified versions of the Hartmann vertical tube [203,204], the 20 l and 1 m<sup>3</sup> spheres are used to determine whether a dust is explosible and to measure the maximum explosion pressure as also the rate of the pressure rise. The minimum explosible dust concentration and the minimum explosion energy are also determined using 20 l/1 m<sup>3</sup> spheres.

The experimental conditions required to obtain agreement with the 1 m<sup>3</sup> ISO vessel were specified in a standard issued by the American Society for Testing and Materials (ASTM) in 1988 [6]. The ignition source has to be the same type of 10 kJ chemical ignitor as used in the 1 m<sup>3</sup> ISO test but the ignition delay can be shorter (60 ms) because of the smaller vessel size. The ignitors with metal capsules could give significantly different  $K_{St}$  values from those obtained for the same dusts with plastics capsules.

With dusts of small particle size, Siwek [51,52] obtained quite good correlations between data from the 1 m<sup>3</sup> ISO vessel and that from his 20 l sphere. It was however seen that many cohesive dusts, in particular those of fibrous particles, can easily get packed and trapped inside the perforated dispersion tube of the original dust dispersion system (Fig. 3b). This led to the development of an open nozzle system named *rebound* nozzle, which

has gradually replaced the original perforated ring in the Siwek 20 l spheres [6]. The rebound nozzle produces both maximum pressures and  $K_{St}$  values in reasonable agreement with those generated by the original perforated-ring system [6,43].

Besides the size and the geometry of test vessel – which, as stated above, should be at least 20 l and spherical, respectively – the ignition source strength should also be appropriate if realistic estimates of dust explosibility are to be obtained. If the ignition source is too large compared to the vessel volume, ‘overdriving’ of the explosion may occur. Overdriving may increase the temperature of the dust cloud, rendering a non-explosible dust explosible. It may also result in burning of the dust within the igniter flame, but with no real propagation beyond the ignition source [63]. The large volume of the ignition source in the context of the vessel volume may make the dust appear to explode, and result in the overestimation of the overpressure as well as the rate of pressure rise.

For several dusts, Going et al. [57] found that best agreement occurred between 20 l vessel data with 2.5 kJ igniters and 1 m<sup>3</sup> vessel data with 10 kJ igniters. In a series of studies on explosion suppression by Amyotte [63] it was noted that the results from 20 l vessel begin to approach those measured in 1 m<sup>3</sup> chamber when the ignition energy is decreased below 5 kJ in the former.

Whereas a large amount of the earlier work on dust explosibility was done using the Hartmann apparatus, most later work has been with the 20 l and 1 m<sup>3</sup> vessels. Unfortunately, the results obtained from the two types of apparatus often do not agree and may even give different rankings. For example the Hartmann vertical tube overestimates the minimum ignition energy by a factor of between 2 and 5, and is therefore not conservative. The ignition source in the Hartmann tube is a capacitive spark igniter, which has two disadvantages: the spark energy tends to be less than the theoretical energy ( $1/2VC^2$ ) due mainly to loss in the transformer, and it is not possible to control the duration of the spark, which is a significant variable. Due to these reasons nearly all of the very considerable explosibility data generated earlier with the original Hartmann apparatus is not utilizable.

#### 4.4. Safety codes

A number of safety codes now address the dust/vapour explosion potential depending on the type of industry or operations [64]. An illustrative example, are the National Fire Protection Association (NFPA) codes:

- (i) Combustible metals and metal dusts (NFPA 65, 480, 481).
- (ii) Explosion protection systems (NFPA 68, 69).
- (iii) Handling and conveying of dusts, vapour, and gases (NFPA 91, 650, 654, 655).
- (iv) Prevention of sulphur fires and explosions (NFPA 655).
- (v) Prevention of fires and explosions in wood processing and woodworking facilities (NFPA 664).

The ‘Atex 100a’ Directive of the European Parliament [65] provides the conceptual basis for the European apparatus standards for prevention and mitigation of accidental gas, vapour, mist, and dust explosion. It has also had a major impact on the

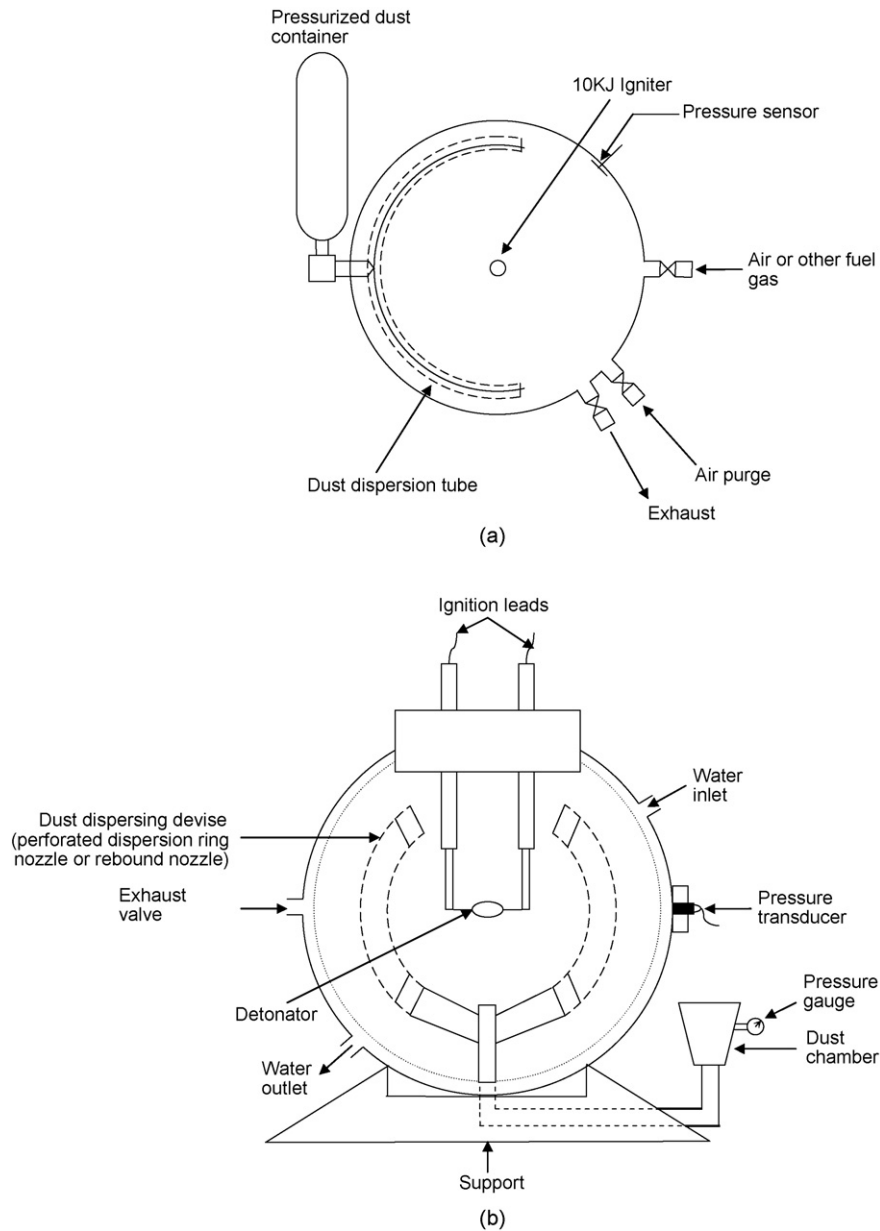


Fig. 3. Spherical vessels for  $P_{\max}$  and  $K_{St}$  testing: (a) ISO 1 m<sup>3</sup> vessel and (b) 201 spherical vessel.

International Electrotechnical Commission (IEC) effort to ‘harmonize’ dust standards with gas standards. But the Directive pays only modest attention to the very different physical and chemical properties of dust clouds and layers, and needs to be suitably revised [6].

The IEC [66] subscope for ‘standardization of uniform practices in areas where combustible dusts are present’ contains the following two points specifying the basic objectives, which are to

- address situations where the presence of dust presents a risk of fire or explosion with respect to use of electrical apparatus;
- test the properties of dusts relating to the risk of fire or explosion.

However, in the revised global scope of IEC’s [67] Ex-standardization work, fires are not included, only explosions. Hence, there is discrepancy between the two scopes; the confusing situation can be resolved by including dust fires even in the global scope [6].

The IEC [68] has produced a standard, the International Protection (IP) code, which defines various ‘degrees of protection’ against ingress of solid objects, including dust particles and water. It is specified by two digits, the first referring to ingress of solid objects, the second to ingress of water. For solid objects six levels of protection are defined, ranging from objects larger than 50 mm (digit 1) to dusts (digits 5 and 6). For water, the corresponding range is from gentle dripping (digit 1) to continuous complete immersion (digit 8). The code also specifies the test methods by which enclosures can be checked for compliance

with the requirements of the various degrees of protection. But the code does not cover protection against ingress of explosive gases.

#### 4.5. Dust explosion triggers

As we have described later in this paper, in theory, the dust explosion hazard can be eliminated by process modifications but, in practice, few industries can maintain their economic viability if such modifications are executed. So the dust explosion hazard is here to stay. What is currently within our means [69] is to reduce the hazard drastically by identifying ‘flash points’ and covering them with more and more layers of protection (LOP).

The first LOP against dust explosion is to identify factors which trigger it and prevent those factors from coming into play. A number of causes can trigger a dust explosion:

- (i) flames and direct heat;
- (ii) hot work;
- (iii) incandescent material;
- (iv) hot surfaces;
- (v) electrostatic sparks;
- (vi) electrical sparks;
- (vii) friction sparks;
- (viii) impact sparks;
- (ix) self-heating;
- (x) static electricity;
- (xi) lightning;
- (xii) shock waves.

These ignition sources differ in terms of temperature, energy and power; the dusts can be ignited by low energy as well as high energy ignition sources. Of these, the ignition sources which can occur inside the plant are of particular importance, and include incandescent material, hot surfaces, sparks, self-heating and static electricity.

A brief description of the ignition sources is presented below.

##### 4.5.1. Flames and direct heat

This obvious trigger can be eliminated by using indirect heating methods like circulating hot water or steam through pipes and using hot water/steam baths.

##### 4.5.2. Self-heating

Self-heating or spontaneous combustion may occur due to exothermic reactions. A wide variety of reactions can give rise to self-heating. These include oxidation reactions as well as reactions of certain dusts with water or wood. In most cases the reaction rate accelerates with temperature, but there are also autocatalytic reactions which may accelerate due to production of a catalyst or removal of an inhibitor. Induction times may be long and the self-heating may be slow to start but may then proceed undetected for a long period. Contaminants such as oil and products of thermal degradation can also contribute to self-heating.

The dust should be screened to determine whether it is prone to self-heating. The dust temperature during the process and in

storage should be controlled. One aspect of this is control of hot surfaces, which may arise in normal operation. Unintended accumulation of dust deposits, which could undergo self-heating, should be avoided.

Situations in which there is a large mass of dust stored at a high initial temperature (to keep the dust dry) are hazardous. Dust in a pile has a high surface area and sufficient air circulation, both of which favor self-heating. The risk of accident is further enhanced during the discharge of hot dust from a drier into a hopper. It may be necessary to cool the dust prior to storage. Another measure which is sometimes used is to recirculate the hot dust through a cooling system prior to its further use.

##### 4.5.3. Hot work

Excessive heat generated during operations such as welding and cutting is another obvious trigger more so when a dust of low ignition threshold (100–200 °C) is present nearby. Accidents often occur because this hazard is not appreciated and the dust is not cleaned out of the equipment before hot work is started.

##### 4.5.4. Incandescent material

Smoldering particles or other incandescent material can trigger a dust explosion inside dust handling equipment. The explosion may then travel through the ducts and connected vessels. Direct firing systems are potential sources of incandescent particles. In direct-fired driers the air inlet should be protected by a fine screen to prevent ingress of such incandescent material.

##### 4.5.5. Hot surfaces

Equipment with a hot surface such as steam pipe or electric lamp, or overheated moving equipment such as distressed bearing, falls under this category of triggers. The surface temperature that can cause ignition of a dust layer is frequently in the range of 100–200 °C. The ignition temperature moves closer to the lower limit of this range as the thickness of the layer increases. Findings from investigations of dust related accidents often reveal that ignitions occur at unexpectedly low temperatures. A dust may contribute to its own ignition; dusts being poor conductors of heat, a layer of dust on the equipment may prevent heat loss to the atmosphere and thus raise the temperature below the surface of the dust heap to the point of ignition. A smoldering or burning layer can act either directly as an ignition source for a dust cloud or by means of agglomerations or ‘nests’ of burning material that break away from deposits and ignite a dust cloud in another part of the plant [46].

When dusts accumulate on hot surfaces, they may go through different and complex stages before combusting. Some dusts burn directly in solid phase with a flame or by smoldering, others melt and burn as liquids. Some dusts can give off large amounts of flammable gases. The size of the flames produced by different dusts also vary.

Hot-surface ignition is a particular problem with driers of various types. According to the IChemE Drier Guide [70,71], the inlet temperatures should be at least 50 °C below the minimum ignition temperature of the dust suspension and 20 °C below that of any dust layer likely to occur. Hot surfaces may also occur as a result of distress in machinery such as pumps and motors. It may

be necessary in some cases to monitor features such as bearing temperatures. A number of incidents have occurred, particularly in silos, involving the use of wander lamps supplied by the mains. These should be avoided and portable battery lamps used instead.

#### 4.5.6. Electrostatic sparks

Electrostatic discharge from electrical equipment may cause a spark which in turn may ignite a dust cloud. Protection against such discharges is based on hazardous area classification and the associated safeguarding. Electrical equipment is designed so that incendive capacity or inductive discharges cannot occur.

#### 4.5.7. Electrical sparks

Electrical sparks occur in the normal operation of switches and relays and in malfunctioning electrical equipment.

To protect against electrical sparks hazardous area ought to be classified and safeguarded. In particular, flameproof equipment must be used, and should exclude dusts. A distinction may be drawn between equipment which is dust tight and excludes dust entirely and equipment which is dustproof and lets in only an insignificant amount of dust [72].

#### 4.5.8. Friction sparks and hot spots

Frictional sparks can occur wherever there is rubbing of one solid with another or during grinding. Foreign materials such as tramp iron can also cause sparks. The dust itself may block the equipment and cause overloading, leading to spark generation. To prevent frictional sparks, dust flow should be controlled and machine overload trips should be installed. Removal of foreign objects should be effected by magnetic or pneumatic separation, especially when the material is to pass through a mill.

Friction-induced heating can also raise dust temperature. Pulling the dust through drag conveyer heats it up a bit, so do mixing operations. But more serious friction-induced heating can occur when hot spots are formed in localized areas of blenders due to the blenders' shearing action.

A thermal runaway reaction which occurred during the mixing of an oxidizer (sodium dichloroisocyanurate), some organic compounds, and inert compounds, in a 500 kg batch was attributed to such hot spots [73]. The accident generated toxic gas release and extensively damaged the blender. The ejected material also caused thermal damage to nearby objects.

#### 4.5.9. Impact sparks

Hand tools may create an incendive impact spark, although there is little evidence from incidents of single impact ignition. The incendive potential of an impact such as that of a metal tool on a metal surface arises from the heating of that surface. The ignition source is not the spark itself but the heated surface, and the heat is transferred from the metal surface to the dust.

#### 4.5.10. Static electricity

Static electricity may turn to sparks when an object moves rapidly into or out of its field. It is more strongly influenced by the process than by the material. For sieving and pouring the charges are low, but for size reduction they are much higher. In certain types of dust handling plants static electricity is readily

generated. These include mills, conveyor belts and pneumatic conveying systems. As with liquids, static charge can accumulate at the center of a large storage hopper. It may then be discharged by an earthed probe. But there is also a hazard unique to dusts that of sliding of highly charged material towards an earthed container wall [12].

The ignition of a dust cloud by static electricity is influenced considerably by particle size distribution, and the duration and the rate of the application of ignition energy [74]. The capacitance of the electrical discharge system may also influence the minimum ignition voltage and the minimum ignition energy.

According to Matsuda [75], 25.7% of dust explosions recorded in Japan between 1952 and 1990 were triggered by static electricity. Nifuku and Enomoto [76] consider agricultural products as being at high risk of dust explosion due to static electricity.

Plastic surfaces such as those used in chutes may give rise to sparks. Bags used for transporting dusts can also create sparks. Sparks generated during the pouring of powder from polyethylene bags into flammable solvents have led to several accidents.

Charge accumulation can occur on non-conductive materials which are being increasingly used in the process industry. When such a charged isolator comes in the neighborhood of a blunt earthed conducting object, it may lead to a 'brush discharge'. In controlled experiments, Larsen et al. [77] were able to observe five instances when such brush discharges ignited dust clouds of sulphur dispersed in oxygen-enriched air.

The human body can generate charge intense enough to make it a potential ignition source. In the manual handling of dusts, the hazard of static electricity from the human body becomes significant if the dust has a minimum ignition energy less than 25 mJ.

Fortunately, the generally high minimum ignition energy of dusts and the tendency of dusts to give corona discharge, contribute towards reducing the risk from static electricity. Modification in the processing conditions of the plant and generating a humid atmosphere can significantly reduce the dust explosion hazard due to static electricity. Earthing may be provided in the form of wire meshes on the walls of storage bins and of earthed rods in the bulk powder. Passive dischargers on the material entering storage bins can also reduce the ignition hazard.

To prevent the human body from generating sparks, conducting antistatic footwear and conductive flooring should be employed. If the minimum ignition energy of the dust being handled is less than 10 mJ it is imperative that antistatic flooring is installed and antistatic clothing and footwear are used.

#### 4.5.11. Lightning, shock waves

Lightening can initiate dust explosions [5,6]. Initiation of dust explosions by shock waves has also been studied [78,79].

In the records of dust explosion incidents the exact trigger is often not mentioned either due to oversight or because it was not known with certainty (Table 2). But accidents of which triggers were clearly identified reveal that welding and cutting, fire, friction, electrical sparks and lightening are among the major causes of dust explosions (Table 3).

Table 3  
Major dust explosion triggers

Ignition source	Proportion found responsible (%)		
	Primary explosions [187]	Elevator incidents [188]	Feed mills [189]
Welding and cutting	10	24.3	12
Fire	7.8	NRA	12
Friction	8.5	NRA	4
Electrical	4.3	6.0	4
Lightning	2.8	1.5	NRA
Static electricity	4.5	1.5	NRA
Unknown	60	25.7	34

NRA: no record available.

#### 4.6. Domino effect

##### 4.6.1. Primary explosions

The dust concentrations adequate for an explosion rarely build-up outside of process vessels, hence most severe dust explosions start within a piece of equipment (such as mills, mixers, screens, dryers, cyclones, hoppers, filters, bucket elevators, silos, aspiration ducts, and pneumatic transit systems). These are called primary explosions even though, in reality, all dust explosions are events which occur after an initiating accident. It is important to note that one of the main differences between the dust explosion and flammable gas hazard is that gas/vapour explosions rarely happen inside vessels due to a lack of air to support explosions. However, the dust is generally suspended in air in process equipment, which can allow dust explosion con-

ditions to occur. This can then cause the vessel to rupture if it has insufficient pressure release devices/venting or if its design pressure is too low.

Even as it is important to attempt eliminating the possibility of primary dust explosions, it is even more important to reduce the possibility of the first explosion setting off a series of other explosions; in other words to prevent 'domino effect'. More so because secondary dust explosions are often more violent than the primary explosions [2,12,80].

##### 4.6.2. Secondary explosions

A primary explosion can disturb settled dust lying nearby, forming a cloud which may then be ignited by the heat released from the primary explosion (Fig. 4). The settled dust occupies very little space, but once disturbed can easily form dangerous clouds. A 1 mm layer of dust of 500 kg/m<sup>3</sup> can give rise to a 5 m deep cloud of 100 g/m<sup>3</sup> dust.

A dust explosion in one part of a powder handling system can cause pressure and/or flames to propagate to other vessels via connecting pipes. For example, in an explosion in a vented bag filter, where typically the reduced explosion pressure is less than 500 mbar, tests have shown that the explosion can propagate into the inlet pipe. This could lead to an explosion propagating with increasing violence throughout the system [43], because the flame propagating in the duct tends to accelerate due to turbulence. It results in a jet flame entering the second vessel. As a result, high combustion rates are obtained at high pressures, even if the second vessel is vented and the amount of dust it contains does not present much danger in itself [11].

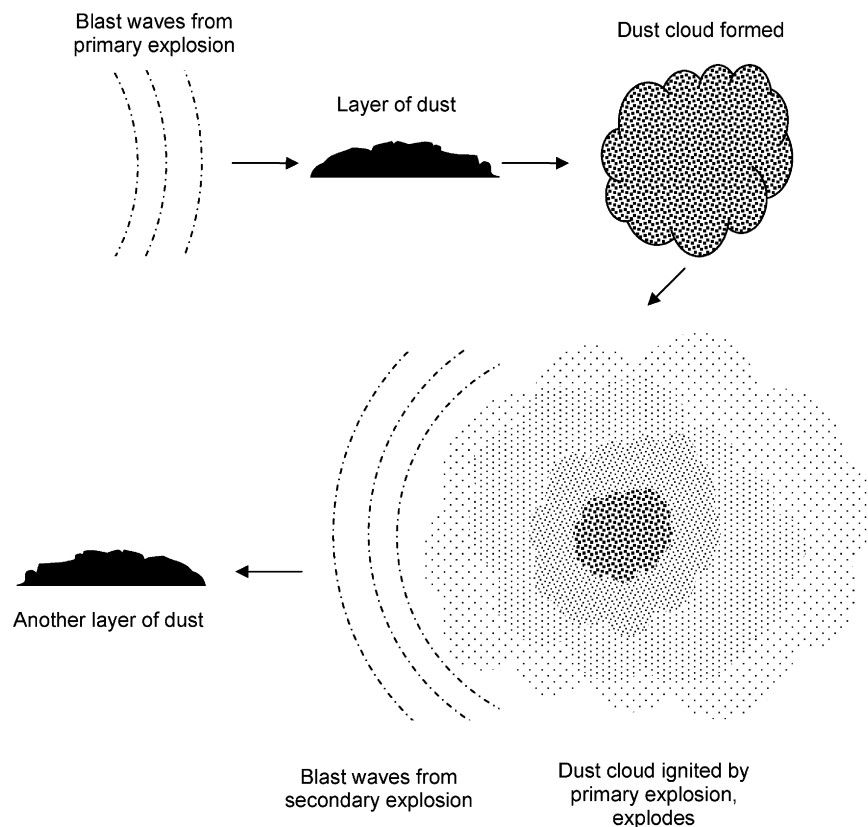


Fig. 4. Domino effect in dust explosions.

If we go by the analogy of the behavior of gas explosions in interconnected enclosed vessels, which has been studied extensively [8,81], ‘pressure-piling’ is likely to occur as secondary and higher order explosions are caused by the primary explosion propagating through interconnected vessels and pipes. The turbulence generated as the first explosion passes through an interconnected pipe into another vessel increases the rate of the dust combustion, hence the rate of pressure rise. As the pressure wave enters the second vessel, it compresses the dust contained in that vessel which then gets ignited by the flame propagated by the first explosion. In this pre-compressed dust cloud, the explosion begins at higher than ambient pressure and the resulting peak pressure of the secondary explosion is correspondingly higher. The energetic nature of the flame jet coming from the first explosion through the interconnecting pipe also contributes to the severity of the second explosion [20].

Lunn et al. [81] produced explosions in coal dust and toner dust in a number of linked systems using vessels ranging in size from 2 to 20 m<sup>3</sup> and connected by 5 cm long pipes with diameters of 0.15 m, 0.25 m, and 0.5 m. They record that the degree of pressure-piling resulting from dust explosions in linked vessels depends on the ratio of the vessel volumes and the diameter of the connecting pipe. Generally an explosion which initiated in a larger vessel and traveled to a smaller one led to very high explosion pressures. Transmission of dust explosions between vessels did not always occur; the narrower the connecting pipe, the lower the probability of a secondary explosion. No transmission of explosion occurred during the experiments of Lunn et al. [81] with a pipe diameter of 0.15 m.

The authors opine that to reduce the hazard of domino effect, volume of the connecting pipe should be added to the volume of the vessel in which initial ignition occurs. The pipe lengths should not be so long that detonation-type explosions can occur in them, nor the pipe volume should be large relative to the vessel volumes.

Of course, as observed by Lunn et al. [81] and others [2,12] the explosive atmosphere in one vessel does not necessarily ignite the dust in the second vessel. It has been found experimentally that the likelihood of ignition breakthrough to the second vessel is a function of the dust properties, the dust concentration, the type of venting of the second vessel, the ducts diameters, the severity of the primary explosions and the presence of any obstructions in the connecting duct [82]. The probability of secondary explosion generally increases with increasing  $K_{St}$  value and the amount of flame produced in the primary explosion [20]. If an obstruction, such as a baffle plate, happens to come in the way of the flame propagating due to the primary explosion, the flame gets distorted, distributing the burning material over a larger area. This increases the size of the potential ignition source.

Proust [83] has developed a software, EFFEX, which models dust flame propagation into a succession of interconnected vessels. At each step of the propagation, the turbulence ahead of the flame is due to the jets coming from the neighboring explosion pressurized rooms. This code implements the flame propagation theory, the turbulence combustion models and the basic combustion parameters reported by earlier authors.

## 5. Frequency of dust explosions

Not all dust explosion events get counted as several, especially the minor ones, are not reported in the media. Hence the historical accounts of dust explosion frequencies are indicative rather than quantitative.

Vijayaraghavan [5] has given an analysis of dust explosions, excluding those in coalmines, in the period 1900–1959. He has listed 1110 explosions and 648 fatalities. Studies by Lunn [84] reveal that during a 10-year span (1958–1967) in the UK, there were 247 reported explosions with 9 fatalities and 324 non-fatal injuries. In the period 1968–1979 there were 474 explosions reported with 25 fatalities and 633 non-fatal injuries; 10 of the 25 fatalities in this latter period occurred in two incidents.

According to the information culled from the UK-based Health & Safety Executive (HSE) by Vijayaraghavan [5], there were 36 dust explosions with injury, and 123 explosions with no injury, during 1979–1988. The principal items of equipment involved were mills, grinders, filters, driers, silos/hoppers and ducts with 51 (17%), 47 (16%), 43 (14%), 19 (6%), and 15 (5%) events, respectively; 95 (31%) events were classified in the category ‘other’. In another survey conducted by the Berufsgenossenschaftliches Institut für Arbeitssicherheit (BIA) 1120 explosions have been identified as having occurred in the United States and Germany (Fig. 5) during the period 1900–1956, of which 536 (48%) have involved industries handling grain, feed and flour [14]. In these 536 explosions, 392 persons were killed, 1015 were injured and the material losses amounted to over \$75 million. The high frequency of explosions in the grain industry occur because grain products can ignite and propagate flames easily, the source of heat required being small.

Eckhoff [85] has examined the details of 75 dust explosions which occurred in the USA from 1900 to 1956, covering a wide range of dusts: wood, food and feed, metal, plastic, coal, paper and chemical. This study, and another by Jeske and Beck [86] which had studied 426 dust explosion occurring between 1965 and 1985, indicated that only about 15% of the dust explosions get recorded. In other words the actual number of dust explosions occurring in the world is over six times higher than the records suggest.

Illustrative example of dust explosions that have occurred in different parts of the world during 1911–2005 have been presented in Table 2. It may be seen that a record is available of only one event from a developing country (China, 1987). This substantiates the fear expressed by us earlier that the dust explosions occurring in the developing world are rarely, if ever, recognized as such but get reported alongside other types of explosions.

Matsuda [75] compiled a report on the dust explosion accidents that occurred in Japan in the period 1952–1990. A total of 248 cases were reported, of which agricultural products were involved in 44 incidents, coal in 13, inorganic materials in 29, metals in 60, chemicals in 32, intermediates and additives in 46 and cellulose materials in 23 instances.

Proust [87] has compiled statistics for dust explosion accidents in five developed countries. According to his estimates, about 160 dust explosions per year occur in Germany. In the USA more than a thousand dust explosion accidents were reported

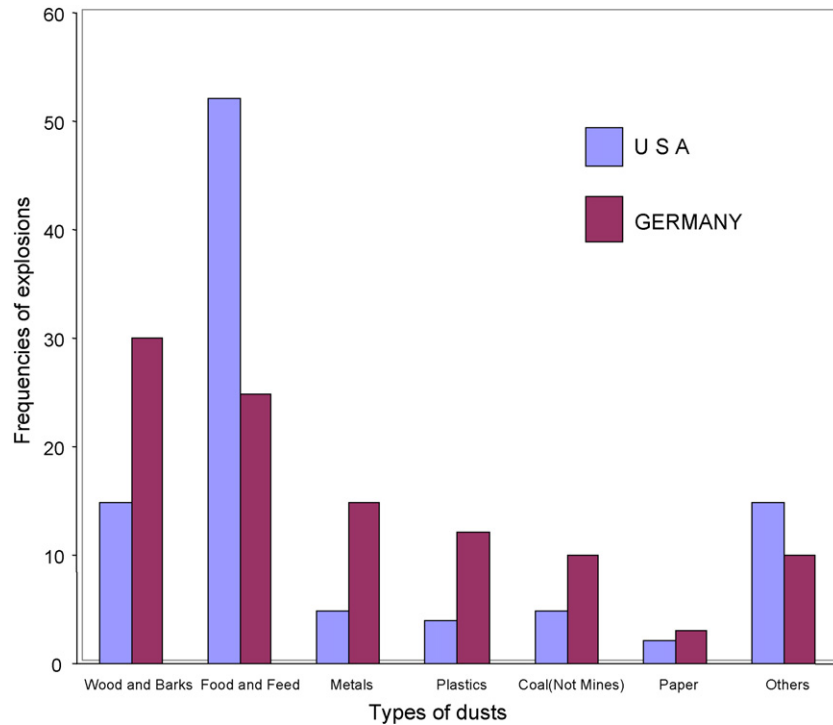


Fig. 5. Frequencies of dust explosion accidents involving different types of dusts (at the USA and Germany).

in the years between 1900 and 1956. Of these, 501 accidents occurred in the food industry during 1958–1978. In Japan, 187 dust explosions occurred in grain storage facilities between 1969 and 1973. In the UK, more than 400 dust explosions were reported during 1969–1976.

A study by Schoeff [27], covering 106 major dust explosions that occurred in agricultural factories in the USA during 1996–2005, indicates that 51 of these occurred in grain elevators and 25 in feed mills. Corn industry accounted for 54% of all accidents in this sector followed by 8.5% in wheat processing facilities.

### 5.1. Historical overview of dust explosion causes

The possible dust explosion triggers have been listed in the previous section. A survey of past dust explosion accidents indicates that of all the triggers of dust explosions, friction, mechanical failure and flames are the ones most often involved.

Abbott [71] and Porter [88] carried out survey of 303 dust explosions that occurred in the UK between 1979 and 1988. Their studies revealed that friction and mechanical failure triggered 18% of the dust explosions surveyed. Flames and flaming material were responsible for 15% of the accidents. Overheating and spontaneous heating were the triggers in 17% of the 303 accidents.

Gummer and Lunn [46] report that a survey on dust explosion triggers carried out by Berufsgenossenschaftliches Institut für Arbeitssicherheit (BIA) identified friction-induced sparks to be the most frequent source of ignition (26%). The BIA reports that, at 11%, smoldering nests were the next most frequent triggers. Mechanical heating caused 9% of the accidents surveyed.

Mechanical sparks have been found to be the most frequent dust explosion ignition source in dust collectors, mills, and grinding plants. Electrostatic discharge caused ignition is responsible for most of the plastic dust explosion accidents in mixers.

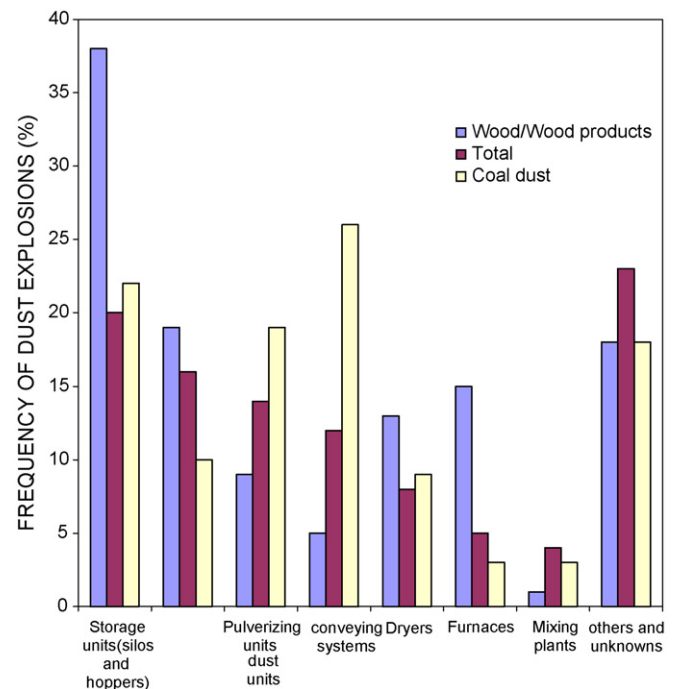


Fig. 6. Frequencies of dust explosions caused in different type of industries in Germany.

Fig. 6 depicts the frequency of explosion in the various processing equipment used in industries processing coal and wood dusts in the Federal Republic of Germany.

Nifuku et al. [89] provide a brief summary of death and injury caused by 269 dust explosions that occurred in Japan between 1952 and 1995 (Table 4). The material-wise inventory and the figures of death or injury per event of these accidents is given in Table 5.

It has been estimated [90] that an average manufacturing facility will have dust explosion once every 20 years, while the probability of such accidents occurring in chemical, pharmaceutical, and milling plants are much higher.

Table 4  
Dust explosion accidents that occurred between 1952 and 1995 in Japan [89]

Year	Explosions	Dead	Injured
1952	6	7	26
1953	9	1	16
1954	9	1	19
1955	4	0	0
1956	7	7	14
1957	4	2	6
1958	8	4	18
1959	7	3	9
1960	6	0	1
1961	3	0	6
1962	8	3	23
1963	11	2	30
1964	7	2	9
1965	12	1	41
1966	6	3	20
1967	8	9	39
1968	12	4	17
1969	8	6	11
1970	6	7	5
1971	7	2	12
1972	7	6	22
1973	12	2	53
1974	7	3	10
1975	9	3	14
1976	4	0	3
1977	6	2	2
1978	8	3	9
1979	9	2	26
1980	3	0	8
1981	2	0	7
1982	3	3	2
1983	3	0	6
1984	3	0	0
1985	5	3	6
1986	8	3	12
1987	1	0	0
1988	3	0	9
1989	1	0	0
1990	6	0	11
1991	4	1	7
1992	4	3	3
1993	1	1	1
1994	6	7	26
1995	6	0	8
Totals	269	106	567

Table 5  
Materials involved in dust explosions and consequent injuries/fatalities [89]

Dust type	Number of accidents	Deaths		Injuries	
		Total	Per accident	Total	Per accident
Cellulosic materials	28	7	0.3	84	3.0
Chemical synthetic materials	36	12	0.3	79	2.2
Coal	13	7	0.5	41	3.2
Food and feed	46	17	0.4	109	2.4
Inorganic	31	9	0.3	28	0.9
Intermediate additives	50	13	0.3	69	1.4
Metals	64	41	0.6	153	2.4
Miscellaneous	1	0	0.0	4	4.0

## 6. Dust explosion mechanism

As explained earlier, when a flammable cloud, formed by the mixing of dust and air in the right proportion and in a confined space, is ignited, a rapid combustion of the fuel takes place, with the propagation of the flame across the cloud. The rate and the extent of flame propagation depend on factors such as nature of dust, dust particle size, and nature of combustion by-products formed. A dust explosion is a complex phenomenon in the sense that it involves simultaneous momentum, energy, and mass transport in a reactive multi-phase system [91].

In order to model the dust explosion phenomenon, assess the dust explosion impacts, and devise prevention and control strategies, a precise understanding of the dust explosion mechanism is required.

### 6.1. The dust combustion process

How similar and how different is the dust combustion process from the combustion of premixed gases? This question has engaged great attention because if the extent of similarity can be established, the dust explosion phenomena can be analyzed using the concepts and tools which exist for studying explosions involving gases.

In the following respects explosive dust clouds behave in a manner similar to explosive gas mixtures [92]:

- (1) flammability/explosibility limits;
- (2) laminar burning velocities and quenching distances;
- (3) the response of the burning velocity to cloud turbulence;
- (4) detonation phenomena;
- (5) adiabatic constant-volume explosion pressures of similar magnitudes;
- (6) well-defined minimum ignition energies;
- (7) minimum ignition temperatures for given experimental conditions.

However, there are two basic differences between dusts and gases which are of substantially greater significance in design of safety standards than these similarities [92]. Firstly, the physics of generation and up-keeping of dust clouds and premixed gas/vapour clouds are substantially different. This means that



in most situations where accidental explosive gas clouds may be produced quite readily, generation of explosive dust clouds would be highly unlikely. Secondly, contrary to premixed gas flame propagation, the propagation of flames in dust/air mixtures is not limited only to the flammable dust concentration range of dynamic clouds. The state of stagnant layers/deposits offers an additional discrete possibility of flame propagation. In the context of these observations, Eckhoff [92] has opined that a revision of the existing European Directives 94/9/EC [93] and 1999/92/EC [94] to clarify important basic differences between dusts and gases/vapours is urgently needed.

In dust clouds inertial forces can produce fuel concentration gradients (displacement of particles in relation to gas phase). Furthermore, thermal radiation may contribute significantly to the heat transfer from the flame to the unburnt cloud, depending on the type of particle material (e.g. light metals).

It has often been assumed that the laminar burning velocity of a given dust cloud is a basic combustion property of the cloud, which is closely related also to the burning velocities at various levels of turbulence, and hence to the flame propagation through that type of cloud at large [95]. A burner apparatus was used by Dahoe [96] and Dahoe et al. [58] to produce stable cornstarch flames in air, and the laminar burning velocity was measured via laser Doppler anemometry (LDA). It was found that the laminar burning velocity varied with flame shape, and this was accounted for by introducing the ‘Markstein length’ of a dust/air flame. This parameter is specific for any given dust cloud. It has a magnitude of the order of the laminar flame thickness of that specific dust cloud, and serves as a measure of the sensitivity of the laminar burning velocity to changes in the flame shape.

When considering turbulent flame propagation in dust clouds, flame distortion and turbulence produced by the propagating explosion itself is central for understanding both dust and gas explosions in practice. Rzal and Veyssiere [97] report possible differences between turbulent combustion of premixed gases and dust clouds. They investigated the interaction of a laminar maize starch/air flame with an obstacle, viz. a sphere, a disk or an annulus. With the annulus, flame quenching phenomena were observed, which were attributed to centrifugal separation of dust particles and air in the turbulent eddies. This is a very important observation, indicating that the burning rate of a dust cloud may not necessarily respond to turbulence in the same way as the burning rate of a premixed gas. Significant differences also exist on the microscopic scale. For example, the basic microscopic turbulence mechanisms that promote the combustion process must be identified. The results of Mitgau [98] and Mitgau et al. [99] indicate that more efficient replacement of gaseous reaction products by fresh air round each particle may be a strong basic turbulent combustion enhancement mechanism.

## 6.2. Role of dust particle size in the combustion process

Flame propagation across a dust cloud occurs in two ways:

- (i) By the combustion of flammable gases emitted by particles heated to the point of vapourization or pyrolysis [100,101].
- (ii) Through direct oxidation at the dust particle surface [102].

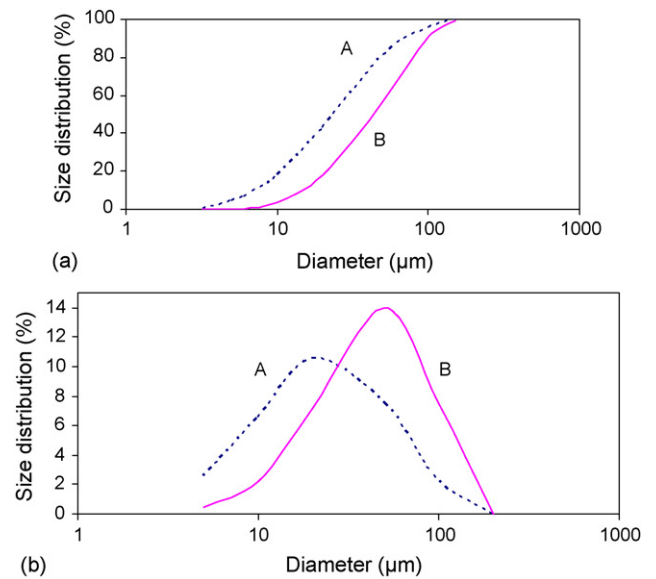


Fig. 7. (a) Cumulative dust particle size distributions. (b) Differential distribution of dust particle size: (A) surface area weighted and (B) mass or volume weighted.

In either case, the particle size plays an important role in the combustion process.

Larger dust particles participate inefficiently in the flame propagation process, whereas finer dust particles of the same material are likely to react faster and more efficiently during combustion [7,103]. This is so because the finer particles have greater surface area per mass, are more easily dispersed in air and remain airborne for longer periods. Fig. 7 shows the cumulative and differential particle size distribution of a typical dust.

Such differential curves are useful in visualizing the dust particle size distribution.

The dust particle size can be calculated from the surface mean diameter ( $D_s < 30 \mu\text{m}$ ) and the mass mean diameter ( $D_w < 50 \mu\text{m}$ ) provided by Fig. 7. But, since it is the surface area of the dust particles that plays a major role in the combustion process, the mean particle diameter which reflects the surface area is a more appropriate characteristic than the one based on mass. The minimum explosible concentration (MEC) are size-independent for very fine powders but begin to increase for particles above  $30 \mu\text{m}$  until a size is reached that cannot be easily ignited [7].

Experimental evidence gathered over the last two decades support the idea that the basic flame propagation mechanism in dust clouds has a *general* similarity with premixed gaseous flames [8] even though differences *also* exist because a dust particle must first volatilize before catching fire. The microscopic turbulence mechanism is also different for dust suspensions in comparison to gaseous mixtures [95]. For particles able to gasify at low temperature (below  $1000^\circ\text{C}$  for instance), flame propagation results from a heat balance between upstream thermal conduction, inducing intense gasification of the particles, and heat release in gaseous phase. Most industrial dusts fall into that category. But, for other powders which have high gasifica-

tion temperature like aluminum, particles present in the burnt products may radiate strongly towards the reactant in such a way that the heat balance will be between thermal radiation and heat release by some heterogeneous combustion process. The interaction between turbulence and combustion in dust clouds is possibly of a lower degree than in gaseous flames, but still roughly of the same nature, with similar correlations linking the turbulent burning velocity to the parameters of the turbulence and of the laminar flames.

Experiments by Chatrathi and Going [104] show that a flammability curve can be created for fuel/inert dust mixtures. This flammability curve for dusts has the same characteristics as a flammability curve for gas mixtures, and is characterized by a lower flammable limit, an upper flammable limit, and a minimum inerting concentration.

It must again be emphasized that the processes leading up to the point of ignition of dusts are different from the ones associated with the ignition of flammable gases. An example is organic dusts (most common in industry) where the actual combustion only starts after a slow devolatilization process. Pilao et al. [105,106] studying cork dust, report that the mechanism of flame propagation in that dust followed the known pattern for organic dusts in which gas phase combustion was preceded by devolatilization of dust particles. This is one of the reasons why the ignition delay for such dusts may be of the order of tens of milliseconds. During this time, a pressure wave would have covered a distance of more than 1 m [11]. The flame consumes the

combustible mixture with a speed corresponding to a burning velocity, which is determined empirically, depending on kind of mixture, its concentration, turbulence and temperature.

### 6.3. Factors influencing dust explosibility

A number of factors influence dust explosibility; in order to assess the dust explosion hazard posed by a substance, it is necessary to quantify the factors that influence it. They are:

- (i) particle size;
- (ii) dust concentration;
- (iii) oxidant concentration;
- (iv) ignition temperature;
- (v) turbulence of the dust cloud;
- (vi) maximum rate of pressure rise;
- (vii) admixed inert dust concentration;
- (viii) presence of flammable gases.

A ready reckoner of the forces and counter-forces that influence dust explosibility is presented as Table 6.

#### 6.3.1. Particle size

As explained in Section 6.2, larger the surface area per unit mass of a dust particle, greater is the hazard it poses. However, in some cases, there is a likelihood of very small particles agglomerating into lumps. If this happens, the explosibility of the dust

Table 6  
Influence of dust properties/characteristics on dust explosion parameters [12]

Parameter	Increases with	Decreases with
Explosibility of the dust	<ol style="list-style-type: none"> <li>1. Lower explosible concentration</li> <li>2. Minimum ignition temperature</li> <li>3. Lower minimum ignition energy</li> <li>4. Burning velocity</li> <li>5. Maximum rate of pressure increase</li> <li>6. Presence of chemical groups such as COOH, OH, NH<sub>2</sub>, NO<sub>2</sub>, C≡N, C=N, N=N</li> <li>7. Presence of volatile matter in the dust at levels above 10%</li> <li>8. Relatively small proportion of fines</li> <li>9. Increasing oxygen concentration</li> </ol>	<ol style="list-style-type: none"> <li>1. Presence of chemical groups such as Cl, Br, F</li> <li>2. Presence of inert material at concentrations above 10–20%</li> <li>3. Dust moisture content above 30%</li> </ol>
Effect of particle size on the likelihood of explosion initiation	50–70 μm < particles size (μm) < 500 μm	500 μm < particles size (μm) < 50–70 μm
Minimum explosive concentration	<ol style="list-style-type: none"> <li>1. Increasing moisture content</li> <li>2. Increasing admixed inert dust concentration</li> </ol>	<ol style="list-style-type: none"> <li>1. Decreasing particle size</li> <li>2. Increasing volatile matter</li> <li>3. Increasing oxygen concentration</li> </ol>
Minimum ignition temperature	<ol style="list-style-type: none"> <li>1. Increasing moisture content</li> <li>2. Increasing admixed inert dust concentration</li> </ol>	<ol style="list-style-type: none"> <li>1. Decreasing particle size</li> <li>2. Increasing volatile matter content</li> <li>3. Increasing oxygen concentration</li> <li>4. Increasing thickness of the dust layer</li> </ol>
Maximum permissible oxygen concentration	Decreasing dust temperature	Increasing dust temperature
Maximum explosion pressure	Decreasing particle size, though weakly	
Maximum rate of pressure rise	<ol style="list-style-type: none"> <li>1. Decreasing particle size</li> <li>2. Increasing volatile matter content</li> <li>3. Increasing oxygen concentration</li> </ol>	<ol style="list-style-type: none"> <li>1. Increasing moisture content</li> <li>2. Increasing concentrations of admixed inert dust</li> </ol>

decreases and when the particle size increases beyond 500  $\mu\text{m}$ , it may even become non-ignitable.

The explosibility of dusts does not vary linearly with particle surface area although it depends on it. This dependence is dictated by the actual speed of combustion of volatiles and the concentration of dusts.

### 6.3.2. Dust concentration

A dust cloud would explode only if the dust concentration is within certain limits. These generally are:

- 50–100  $\text{g}/\text{m}^3$ : lowest concentration;
- 2–3  $\text{kg}/\text{m}^3$ : maximum concentration.

According to Hertzberg et al. [100] a certain stoichiometric concentration of volatiles in air of the solid phase fuel must be generated for a flame to propagate rapidly through the mixture before more fuel volatiles are produced. This indicates that the lower concentration limit is determined by the minimum quantity of fuel particles that must exist in order to sustain combustion. A near parabolic relationship exists between dust concentration and ignition energy. The latter is high at high dust loadings and decreases to a minimum value with decreasing concentrations. A further decrease in dust concentrations result in an increase in the ignition energy [64].

The upper concentration limits are dictated by the minimum amount of oxygen needed for explosion.

As is the case with flammability characteristics, data on the explosibility of the same dust differs from test to test [12,107]. The considerable progress made in dust explosibility testing (summarized in Section 4.3) has reduced this variability, but still significant differences exist and utmost care must be exercised in developing control strategies on its basis.

### 6.3.3. Oxidant concentration

One of the sides of the ‘dust explosion pentagon’ is the oxidant, which usually is oxygen in air. Oxygen influences the dust explosion process to a very large extent. Oxygen concentration greater than 21% tends to increase the burning velocity of the fuel. But for concentration less than 21% the burning velocity is reduced. This happens because oxygen is consumed by the fuel in the combustion process, thereby decreasing the oxygen concentration. As a consequence of this, the rate of combustion of the dust comes down. Eventually, the combustion may die down, or, if an explosion occurs, it may be less severe. Fire is sustained only if the oxygen concentration in air is greater than 10%.

### 6.3.4. Ignition temperature

If a mixture of ignitable dust and air gets heated, it would catch fire at some point. The lowest temperature at which such ignition occurs is characterized as ‘minimum ignition temperature’ (MIT). As summarized in Table 6, MIT increases with the presence of moisture or other inertants in dust cloud but decreases with decreasing particle size and increasing volatile matter, oxygen concentration, and thickness of dust layer.

Measurement of MIT is essential for eliminating ignition sources and for designing explosion suppression systems.

When dust-oxidant clouds are subjected to temperatures higher than MIT, the lower flammability limit and the minimum explosible concentration of the clouds expectedly decrease. But the maximum absolute explosion pressure,  $P_{\text{max}}$ , also decreases [56,108]. For coal dusts, Cashdollar [56] has explained this effect to be occurring because at elevated temperatures fewer oxidant molecules are available to react with the coal. The author measured the MEC at 60 °C and 80 °C using a 201 chamber test apparatus. Applying the modified Burgers–Wheeler law for hydrocarbons, he has extrapolated the two data points to obtain a curve correlating temperature and MEC.

Typical ignition temperatures of common dusts in air of relative humidity 30–90% are:

- wheat flour: 410–430 °C;
- corn starch: 410–450 °C;
- rye dust: 430–500 °C.

Indeed most grain and flour dusts can be ignited if they are blown against hot surfaces bearing temperatures of the order of 400–500 °C. Even at temperatures of  $\sim 200$  °C dusts can be ignited under favorable conditions (low relative humidity, small particle size, high turbulence) and begin to smolder. The smoldering nests can then cause explosion [46].

### 6.3.5. Role of turbulence

The rapid, more or less random, movement of small elements constituting the dust cloud in three-dimensional space generates turbulence. A highly turbulent cloud will have evenly distributed dust in it. When such a cloud catches fire, the turbulence will cause a mill-like effect: mixing the hot burnt/burning parts of the cloud with the cold unburnt parts, generating a three-dimensional laminate of alternating hot burnt/burning and cold unburnt zones. In short, a flame will propagate very quickly through a dust cloud if the latter has high degree of turbulence, resulting in a violent explosion. Turbulence affects the rate of pressure rise much more than the peak pressure [109].

On the other hand, when ignited, a less turbulent cloud releases an initial large amount of heat which is locally concentrated due to its low rate of heat dissipation. Further propagation of any flame produced in the dust cloud is due entirely to the degree of dust dispersion. A more evenly dispersed dust burns more easily.

There are two kinds of turbulence, differing in their origin, that are generated by industrial process involving dusts. The first one is generated by the dust production operations such as by air jet mill, mixer, bag filter, pneumatic transport pipe and bucket elevator. This type of turbulence is often called initial turbulence.

The second kind of turbulence is generated during the combustion process after the dust cloud has ignited. It is an expansion-induced flow of unburnt dust cloud ahead of the propagating flame. The speed of the flow and the geometric constrictions present at the operation site govern the degree of turbulence generated. For example, vent openings and obstacles like buckets in a bucket elevator, enhance the turbulence gen-

eration process. By and large, the turbulence generated by the flame front is much greater than the initial turbulence [110].

Given that the rate of combustion and other chemical reactions associated with dust explosions are characterized through a set of fundamental properties such as burning velocity, turbulence has been regarded by some authors [111] as ‘the single most important factor whose effects need to be incorporated in a model of dust explosions’.

### 6.3.6. Maximum rate of pressure rise

The rate of pressure rise, when a dust is ignited, is not only a measure of the ‘explosibility’ of a dust but is also a key property on which the design of several explosion detection systems and vents are based.

According to the classical combustion theory [112], for the ideal case, the absolute pressure as a function of time,  $P(t)$ , in a constant volume, spherical explosion, is related to the fractional volume,  $V(t)$ , occupied by the fireball during the time of propagation,  $t$ , as follows [113]:

$$\frac{P(t) - P_0}{P_{\max} - P_0} = k \frac{V(t)}{V_0} \quad (1)$$

where  $P_0$  is the initial absolute pressure,  $V_0$  the chamber volume, and  $k$  is a correction factor related to the difference in compressibility between burned and unburned gases. For spherical propagation from a point source:

$$\frac{V(t)}{V_0} = \left[ \frac{r(t)}{r_0} \right]^3 = \left[ \frac{S_b t}{r_0} \right]^3 \quad (2)$$

where  $r(t)$  is the fireball radius,  $r_0$  the chamber radius, and  $S_b$  is the flame speed given by

$$S_b = \frac{dr(t)}{dt} = \left( \frac{\rho_u}{\rho_b} \right) S_u \quad (3)$$

where  $\rho_u/\rho_b$  is the density ratio of unburned to burned gases (at constant pressure). The burning velocity,  $S_u$ , is the rate of flame propagation relative to the unburned gas ahead of it. The flame speed,  $S_b$ , is relative to a fixed reference point. Note that both  $S_b$  and  $S_u$  are for turbulent non-laminar conditions for dust explosions. For spherical propagation in a spherical chamber, the maximum pressure is reached just as the flame contacts the wall. At that instant,  $k = 1$ . Differentiating Eq. (1) with respect to time and substituting Eqs. (2) and (3) yields:

$$\frac{dP(t)}{dt} = 3(P_{\max} - P_0) \left( \frac{\rho_u}{\rho_b} \right) S_u \frac{r(t)^2}{r_0^3} \quad (4)$$

Setting

$$r(t) = r_0 = \left( \frac{3V_0}{4\pi} \right)^{1/3}$$

and letting

$$\frac{\rho_u}{\rho_b} \approx \frac{T_b}{T_0} \approx \frac{P_{\max}}{P_0},$$

we get

$$K_{St} = \left[ \frac{dP(t)}{dt} \right]_{\max} V_0^{1/3} = 4.84 \left( \frac{P_{\max}}{P_0} - 1 \right) P_{\max} S_u \quad (5)$$

$K_{St}$  is the size normalized maximum rate of pressure rise. As stated earlier, the subscript ‘‘St’’ refers to staub, the German word for dust.

Bartknecht [114] and Wiemann [108] who studied the effect of initial pressure on the  $P_{\max}$  and  $K_{St}$  values found that  $P_{\max}$  increases linearly with increase in initial pressure, over the range of 1–4 bar. They also found that  $K_{St}$  increases with initial pressure.

Even though the  $K_{St}$  concept has its limitations, as elaborated in Section 4.2, it nevertheless remains a key parameter for the design of explosion vents [115]. It has also been used in expert systems developed for assisting in vent design [116].

### 6.3.7. Admixed inert dust concentration

Experiments by Chatrathi and Going [104] on the suppressing effect of sodium bicarbonate, potassium bicarbonate, mono-ammonium phosphate, and calcium carbonate on the dusts of coal, cornstarch, polyethylene, anthraquinone, etc. showed that flammability curves can be created for fuel/inert dust mixtures. The flammability curves for dusts had the same characteristics as the flammability curves for gas mixtures, and were characterized by a lower flammable limit, an upper flammable limit and a minimum inerting concentration. The suppression results showed that high  $K_{St}$  deflagrations and metal dust deflagrations can be extinguished and the maximum explosion pressure can be reduced to an acceptable level. The effectiveness of the extinguishing agent used was dependent upon the compatibility of the fuel dust with the inert dust. Specific heat, thermal conductivity, absorptivity, particle geometry and particle decomposition seemed to play a critical role in the effectiveness of suppressants. These factors along with flame temperature and heat of combustion may explain the variation in explosion severity and extinguishment effectiveness.

### 6.3.8. Presence of flammable gases

If a flammable gas is also present in the midst of the dust, the explosibility of the latter is enhanced. The minimum explosive concentration, minimum ignition temperature, and minimum ignition energy are all reduced, and the increase in maximum rate of pressure rise goes up. Thus a flammable gas can render explosive a dust–gas mixture at a dust concentration which is below the normal lower explosive limit for the dust and at a gas concentration below the normal lower explosive limit for the gas [117]. It may also make explosive a dust of such large particle size which would otherwise have been non-explosive. For example dust of class St 0 changes to classes St 1, 1/2, 2 and 3 at methane concentrations of 1%, 3%, 5% and 7%, respectively, and to classes 1, 2/3 and 3 at propane concentrations of 0.9%, 2.7% and 4.5%, respectively [2,12]. For hybrid air/methane-cork dust mixtures Pilao et al. [105] observed that the risk of explosion rises with the increase of methane concentration. Hybrid mixtures of polyurethane–cyclopentane and

plastic dust–cyclopentane were two times more sensitive to dust explosion than the dust without the cyclopentane gas [40]. For mixture of hydrocarbon dusts and gases, the mixing law of Le Chatelier is applicable [7,118,119].

The minimum ignition energy of the dust–flammable gas mixtures is also lower than that of the dust alone.

## 7. Prevention of dust explosion

### 7.1. Dust explosion prevention strategies

The right conditions that must prevail for a dust explosion to occur is summed up under the ‘dust explosion pentagon’ (Section 3). The most obvious way to prevent a dust explosion from happening is to not allow the dust pentagon to be closed. This can be attempted in the following ways:

- (i) Effectively modifying the process to reduce dust handling hazards.
- (ii) Preventing suspensions of flammable dusts.
- (iii) Completely removing or minimizing the presence of ignition sources.
- (iv) Inerting.

#### 7.1.1. Process modification

The most obvious and fool-proof way to prevent dust explosions is to replace existing processes with the ones which do not deal with combustible dusts. But, sadly, it also happens to be a strategy confirming to the axiom ‘easier said than done’ because considerations of process economics, not to speak of viability, may drastically reduce the options of process modifications available to existing industries.

Inherently safe process design to prevent or reduce dust explosion hazard involve use of such production, treatment, transportation and storage operations where dust cloud generation is kept at a minimum. One example is use of mass flow silos and hoppers instead of the frequently used funnel flow types. Eckhoff [120] has emphasized the importance of knowing powder science and technology when striving for inherently safe process design in industries having a dust explosion hazard. Amyotte and Khan [121] have proposed a framework for directing the concept of inherently safe process design specifically towards reducing the dust explosion hazard in industry.

Amyotte et al. [122] have identified the following four elements associated with inherently safe design which may reduce the risk of accidents:

- (1) Minimize (intensification): use smaller quantities of hazardous materials when the use of such materials cannot be avoided.
- (2) Substitute (substitution): replace a hazardous substance with one that is less hazardous or a hazardous process route with one that does not involve hazardous material.
- (3) Moderate (attenuation/limitation of effects): use hazardous material in their least hazardous forms or identify options that involve less severe operating conditions.

- (4) Simplify (simplifications/error tolerance): design processes and equipment to eliminate opportunities for errors by identifying ways to eliminate excessive use of add-on safety features and protective devices.

Amyotte et al. [122] present the following recommendations:

- Use nitrogen as a conveying gas instead of air.
- Use nitrogen sealing in silos.
- Fill silos using a cyclone to reduce dust cloud dispersion.
- Carefully control the particle size.
- Reduce electrostatic problems with silos and bag filters by checking the relative potential of metal construction parts.
- Control moisture in pipes and silos.
- Use lower mass flow rates.
- Use online monitoring of the electric field of compacted powders in silos.
- Use conduction of the electric field, if required.
- Design silos and explosion isolation valves between silos for explosion venting, so that sequential dust explosion will be avoided.
- Keep the dust concentration below the minimum explosible concentration.
- Design and test explosion blocks in conveying pipes.

#### 7.1.2. Preventing flammable dust suspensions

It is difficult to keep the flammable dust cloud concentrations below certain levels in order to prevent an explosion, because the minimum explosive concentration is usually far below the economic operational conditions [41].

The following measures may be effective:

- (i) In cases where high dust concentration may be unavoidable, it would be appropriate to work with smaller piles of dust than with one large one.
- (ii) Situations such as the free fall of dust from a height into a hopper, which may encourage dust cloud formation, should be avoided.
- (iii) The dust removal process, say from a gas stream, must be done at as early a stage as process considerations permit in order to avoid dust suspensions.
- (iv) Plants handling flammable dusts should be appropriately designed to minimize the accumulation of dusts. Cleaning of dusts collected in places like ducts should be facilitated as often as permissible.

It must be emphasized that even if a dust suspension within the explosive range is not present during normal operations, it may be so during startup, shutdown or fault conditions. It may not cause what we normally call ‘primary explosion’ yet it may trigger a ‘secondary explosion’ by disturbing the nearby dust heaps.

By adhering to certain safe housekeeping practices, the presence of dust can be limited to controlled locations thereby reducing the potential for the formation of hazardous dust clouds. Once a dust explosion is initiated, the expanding gases

behind the flame of such incipient dust explosion can whirl-up the otherwise settled dust lying nearby, thus feeding the explosion. NFPA 654 [123] provides guidance on housekeeping practices which maintain that: (a) dust layers 1/32 in. (0.8 mm) thick can be sufficient to warrant immediate cleaning of the area; (b) a dust layer this thick can create a hazardous condition if it covers more than 5% of the building floor area, with 1000 ft<sup>2</sup> (93 m<sup>2</sup>) of dust layer as the upper limit for large facilities; (c) dust accumulations on other surfaces, such as overhead beams and joists, ductwork, conduit and cabling, piping, light fixtures, or tops of equipment, can also contribute significantly to the secondary dust cloud potential, and should be considered in estimating the dust loading in a room; (d) dust adhering to walls and other vertical surfaces should also be considered.

These are non-mandatory recommendations but NFPA 654 does stipulate the dust layer thickness of 0.8 mm as a basis for defining dust explosion hazardous areas in the main (mandatory) portion of its standard. Considering that the measure of 0.8 mm represents extreme thinness, NFPA 654 has strongly emphasized the necessity of reducing dust accumulation.

Physical barriers to limit dust migration have been permitted in NFPA 654 to minimize the extent of the housekeeping zone, but, expectedly it has also stipulated that all penetrations of floors, walls, ceilings, and partitions defining such barriers be dust tight. The standard also stipulates that all surfaces where dust might accumulate be designed and constructed to minimize dust accumulations and to facilitate cleaning (for example, interior window ledges can be sloped, beams can be boxed in, and concrete walls can be painted to limit dust adherence). The standard also requires sealing of spaces that may be inaccessible for cleaning and the installation of localized dust collection systems to limit dust migration. Such systems, however, must be carefully designed, operated, and maintained to control their own inherent dust explosion hazards. For example a small explosion inside a filter or a pressure wave arriving at the filter bag from the connecting pipe work can rupture the filter, blow the contents of the filter into open space and cause a violent secondary explosion.

One of the most effective ways of limiting the spread of dust through a facility is to keep it inside the equipment. Proper design, maintenance, and operation of equipment to minimize dust emissions is, therefore, of prime importance.

Unsafe housekeeping such as vigorous sweeping or the use of steam or compressed air to blow down equipment in dusty areas may lead to the formation of combustible dust clouds. NFPA 654 permits the use of steam or compressed air only when (a) the area and equipment have been vacuumed before blowdown; (b) electrical power and other sources of ignition have been shut down or removed; (c) the steam or air pressure is limited to 15 psig; (d) there are no hot surfaces in the area capable of igniting a dust cloud or layer.

If vacuuming is intended as part of the housekeeping program, NFPA requires either the use of a fixed-pipe (house) system with a remotely located exhauster and dust collector (properly protected against explosions), or a portable vacuum cleaner listed for use in Class II hazardous locations. Other commodity-specific standards generally parallel the require-

ments in NFPA 654. More restrictive requirements, however, may exist for certain commodities, for example, NFPA 484 (combustible metals, metal powders, and metal dusts) addresses the ease with which aluminum dust can be ignited and requires that: (a) the “Preliminary cleanup... shall be accomplished by using conductive, non-sparking scoops and soft brooms, as well as brushes that have natural fiber bristles”; (b) very stringent controls on the design and use of vacuum cleaners; (c) restrictions on the use of water due to the reaction of water and aluminum to produce flammable hydrogen gas.

Based on an analysis of six past accidents which have occurred between 1995 and 2003, Frank [25] has identified the following housekeeping imperatives:

- (1) Specific characteristics of the dust involved (such as MIE, conductivity, chemical incompatibilities) must be considered in planning safe housekeeping procedures.
- (2) Housekeeping programs must comprehensively address all areas where combustible dust may accumulate.
- (3) One must strive to limit the production of dust clouds during housekeeping, after first de-energizing or removing all ignition sources.
- (4) Regular cleaning frequencies must be established (an NFPA 654 requirement) and complied with; cleaning may be done if dust accumulations necessitate.
- (5) Extreme caution must be exercised when using compressed air for cleaning.
- (6) One must learn from near misses, and treat them as fore-warnings.

### 7.1.3. Elimination of ignition sources

A brief description of the various types of dust triggers was given in Section 4.4. In situations where the minimum electrical spark ignition energy of the working dust is considerably greater than 10 mJ, elimination of ignition sources would provide adequate protection against dust explosions.

The ignition sources, which are traceable to routine operations or worker habits such as smoking, open flames, open light (bulbs), welding, cutting, and grinding, can be eliminated by sufficient staff training and enforcement of discipline.

The ignition sources that originate in the process itself involve factors such as open flames, hot surfaces, self-heating, smouldering nests and exothermic decomposition, heat from mechanical impacts, exothermic decomposition of dust via mechanical impacts, and electric sparks and electrostatic discharges. As these ignition conditions are inherent in the actual process, the hazard can be reduced by employing the right precautionary measures like regular cleaning of accumulated dust at the process site, earthing of equipment that may develop charges, inspection of odd noises; and strict adherence to the process operation norms.

Which ignition sources have greater probability of initiating combustion and which have lesser?

Hesby [124] reports that the number of sparks from single accidental impacts of steel objects is too low to be able to cause ignition of the layers of organic dusts studied. According to

the studies of Gummer and Lunn [46], smoldering nests are poor ignition sources for most dust clouds, whereas, flaming nests cause ignition more readily. Further work is needed to clarify both the conditions under which smoldering or flaming nests of various materials are generated in industrial plant, and the circumstances under which such nests will ignite explosive clouds of various dusts [95].

The minimum hot-surface temperature for ignition of a dust cloud varies with scale as well as the geometry of the hot surface in relation to the dust cloud. Consequently, results from small-scale laboratory tests ought not to be directly applied in design of large-scale industrial plants. Development of numerical models for dynamic simulation of hot-surface ignition processes would be helpful in this regard [95].

The parameters influencing the minimum energy required for igniting a dust cloud by an electric spark include voltage and current characteristics across the spark gap, spark gap geometry and electrode material, as well as all the dust cloud parameters. The latter include particle material and particle size/shape distributions, dust moisture content, dust concentration, and the dynamic state of the dust cloud with respect to the spark gap. Minimum ignition energies (MIE) of clouds of a given dust material decreases strongly with the fineness of the dust [95].

Randeberg and Eckhoff [125] have investigated an alternative method for measuring MIEs of explosive dust clouds, which may approximate accidental electrostatic spark ignition in industrial plant better than other methods. In the conventional method a special electronic system is employed for optimal synchronization of the dust cloud and the spark discharge. Randeberg and Eckhoff [125] have used the transient dust cloud itself to initiate spark breakdown between a pair of electrodes pre-set at a high voltage somewhat below the breakdown voltage in dust-free air. Using this method, the MIEs of three dusts were determined. The results were of the same order, although somewhat higher than those obtained using the conventional method.

The issue of whether one-electrode discharges such as brush discharge can ignite dust clouds was examined by Larsen et al. [77]. They were able to ignite clouds of sulphur dust in oxygen-enriched air by true brush discharges. However, ignition in air only was never observed. Because of the very low MIE of clouds of sulphur dust in air, this indicates that ignition of even the most sensitive dust clouds by brush discharges in air is unlikely.

#### 7.1.4. Inerting

‘Inerting’ refers to ways and means by which the oxygen concentration in a process area or a vessel is reduced by adding an inert gas to a level at which the dust cloud can no longer propagate a self-sustaining flame. Such inerting would slow down or totally prevent the dust explosion pentagon (Section 3) from taking shape, thereby reducing the explosion hazard. ‘Inerting’ is also practiced, though much less frequently, by mixing a combustible dust with a non-combustible one.

**7.1.4.1. Use of inert gases.** The gases commonly used for inerting of hazardous dusts are nitrogen, carbon dioxide, water vapour and rare gases. Selecting a suitable gas depends on various factors the principle one being the reactivity (or rather the

lack of it) of a gas with the dust for which it is used. For example, CO<sub>2</sub>, which is otherwise a useful inerting gas for several dusts, cannot be used with aluminum dust as it reacts violently with it. At high temperatures, nitrogen reacts strongly with magnesium dust and hence cannot be used in process involving the latter. Other factors are the availability and cost of supply of the relevant gas. In situations where nitrogen or carbon dioxide is incompatible with some powders, it is advisable to use rare gases. Applying water spray or increasing the relative humidity in the work area are the practical ways of inerting the dusts during open operations such as shredding [41].

Inerting is done in the following way. The system is slightly evacuated and then flushed with the inert gas until the original pressure is regained. This is repeated until the desired level of inerting is accomplished. If a high pressure system is being used, the inert gas may simply be pumped into the process vessels until the desired pressure is reached. Once inerting has been done, care must be taken that no air leaks into the process. If a new gas is introduced with the feed, it should also be inerted.

Often partial inerting is used where total inerting may be too costly; this does not eliminate the chance of explosion, but limits it substantively.

To accomplish partial inerting the gas (most often air) in which the explosible dust is dispersed is mixed with a fraction of inert gas (e.g. nitrogen) considerably smaller than that required for complete inerting. This reduces both the explosibility and the ignition sensitivity of the dust cloud. The effects on  $K_{St}$  (explosion violence) and MIE (minimum ignition energy) are particularly pronounced. This can facilitate the implementation of conventional protective methods that would otherwise have been difficult to use. By using published data for coal dust and the new European CEN standard for vent sizing, Eckhoff [47] has shown that the minimum required areas for explosion venting are reduced considerably, due to reduced  $K_{St}$  and  $P_{max}$  values, by even a moderate reduction in the content of oxygen in the atmosphere. The author has also shown, using a qualitative probabilistic argument, how the marked increase of MIE obtained by partial inerting would justify a further reduction of minimum required vent areas.

In many cases the explosion hazard may be reduced markedly by only a moderate reduction of the oxygen content. It has been shown experimentally by Glor and Schwenzfeuer [126] that even modest reductions of the oxygen content, can increase the minimum ignition energies of dust clouds substantially. Devlikanov et al. [127] report that  $K_{St}$  is a linear function of the percentage of oxygen in the gas phase (mixture of nitrogen and oxygen).

The IChemE Guide provides information on the factors governing the selection of a suitable gas for a process involving dust. The IChemE Guide also cites certain rules of thumb relating the maximum permissible oxygen concentration for carbon dioxide ( $C_0$ ) and nitrogen ( $N_0$ ):

$$N_0 = C_0 - 2 = 1.3C_0 - 6.3$$

These rules are attributed to Germany and NFPA 69: 1978, respectively. The maximum permissible oxygen concentrations to prevent ignition, which are reported in the literature, are normally measured at ambient temperature. If it is proposed to use

inerting for dust at high temperatures ( $>100\text{ }^{\circ}\text{C}$ ), the maximum permissible oxygen concentration to prevent ignition should be determined by tests. Likewise tests are called for if there is a hybrid vapour–dust mixture. With regard to the safety margin to be employed, the IChemE Guide suggests a minimum margin of 2%. Thus if the maximum permissible oxygen concentration to prevent ignition is 11%, the oxygen concentration should be kept below 9%. It also recommends partial inerting of larger plant volumes, extended hot surfaces or high explosibility dusts (St 3). There should be continuous monitoring of the oxygen content of the gas in the plant and a trip system to shut the plant down if the concentration rises towards a hazardous level. Since the gas contains dust, there may be problems in the measurement of the oxygen level. Experiments by Wilen et al. [128] reveal that with the increasing initial pressure in the range 5–18 bar, the limiting oxygen concentration (LOC) for inerting biomass dust cloud goes up.

Schwenzfeuer et al. [129] have observed that LOC for ignition of dust clouds by electrostatic discharges, or metal sparks from mechanical impact, were significantly higher than the conservative limit determined in standard tests, using a very strong pyrotechnical ignition source.

Even as a reduction in the oxygen level at the process site can prevent dust explosions, it may introduce a suffocation risk. Adding a few vol.% of  $\text{CO}_2$  to the gas mixture reduces the critical oxygen threshold for suffocation considerably. An inert gas mixture utilizing this effect has been reported [130].

**7.1.4.2. Use of particulate or liquid inertants.** Use of solid/liquid inertants can be made to achieve two objectives: prevention of dust explosion or mitigation/control of dust explosion. To achieve the first objective, non-gaseous inertants must be added to an otherwise explosible dust in sufficient quantities to render the latter non-explosible. To achieve the second objective a non-gaseous inertant should be quickly released in adequate quantities in a process vessel, as soon as dust in that vessel catches fire to prevent the flame from propagating further.

Solid/liquid inertants are extensively used to realize the second of these objectives—this aspect has been dwelt upon at some length in Section 8.3. But the use of non-gaseous inertants in dust explosion prevention is limited. It is principally done in coal mines where rock dust (calcium carbonate, with or without magnesium carbonate) is often sprayed in mine galleries to reduce the explosion hazard posed by coal dust. The rock dust acts as a thermal inhibitor by absorbing heat from the flame front of an explosion. This quenches the flame and arrests its propagation. Smaller particles of rock dust are more effective than larger particles though very fine particles suffer from the disadvantage that agglomeration can occur [54]. Similar correlation between the size and the effectiveness of an inertant has been observed in several cases [117].

Inertant dusts are rarely used outside coal mines in preventing explosions. This is to prevent contamination of the original material. Attempts to develop inertant dusts which can go on with the main material through the process have been made; for example Mintz et al. [131] have patented a system which

uses finely powdered refractories like MgO to completely inert highly explosible dusts such as a 50:50 Al–Mg alloy.

## 7.2. Preventive steps for specific dusts/operations

### 7.2.1. Explosion protection when handling metal dusts [132]

Metal dusts suspended in air during operations like grinding pose explosion hazard which can be mitigated by the use of wet grinding or the collection of the grinded dust by a wet medium. In the manufacturing of metal powders such as aluminum or magnesium, the process of inerting technique may be employed to reduce the oxygen concentration present at the process site. Nitrogen may be used as the inerting gas in the manufacture of aluminum, and helium or argon may be used for magnesium. Other prevention practices, such as good housekeeping, grounding/bonding, hot work permits, etc. should be routinely observed. In addition, protective measures must also be taken.

### 7.2.2. Explosion protection when handling carbonaceous dusts [133]

It is very difficult to prevent formation of an atmosphere which is not susceptible to explosion hazard, in processes involving carbonaceous dusts. In theory, an explosive atmosphere can be avoided by reducing the oxygen concentration using inerting techniques. In practice, however, the very nature of the given process using carbonaceous dust may preclude inerting. For such situations, the only options are avoiding ignition sources or using explosion proof equipment.

If a mixer is closed, the circumferential speed of the mixing element need not be limited. Circumferential speeds up to 10 m/s can be tolerated during filling and emptying with a mixer not filled to less than 70% of its volume, provided that the MIT of the processed product lies above the limit values.

### 7.2.3. Precautions during dust separation

Dust separation may cause electrostatic charging which must be inhibited by the following measures:

- (i) Grounding of all conductive parts.
- (ii) Use of electrically conducting filler material in the case where the MIE  $< 3\text{ mJ}$ ; or there is a presence of flammable gas; or vapours in the air are being cleaned.
- (iii) Ensuring that all inner walls on which dust can impact at high speed do not have any insulating inner coatings with a high electrical break down strength.

## 8. Dust explosion damage control

Due to the myriad and complex ways in which dust explosions can occur, it is more or less impossible to eliminate the dust explosion hazard [90]. But control measures can drastically reduce the damage caused by the explosions, both in terms of lesser property losses and trauma, and lesser process shutdown time.

The dust explosion damage control strategies revolve round:



- (i) explosion containment;
- (ii) explosion isolation;
- (iii) explosion suppression;
- (iv) explosion venting.

### 8.1. Explosion containment

If a dust explosion can be contained within a designated space, much of the damage it may cause to the surroundings can be controlled.

Containment is an attractive option, since it is an essentially passive method and avoids the problem of relief disposal. It is not usually practicable, however, to design the whole of a dust handling plant so that it can withstand the pressures generated by dust explosions. This is particularly the case with large plants.

Nevertheless containment is practicable in small-scale units and on certain equipment. A grinding mill, for example, can be made strong enough to withstand a dust explosion. The maximum explosion pressure for most flammable gases and dusts is in the range 7–10 bar(g) [2,12]. But the static pressure is not the sole criterion; the rate of pressure rise in a dust explosion being high, the equipment must be able to withstand this dynamic loading. The equipment should be designed on the basis of rotational symmetry and avoid large flat surfaces and angular parts. Particular attention should be paid to the points at which dust is fed or withdrawn from the plant and to the connections between units.

When the powder/dust is highly toxic, complete and reliable confinement is absolutely necessary.

### 8.2. Explosion isolation

The objective of explosion isolation is to prevent dust explosions from spreading from the primary explosion location to other process units, workrooms, etc. As described earlier in Section 4.6.2, due to pressure-piling, jet-initiated high initial turbulence and turbulent jet ignition, very high pressure peaks can be generated even in generously vented vessels necessitating effective means of explosion isolation in inter-connected systems. Two approaches are commonly adopted: use of quick acting shut-off valves, and material chokes.

For explosion isolation involving quick acting shut-off valves, the valves are installed in pipes connecting one vessel with another. The valves are activated by explosion detectors which are equipped with pressure and/or optical sensors. The former type is usually preferred, since an optical detector can be blinded. But a pressure sensor may not detect very weak pressure waves which an optical sensor is capable of detecting.

The time required for the valve to close depends on the distance between the remote pressure or flame sensor and the valve. Also on the type of dust. The valve must close within 50 ms or even quicker. This level of swiftness can be achieved by using an electrically triggered explosive charge for releasing the compressed air or nitrogen that operates the valve. The slide valve must be sufficiently strong to resist the high pressures of 5–10 bar(g) that can occur on the explosion side after the valve closure. The pressures may even have a higher peak than this in the event of pressure-piling or detonation [6,134].

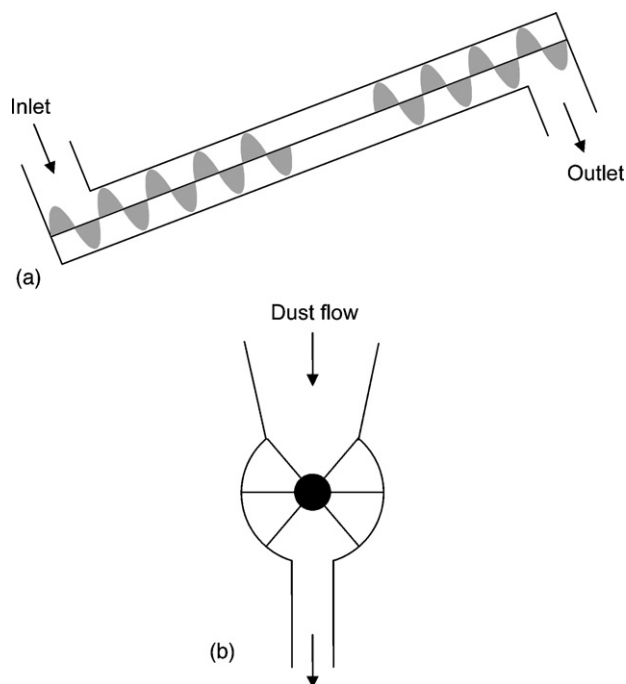


Fig. 8. Explosion isolation devices: (a) worm (screw), conveyor and (b) rotary valve.

Bartknecht [135] has discussed the ability of various types of fast-closing slide valves to interrupt dust explosions in ducts. The required closing time depends on the distance between the remote pressure or flame sensor and the valve and on the type of dust. This most often is obtained by using an electrically triggered explosive charge for releasing the compressed air or nitrogen that operates the valve. The slide valve must be sufficiently strong to resist the high pressures of 5–10 bar(g) that can occur on the explosion side after valve closure (in the case of pressure-piling effects and detonation, the pressures may transiently be even higher than this).

Isolation can also be effected with a material choke and may be considered where it is necessary to have a flow of dust between units. Two commonly used chokes are rotary valves and worm conveyors (Fig. 8). Of these, a rotary valve, which is installed vertically and is driven by a motor, is generally designed to act as a dust explosion choke even in the absence of a head of dust above it. If the valve is installed on the inlet of a hopper, there should be arrangements to prevent the former from continuing to turn if the hopper becomes overfilled, since this could cause ignition by overheating the dust or overloading the motor.

Schuber [136] and Siwek [137] have conducted extensive studies of the conditions under which a rotary lock is capable of preventing transmission of dust explosions. Schuber has provided a nomograph by which critical design parameters for explosion-transmission-resistant rotary locks can be determined. The minimum ignition energy and minimum ignition temperature of the dust must be known. However, the nomograph does not apply to metal dust explosions. Moreover, explosions of fine aluminum are difficult to stop by rotary locks.

A worm conveyor may be installed horizontally, vertically, or inclined, and is driven by a motor. One turn of the helix is

removed to prevent the choke emptying if the feed is interrupted. If the conveyor is mounted horizontally, a baffle plate is also necessary.

There have not been many tests of the effectiveness of rotary valves and worm conveyers as dust explosion chokes. Since the residence time in a worm conveyor is much longer than that in a rotary valve, the former may be expected to be more effective as a dust explosion choke, but smoldering dust could still pass through unless the motor is tripped [2,5,12].

### 8.3. Explosion suppression

If, in process equipment which harbor dust explosion hazard, a system can be put in place which gets activated as soon as an explosion begins to occur, suppresses it by swiftly adding suitable inertants, and prevents it from re-building (Fig. 9), the risk of explosions can be greatly reduced. Automatic explosion suppression devices aim to achieve this objective.

An explosion suppression system must have four basic attributes:

- (i) It should respond to an explosion with minimum time delay by getting activated quickly.
- (ii) It should inject a suppressant in adequate quantities within a very short time in a manner as to counter the incipient explosion and arrest the propagating flame.
- (iii) To shut down the plant.
- (iv) To prevent the plant from getting restarted until the explosion hazard has been mitigated.

The effectiveness of suppression system depends on how quickly the sensors respond to the emerging explosion and also on the efficiency of the suppressant injection system. The latter should be capable of injecting a large quantity of suppressant in a very short time and with adequate reach to all parts of the space being protected. It should be capable of a high mass discharge rate, a high discharge velocity and hence good 'throw' and good angular coverage.

The explosion detectors in suppression systems are similar to the ones described in the preceding section. Innovations are also done from time to time; for example Cybulski et al. [138], addressing the problem of coal dust explosions in coal mines, used solar panels for automatic detection of the coal dust flame and simultaneous actuation of water barriers. The water was contained in plastic bags, which were opened by a detonating cord triggered by the flame generated power from the solar panel. Attempts to develop ever more quick explosion detection and suppression systems also continue [79,95,117].

Detectors are available which are robust to most materials, to condensation and corrosion, and to shock. In some cases use is made of two detectors oriented in different planes. Activation occurs when the pressure reaches its threshold value, typically of the order of 0.05 bar. In some systems use is also made of the rate of pressure rise.

The mechanisms of suppression of the explosion are: (1) quenching, (2) free radical scavenging, (3) wetting and (4) inerting. Of these the principal mechanisms are quenching, or

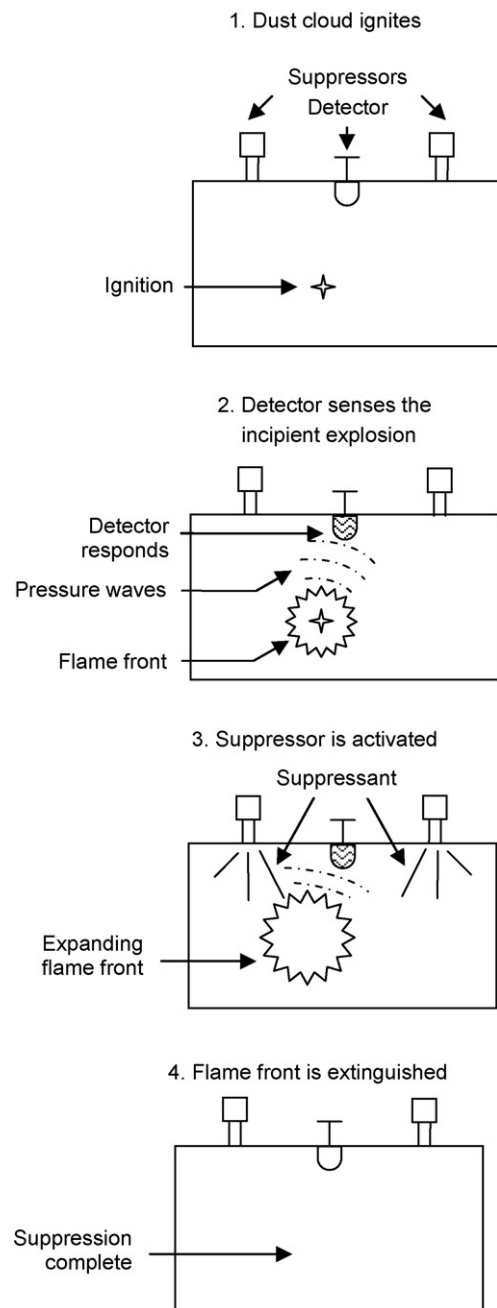


Fig. 9. Automatic suppression of an explosion in a vessel adapted from Barton [20] and Lees [12].

abstraction of heat. The contribution of free radical scavenging is specific to the particular explosion reaction. Wetting of unburned particles is applicable to liquid suppressants. There is also some inerting effect.

The response time of the sensor to the explosion depends on the size of the vessel. The explosion will propagate at the speed of sound, giving a delay, or equalization time, of the order of 4 ms/m distance between the ignition source and the detector. The response of the suppression system must be very rapid; the time taken for the suppressant to reach the flame depends on the discharge velocity, which initially is of the order of 40 m/s.

Two injection devices are in common use. One is a hemispherical suppressor with chemical detonator installed inside the vessel. This type is used with a liquid suppressant. It has a limited throw of about 2.5 m, and contains only a small quantity of suppressant. The time to complete discharge is about 10–30 ms. It is suitable for protection of smaller vessels. Multiple injection points can be more effective in containing the pressure build-up than single injection ports.

The other injection device is the high discharge rate (HDR) bottle, installed on the outside of the vessel and pressurized with nitrogen at 20–100 bar. This type is much more versatile, being able to handle most common suppressants and available in capacities from 0.003 to 0.1 m<sup>3</sup>. The time to complete discharge is about 100 ms.

There are limits to the size of vessel which can be protected by suppression; for very large vessels it may not be possible to douse propagating flames or enfeeble emerging explosions by suppressant injection. Over the years various figures have been quoted for the limiting size. The figure given in FS 6016: 1974 was 115 m<sup>3</sup>. That quoted in the IChemE Guide is about 10 times bigger: 1000 m<sup>3</sup> [139].

The design of a suppression system being a complex function of the triggering pressure, geometry of the area to be protected, nature of suppressant, the suppression system hardware, etc., computer-aided design (CAD) is being increasingly utilized to cognize and balance various contributing factors, including, of course, the costs. CAD also provides a means of quantifying the effectiveness of a deployed explosion suppression system. Likewise, insights for system design, operation, and control can be obtained through mathematical modeling [11,110,140,141]. Simple design nomograms and equations, based on the ‘cubic law’ of the maximum rate of pressure rise (Section 6.3.6), have also been derived to aid the design engineer in assessing the effectiveness of explosion suppression in practice [139].

*The common inertants.* Halons have been widely deployed in the past as both fire extinguishing agents and explosion suppressants. When released into processing equipment, halons could suppress the explosion and then provide an inert blanket of gas to preempt risk of re-ignition.

However, use of halons was fraught with the danger that if the quantity deployed was insufficient, it would not only fail to suppress an explosion but, instead, would enhance the latter’s severity. There was also a toxicity hazard because halogen acid gases would be generated as the consequence of halon dissociation [139]. For these reasons, as also their role as depletors of stratospheric ozone layers, halons have been more or less phased out of use.

Water, deployed as a rapidly injected water spray is an efficient suppressant, particularly of explosions occurring in hydrophilic dust. Water is ineffective against flammable gas explosions and against hybrid gas/dust explosions. However, if deployed very early in an explosion, water can have certain degree of effectiveness in diffusing an incipient explosion. Water has no chemical specificity and acts strictly as a thermal quenching agent. It provides minimal post-explosion inerting, save that it effectively wets hydrophilic materials rendering them less ignitable. If injected in a super-heated state, water can be more

effective because steam provides a degree of post-explosion inerting. The effectiveness of water can also be enhanced by spiking it with suitable additives.

With the passage of time, powders such as the proprietary mono-ammonium phosphate (MAP) are being increasingly preferred in explosion suppression systems. They have superior suppression effectiveness to both halons and water and have been shown to suppress even intense St<sub>3</sub> organic dust explosions and also provide protection against metal dust explosion hazards.

Dry chemical suppressants basically act as post-explosion inerting agents provided that they remain in suspension in the transport medium. They mitigate the primary explosion but do not prevent a second or subsequent ignition of flammable gases or of dust/gas hybrid explosion threats.

For the use of food processing industry, which cannot use MAP due to its potential for product contamination, effective water-soluble and food-grade-compatible inertants such as sodium bicarbonate-based dry powder suppressant (Dessikarb<sup>TM</sup>) have been developed.

The other two features associated with dust explosion suppression systems – shutting down the plant, and preventing it from restarting till it is safe to do so – are plant-specific. Standard techniques of instrumentation and control are used to link these with the pressure sensors.

Attempts to standardize the design of explosion suppression systems have led the European benchmark organization, CEN, to propose a draft guideline [142] which, apparently, provides greater flexibility than the traditional approach [95].

#### 8.4. Explosion venting

When all attempts to prevent a dust explosion have failed, the explosion would occur. If the explosion can be vented effectively, its adverse impact can be minimized. Except when toxic dusts are involved, venting can significantly reduce the destructive potential of a dust explosion.

At first glance venting appears a rather simple operation to execute. If a portion of the structure, prone to dust explosion, is made considerably less pressure-resistant than the rest of the structure by using members or thinner panels, the portion may fail in the early stages of an explosion. By sizing such a vent properly, it may be ensured that the vent becomes operative as soon as the overpressure exceeds a certain safe threshold and sufficient quantities of gas (and particulates) are let off quickly to prevent the pressure in the protected area from reaching destructive levels.

To size the vent area one needs an understanding of all the factors which determine the severity of a dust explosion, including *inter alia*, the geometry of the unit in which the provision of explosion venting is being made, dust concentration, initial pressure, initial temperature, initial turbulence, ignition source, presence of flammable gas or inert gas/dust (Table 7). All these factors influence the explosion pressure—in terms of *rate* as well as *extent* of pressure rise. In addition one needs to consider the reduced explosion pressure, vent opening pressure, vent area, vent distribution, vent opening, and vent panel. If a duct is to

Table 7  
Influence of various factors on two key parameters associated with the design of dust explosion vents

Factor	Effect on maximum explosion pressure	Effect on maximum rate of pressure rise
Vessel size and shape	Independent of vessel size, provided heat effects are disregarded. For elongated vessels, heat loss is significant and the maximum explosion pressure less	Follows the cube root law, it is inversely proportional to $V^{1/3}$ , and proportional to the surface area/volume ratio
Dust concentration	Increasing concentration leads to higher explosion pressures, up to the maximum explosion pressure after which it starts to fall	Up to the maximum rate of pressure rise, the rate of pressure rise increases with concentration, after which it decreases with further increase in concentration
Initial pressure	There is little information on the effect of initial pressure, perhaps because most dust handling plants operate at atmospheric pressures <sup>a</sup>	There is little information on the effect of initial pressure, perhaps because most dust handling plants operate at atmospheric pressures <sup>a</sup>
Initial temperature	Net effect of higher initial temperatures is to reduce the maximum explosion pressure	Increasing initial temperatures increase the maximum rate of pressure rise (due to effect on combustion rate and reduction in the moisture content)
Initial turbulence	Weak effect on maximum explosion pressure	Strong effect on maximum rate of pressure rise
Ignition source	Effect of ignition source is complex and depends essentially on the nature and strength of the ignition source	Effect of ignition source is complex and depends essentially on the nature and strength of the ignition source
Presence of flammable gas	Not much data is available on the effect of presence of flammable gas	Not much data is available on the effect of presence of flammable gas
Presence of flammable gas or an inert gas/dust	Not much data is available on the effect of presence of flammable gas	Increasing concentration of inerts decreases the maximum rate of pressure rise

<sup>a</sup> For starch dusts, the maximum explosion pressure and the maximum rate of pressure rise is proportional to the initial pressure, for initial pressures up to 2 bar.

be fixed to the vent, its impact on the explosion – venting profiles is to be considered, too. But, in spite of extensive R&D, the basic understanding of these processes inside and outside vented enclosures is still incomplete; the ways in which the dust clouds are generated or the manner in which the clouds burn, are not clearly understood either [95,143].

Furthermore, protecting the vented enclosure is not the only important concern in dust explosion venting. The blast waves and flames which are emitted into the surroundings by the vent can be hazardous, not to speak of the risk posed to the personnel by unburned dust, smoke and soot which may suddenly spew from the vent in copious manner [144–146]. The maximum flame length emitted from a vent can be up to 10 times the cube root of the vented vessel volume [147]. If too much unburned flammable material is ejected by a vent it may even get ignited by the vented flame to cause a secondary explosion [148,149]. The challenge of vent design thus extends to eliminating hazardous effects of vented material, especially the flames. Then, again, care has to be taken so that the provisions for safe venting do not interfere with the effectiveness of the venting. For example a duct if fitted to a vent may safely discharge the vented mass and energy but may, in the process, prevent sufficient dampening of the explosion pressure [95,150,151].

These perils notwithstanding, venting is arguably the most widely used method for mitigating dust explosions.

#### 8.4.1. Approaches to vent design

The design of dust explosion venting has been approached in a manner similar to the one for gas explosion venting. However, certain essential differences between gas explosions and dust explosions are cognized: whereas a gas explosion can occur

even when the gas mixture is initially quiescent, a dust explosion can occur only if there is turbulence. Also, dust clouds are rarely uniform unless they are formed under highly turbulent conditions [12].

The design approaches may be empirical (Table 8a) or theoretical (Table 8b). Future work in this field may include development of methods based on computational fluid dynamics and other computer-aided mathematical modeling techniques.

Empirical approaches, summarized in Table 8a, suffer from difficulties inherent in characterizing dust explosibility. Furthermore, dust explosions are sensitive to so many factors (Table 6) that it is very difficult to reproduce them experimentally. Nevertheless protocols for scaling up experimental studies have been developed leading to the several empirical methods summarized in Table 8a. Comparative studies on the efficacy of these methods [2,12,152] reveal that: (a) the vent ratio method tends to overestimate the explosion pressure while the vent coefficient method underestimates it; (b) the  $K_{St}$  method gives good predictions in its low ignition energy source nomographs but tends to underestimate the explosion pressure; (c) if reliable Hartman bomb data are available, the method of Schwab and Othmer yields good results. The  $K_{St}$  method can yield vent areas up to three times larger than the experimentally determined values but it has been argued that this provides a valuable safety factor, more so when the standard test conditions under which the experimental results are generated can differ substantially from the happenings in an actual plant [149]. Eckhoff and Fuhre [153] and others have shown that in most industrial processes the dust clouds produce less violent explosion than those to which the monographs apply, but the method continues to be favored due to the safety margin it generates.

Table 8a  
Empirical methods for dust explosion vent design

Method	Formula/procedure	Description	References
Vent ratio method	Graphical correlations are developed between explosion pressure in vessels and the vent ratio, $f$ , required to dampen the explosion. Where, $f = \frac{A_v}{V}$ $A_v$ is the vent area and $V$ is the volume of the enclosure.	Plots of required vent area vs. the reduced explosion pressure, covering a wide range of vessel volumes have been published. This can be used under the following conditions: 1. $P_{red} \leq 0.14$ bar(g) 2. Opening pressure and vent inertia are small	[4,6,152,190,191]
Vent coefficient method	The vent coefficient, $K$ , is given by $K = \frac{A_c}{A_v}$ $A_c = L_1 L_2$ where $A_c$ is the area of the smallest cross-section of the enclosure, and $L_1$ and $L_2$ are the two smallest dimensions of the enclosure.	The results from the experiments conducted on St 2 dusts give a good fit for $K=3$ , when vent area is plotted vs. vessel volume.	[192]
$K_{St}$ method for vessels without ducts	Nomographs for calculating vent area are developed, based on the 'cube root law'.	Three separate nomographs have been made, for vent opening pressures $P_{stat}$ values of 0.1, 0.2 and 0.5 bar(g) and reduced pressure $P_{red}$ range from 0.2 to 2 bar(g). The $K_{St}$ used here has to be measured using the 1 m <sup>3</sup> ISO standard apparatus. The nomographs can be used for the following conditions: 1. Dusts with $K_{St} \leq 300$ bar m/s, $P_{max} \leq 10$ bar(g) 2. Dusts with $300 < K_{St} \leq 600$ bar m/s, $P_{max} \leq 12$ bar(g) 3. Vessel volume $V \leq 1000$ m <sup>3</sup> 4. $P_{stat} \geq 1.1$ bar(a) 5. $P_{red} \geq 1.2$ bar(a) 6. $10 \text{ bar m/s} \leq K_{St} \leq 600 \text{ bar m/s}$ 7. Vessel length to diameter ratio less than 5:1 8. No ducts fitted to the vents	[114,193,194]
$K_{St}$ – nomograph method for vents with ducts	Four sets of graphs are used in conjunction with the $K_{St}$ – nomograph method described above	The method enables the following assessments: 1. Effect on the reduced explosion pressure of straight vent ducts 2. Effect on the reduced explosion pressure of vent ducts having a 45° bend 3. Effect on the reduced explosion pressure of vent ducts having a sharp 90° bend 4. Effect on the reduced explosion pressure of straight vent ducts for metal dusts in the St 3 group The graphs are valid for vent ducts of circular cross-section, lengths less than 16 m, and the vent cross-section area equal to the duct's. This equation is valid only if:	[144]
The draft European standard equation for single enclosures, for $P_{red} \geq 0.1$ bar(g)	$A_v = [3.264 \times 10^{-5} P_{max} K_{St} P_{red,max}^{-0.569} + 0.27(P_{stat} - 0.1)P_{red,max}^{-0.5}] \times V^{0.753} [1 + (-4.305 \log P_{red,max} + 0.758) \log (\frac{L}{D})]$		[175,195]

Table 8a (Continued)

Method	Formula/procedure	Description	References
	Where $A_v$ is the vent area, $P_{\max}$ is the maximum explosion pressure, $P_{\text{red,max}}$ is the reduced explosion pressure, $L$ is the enclosure length and $D$ the enclosure diameter.	<ol style="list-style-type: none"> <li>1. <math>0.1 \text{ m}^3 \leq V \leq 10,000 \text{ m}^3</math></li> <li>2. <math>0.1 \text{ bar(g)} \leq P_{\text{stat}} \leq 1 \text{ bar(g)}</math></li> <li>3. <math>P_{\text{stat}} &lt; P_{\text{red,max}} \leq 2 \text{ bar(g)}</math></li> <li>4. <math>5 \text{ bar(g)} \leq P_{\max} \leq 10 \text{ bar(g)}</math> for <math>K_{\text{St}} \leq 300 \text{ bar m/s}</math></li> <li>5. <math>5 \text{ bar(g)} \leq P_{\max} \leq 12 \text{ bar(g)}</math> for <math>300 \text{ bar m/s} &lt; K_{\text{St}} \leq 800 \text{ bar m/s}</math></li> <li>6. Dust cloud is homogenous</li> <li>7. <math>L/D \leq 20</math>, or minimum vent area greater than the cross-sectional area of the vessel for end venting</li> </ol>	
The draft European standard (prEN) method for vent with ducts	$\frac{P_{\text{red,max}}}{P'_{\text{red,max}}} = 1 + 17.3 \left[ \frac{A}{V^{0.753}} \right]^{1.6} L$ <p>Where <math>A</math> is the vent area (<math>\text{m}^2</math>), <math>V</math> is the vessel volume (<math>\text{m}^3</math>), <math>L</math> is the vent duct length to diameter ratio, <math>P_{\text{red,max}}</math> is the reduced explosion pressure without a vent duct (bar), and <math>P'_{\text{red,max}}</math> is the reduced explosion pressure with a vent duct (bar)</p>	<p>This method is applicable under the following conditions:</p> <ol style="list-style-type: none"> <li>1. The initial sizing of the vent, before the duct is fitted, should have been done by using the method described above</li> <li>2. The duct is straight, without any bends</li> <li>3. The duct is fitted to an isolated enclosure</li> <li>4. <math>0.1 \text{ bar} \leq P_{\text{red,max}} \leq 2 \text{ bar}</math></li> <li>5. The vessel is cubicle</li> <li>6. Dust–air mixture is homogenous</li> </ol>	[195]
NFPA formula for dust explosions in buildings	$A_v = \frac{C_1 A_s}{P_{\text{red}}^{0.5}}$ <p>Where <math>A_v</math> is the vent area, <math>A_s</math> is the total internal surface area of the enclosure and <math>C_1</math> is a factor whose value depends on the <math>K_{\text{St}}</math> of the dust</p>	This formula applies principally to buildings, but also to low-strength dust handling plants and similar enclosures [20], $P_{\text{red}} < 0.1 \text{ bar}$	[196]
Schwab and Othmer method	$Y = \frac{A}{10^{KX}}$ <p>Where <math>A</math> is the maximum pressure in the unvented explosion (psi), <math>X</math> is the vent ratio (<math>\text{ft}^2/100 \text{ ft}^3</math>), <math>Y</math> is the reduced pressure in the vented explosion (psi) and <math>K</math> is the slope of the line of the maximum pressure (<math>A</math>) vs. vent ratio (<math>X</math>)</p> <p>This is not a strictly empirical method but has some theoretical basis: the rate of volume increase is equated with volumetric vent outflow, leading to the Runes equation:</p>	Nomographs have been developed based on this method to determine vent ratio, given the maximum explosion pressure, maximum explosion pressure in vented explosion and the maximum rate of pressure rise.	[197]
Runes' method	$A_v = \frac{CA_c}{(\Delta P)^{1/2}}$ <p>Where <math>\Delta P</math> is the explosion overpressure, <math>A_c</math> is the area of the smallest cross-section of the enclosure and <math>C</math> is the Runes' constant</p> <p><math>A = \frac{0.019V^{0.635}}{P_{\text{red}}^{0.5}}</math> for <math>\left(\frac{dP}{dt}\right)_{\max} \leq 300 \text{ bar/s}</math>, and hinged vent panels have a maximum mass of <math>20 \text{ kg/m}^2</math></p> <p><math>A = \frac{0.044V^{0.685}}{P_{\text{red}}^{0.5}}</math> for <math>300 &lt; \left(\frac{dP}{dt}\right)_{\max} &lt; 600 \text{ bar/s}</math>, and hinged vent panels have a maximum mass of <math>12 \text{ kg/m}^2</math></p> <p>Where <math>A</math> is vent area (<math>\text{m}^2</math>), <math>V</math> is volume (<math>\text{m}^3</math>) and <math>P_{\text{red}}</math> is in bar (g)</p>	NFPA (1978) has reported Runes' constant for organic dusts and high flame speed metal dusts. Later in 1994, NFPA (1994) has given a relationship similar to Runes' equation containing a constant, $C$ , of which values for different St dusts have been reported.	[191,193,198]
Swedish method	<p>Where <math>A</math> is vent area (<math>\text{m}^2</math>), <math>V</math> is volume (<math>\text{m}^3</math>) and <math>P_{\text{red}}</math> is in bar (g)</p>	The $\left(\frac{dP}{dt}\right)_{\max}$ to be used in this method has to be determined from standard 1.2 Hartmann bomb tests.	[6,199]

Table 8b  
Theoretical methods for dust explosion vent design

Method	Formula/procedure	Description	Reference
Heinrich method	<p>The vent area, <math>A_v</math>, is given by:</p> $A_v = \frac{V^{2/3} V_L^{1/3} (dP_{ex}/dt)_{P_{red}} \cdot V_L}{C_d (2RT/M)^{1/2} P_{red}^{1/2} (P_{red} - P_a)^{1/2}}$ <p>Where <math>V</math> is the volume of the vessel to be vented, <math>V_L</math> is the volume of the test vessel, <math>C_d</math> is the coefficient of discharge, <math>M</math> is the molecular weight of the gas, <math>T</math> is the absolute temperature of the burnt gas, <math>P_a</math> is the atmospheric pressure, <math>P_{red}</math> is the reduced pressure and <math>P_{ex}</math> is the pressure due to the explosion.</p>	This formula has been derived by equating the rate of generation of burned gas to the mass velocity of vent outflow, assuming isothermal conditions. Nomographs have been developed based on this approach.	[200]
Palmer's method	<p>For the case when the venting pressure is close to atmospheric pressure (low pressure case):</p> $P - P_0 = \frac{2.3 P_0 \rho_c}{C_d^2 \gamma^2 P_{max}^3} \left( \frac{V}{A_v} \right)^2 \left( \frac{dP}{dt} \right)_{max}^2$ <p>For the high pressure case:</p> $\frac{1}{P - P_0} = \frac{1}{P_{max} - P_0} + \frac{K A_v P_{max}}{0.8 V \rho_c (dP/dt)_{max}}$ $K = C_d \left[ \frac{\rho_0 \gamma}{P_0} \left( \frac{2}{\gamma + 1} \right)^{(\gamma + 1)/(\gamma - 1)} \right]^{1/2}$ <p>Where <math>P_0</math> is atmospheric pressure, <math>\rho_c</math> is the density of the unburned gas at <math>P_0</math>, <math>C_d</math> is the coefficient of discharge, <math>\rho_0</math> is the density of the gas at atmospheric pressure, <math>\gamma</math> is the ratio of specific heats, <math>P_{max}</math> is the maximum unvented explosion pressure, <math>V</math> is the volume of the enclosure, <math>A_v</math> is the vent area</p>		[201]
Rust's method	$A_v = \frac{k F (P_{max} V)^{2/3} K_D^{1/3}}{P_{red}^{1/2}}$ <p>Where <math>A_v</math> is the vent area, <math>K_D</math> is an explosion parameter which can be determined from the pressure rise in tests, <math>F</math> is a shape factor to take into account non-spherical vessels, <math>k</math> is a constant (<math>= 8.35 \times 10^{-5}</math>, when <math>A_v</math> is expressed in FPS units), <math>P_{max}</math> is the absolute maximum pressure of the unvented explosion and <math>P_{red}</math> is the reduced pressure (gauge).</p>	This method was derived for the venting of dust explosions in low-strength plants, for a dust mixture initially at atmospheric pressure.	[202]
Nagy and Verakis' method	<p>For the case of dust explosion in a cylinder with an open vent, assuming that the ignition takes place at the closed end and there is subsonic flow of the vented burnt gas:</p> $\frac{P_{red}}{(P_{red} - P_0)^{1/2}} = \frac{RT_u k_{v1} P_0 L (A_v / V)}{T_b^{1/2} \alpha S_u (P_m - P_0)}$ <p>For the case of initially closed vent, with all the other assumptions same as above, the equation is:</p> $\frac{P_{red} + P_v}{(P_{red} - P_0)^{1/2}} = \frac{RT_u k_{v1} P_0 L (A_v / V)}{T_b^{1/2} \alpha S_u (P_m - P_0)}$ <p>Where <math>T</math> is the absolute temperature, <math>P</math> is the absolute pressure, <math>V</math> is the vessel volume, <math>A_v</math> is the vent area, <math>L</math> is the length of the cylinder, <math>k_{v1}</math> is a constant, <math>S_u</math> is the burning velocity, <math>\alpha</math> is the coefficient of turbulence, the subscripts b, m, o, u denote burned, final, initial and unburned, respectively, <math>P_v</math> is the vent opening pressure.</p>	This model has been adapted from the one developed by the authors for gas explosion.	[107]

Note:  $P_{red}$  the residual overpressure from a vented explosion,  $P_{stat}$  is the static activation gauge pressure.

The theoretical methods for dust explosion vent design (Table 8b) also rely heavily on experimental data [2,12]. The Heinrich method gives upper limits of the reduced pressure for St 1 and 2 dusts, but should not be used for St 3 dusts. The Palmer method gives good predictions for St 1 and 2 dusts, but tends to underestimate the reduced pressure for St 3 dusts in larger vessels. The Rust method is better for St 1 dusts than for St 2 and 3 dusts, and can be highly inaccurate in some cases. The Pittsburgh method requires information on the burning velocity and turbulence factor which are generally not available; the method is therefore, of limited application [2,12]. Tamanini [154], in the course of a comparative study of scaling parameters of vented gas and dust explosions has recorded that different vent sizing formulae (tables) lead to proportionality between the reduced pressure and the peak unvented pressure raised to a power between  $-5/3$  and  $-1$ .

An excellent do-how manual on dust explosion control, especially vent design, has been produced by Barton [20].

#### 8.4.2. Factors influencing the venting process

Several factors, besides the dimension of the vent opening, influence the effectiveness of dust explosion venting.

**8.4.2.1. Vent ducts.** Ducts have to be attached with most vents so that the blast waves, unburnt material, and flame exiting from a vented dust explosion do not injure staff working outside the vented enclosure, damage other units, or cause secondary explosion. But flow through the vent is impeded to some extent by ducts, causing an increase in the explosion overpressure inside the vented plant. The increased overpressure may be due to a secondary explosion in the duct (burn-up) [155,156], frictional drag and inertia of the gas column in the duct [157], acoustic [158] and Helmholtz oscillations [159].

The turbulent mixing of hot and fresh gases in the initial section of the duct after the flame entrance promotes a violent burning therein (an explosion-like combustion or 'burn-up'). The resultant pressure impulse in the duct induces backflow of gases from the duct to the vessel with the possible consequent turbulence of residual combustion in the vessel and the blockage of the gas efflux. Some authors have singled out 'burn-up' as the main factor responsible for the dramatic increase of the pressure in the vented vessel [160,161]. Some others believe that the additional pressure drops due to the resistance of the gas flow in the vessel–duct assembly is the main cause for the higher pressure rise in the ducted vessel with respect to simply vented vessels [162,163]. Substantial pressure drops may occur due to the very high flow velocities attained at the duct entrance and the concentrated losses in the sudden flow area changes at the duct exit. Zero-dimensional and one-dimensional mathematical models [155,162] suggest that the enhancement of the burning rate through turbulence and the friction losses are the most important phenomena affecting overpressure [164].

Ferrara et al. [165], using a CFD model based on the unsteady Reynolds Averaged Navier Stokes (RANS) approach for the numerical simulation of a ducted explosion, saw that the burn-up related effects are the key phenomena in determining the pressure rise in the ducted venting configuration. This was in

agreement with the experimental findings of Ponizy and Leyer [156] and Molkov [166,167]. It was seen that the reduction of venting rate (mechanical effect) rather than the burning rate enhancement through turbulence (combustion related effect) was responsible for the recorded overpressure. Simulation, carried out by varying the duct diameter and length and the ignition position revealed that the latter affected strongly the vessel overpressure through the combustion of the residual unburnt mixture in the vessel after burn-up. With respect to geometrical parameters, numerical flow field representations indicated that larger duct sections are not *a priori* beneficial to relieve the pressure in the vessel, due to the flame distortion effects [165].

In a study with four types of dusts varying widely in explosion characteristics, and ducts of two sizes, Lunn et al. [149], have observed that the difference between the vent duct area and the effective vent area, and the explosion characteristics of the dusts are important factors in decreasing the effect of vent ducts below the levels predicted by theory. The logistics of an industrial unit often makes it necessary for the ducts to have one or more bends. Such bends hamper reduction in the explosion pressure; closer a bend from the vent, greater the effect.

Design considerations for vents with and without ducts have been reviewed by Barton [20]. Vent ducts with area less than, or much greater than, the vent area are forbidden. Gradual bends are recommended instead of sharp bends as their interference in the venting efficiency is lesser. Multiple bends must be avoided as far as possible; their effect on the reduced explosion pressure is far greater than a single bend. No bend should be located closer than 2 m from the vent opening. Explosion pressure in the vent duct can be as high as that in the explosion vessel; some pressure pulses in the duct can even be higher. The ducting should be made correspondingly strong to withstand these pressure loads.

Even light-weight weather-protection devices fitted to vent ducts can hamper the vent's action but a grating over the end of a vent duct (to keep out pests) does not have a noticeable effect on the explosion pressure.

**8.4.2.2. Recoil.** When a dust explosion is being vented, the flow exiting from the vent produces reaction forces in the direction opposite to that of the flow. If there is a resulting imbalance, a net force would develop in the nature of a recoil which needs to be counteracted by providing appropriate supports to the structure.

Expressions correlating maximum recoil force developed, vent area, maximum pressure rise inside the vented enclosure, and other related parameters have been developed by Faber [168] and Harmanny [169]. Building upon this work, Tamanini and Valiulis [151] have developed a new correlation which links the average recoil force,  $F_R$ , the duration,  $t_D$ , and the impulse,  $I$ , of vented explosions as under:

$$I = F_R t_D, \quad F_R = 0.6(p_r - p_c)A_v$$

and  $t_D = \frac{v(p_m - p_0^{1/2})}{a_{cd} A_v (p_r - p_0)}$  where  $p_0$ ,  $p_c$ ,  $p_m$ , and  $p_r$  are initial pressure, external pressure, maximum pressure, and maximum unvented pressure, respectively;  $V$  ( $m^3$ ) the volume of the vented enclosure, and  $a_{cd}$  is a dimensional constant equal to 232.5 m/s. According to the authors, their correlation has a surer theoretic



cal foundation and has been more rigorously validated than the earlier treatments.

Their preliminary results have shown that the correlations also apply to situations where final inertia effects are present, provided that such effects are contained.

If a vent is fitted with a duct, the recoil is experienced by the duct as well. Lateral reaction forces also occur if the duct contains a bend. Hence provision for withstanding the reaction forces must be made for ducts as well as the vents.

**8.4.2.3. Venting devices.** Venting devices range from light-weight burst disks or membranes to much heavier explosion doors. An ideal device should begin to open only at a predictable pressure. The inertia of the device should be such that it begins to open when the overpressure reaches a threshold level.

A venting device should not open inadvertently or leak significant quantities of dust but should open quickly enough when it is supposed to [20].

It is preferable that vent areas are located on different walls of a building rather than be all placed on the same wall. Restrictive devices such as a hinge or a chain should be attached to vent panels to prevent the panels from flying off in the event of a vented explosion. As the panels can swing outwards violently, get unhinged and propelled, or even shatter into flying fragments during the venting process, consideration should be given to the space into which the panel would open. Materials of a brittle nature which may shatter into pointed projectiles should not be used for vent panels.

Vent closures should be pre-tested both for their mechanical strength and opening pressure,  $P_{\text{star}}$ .

The venting process is strongly influenced by the inertia of the vent closure, expressed as  $\text{kg/m}^2$ . As this increases, the venting process is progressively impeded and, for a specified vent area, the reduced explosion pressure,  $P_{\text{red}}$ , begins to rise.

The effect of inertia is determined by comparing the venting effectiveness of a given vent closure with one that is essentially inertia free and so has a venting efficiency of 100%; for example, bursting diaphragm. The venting efficiency is given by  $(A_e \times 100/A_d)\%$ , where  $A_d$  is the geometrical vent area of the vent closure and  $A_e$  is the effective vent area. Thus, in practice, the geometric vent area necessary to give safe venting of a given enclosure would be greater than the vent area estimated by the vent sizing methods to an extent depending on the venting efficiency. If the estimated vent area necessary to protect a given enclosure was, say,  $2 \text{ m}^2$ , then a geometric area of  $4 \text{ m}^2$  would be necessary if the vent closures had a venting efficiency of 50% [20].

It has been experimentally determined that, except at very small volumes, the effect of panel inertia up to  $10 \text{ kg/m}^2$  is negligible. Thus, for closures with inertia of  $\leq 10 \text{ kg/m}^2$ , 100% efficiency can be assumed in vent area calculations [20,170,171].

**8.4.2.4. Turbulence.** Effect of turbulence on explosion venting has been studied extensively [5,121,170,172–174]. Three types of turbulence have been identified:

- (i) Turbulent motions which exist before ignition of the flammable mixture, for example these driven by fans.
- (ii) Turbulence generated during flame propagation due to different phenomena like flame wrinkling, expansion of hot gases, flame stretching, obstacles and others.
- (iii) Turbulence created by the ejection of gas during venting.

The interactions between flow field, turbulence, and flame are not only complex but very from process equipment to process equipment. Even pre-ignition turbulence of a moderate level (about  $0.45 \text{ m/s}$ ) may lead to a considerable increase in the reduced explosion pressure and in the maximum rate of pressure rise during a vented explosion [174].

**8.4.2.5. Venting of interconnected vessels.** Most industrial dust handling plant consists of vessels linked by pipelines. As detailed earlier in Section 4.6.2, if dust explosion occurs in one of the vessels, there is increased turbulence and pressure-piling in the other vessels which may cause secondary explosions to be more severe than the primary explosion.

The Germany-based institution VDI in its guidebook no. 3673 [175] has given the following rules of the thumb:

- (i) When the larger of the enclosures cannot be vented, then the entire system must be designed for full containment.
- (ii) When the smaller of the enclosures cannot be vented, then it must be designed for containment and the vent area of the larger vessel determined directly from representative explosion trials or appropriate published data.
- (iii) When the enclosures are of equal size, and one enclosure cannot be vented, (ii) applies.

Holbrow et al. [176] have given guidelines for estimating the vent areas for some linked systems based on the results of an experimental investigation. The following rules have been derived that can be applied up to vessel volumes of  $\sim 20 \text{ m}^3$ :

- (i) For  $K_{St}$  values of  $150 \text{ bar m/s}$  or less, a dimensionless vent area ( $A_D$ ) in both vessels of greater than 0.25 will limit the reduced explosion pressure to 0.5 bar.  $A_D = A_v/V^{2/3}$ , where  $A_v$  is the vent area and  $V$  is the vessel volume.
- (ii) For  $K_{St}$  values between 150 and  $250 \text{ bar m/s}$ , a value of  $A_D$  in both vessels of 0.4 will limit the reduced explosion pressure to 0.5 bar.

The venting area shall be divided between enclosures so that  $A_D$  is the same in both enclosures. When venting a system of linked enclosures, the venting devices should be designed for a low static activation overpressure,  $P_{\text{stat}} \leq 0.1 \text{ bar}$ .

As the vent area decreases, the relative effect of linking the vessels increases. The increase in pressure is greatest when primary ignition occurs in the larger of the linked vessels. Generally, the longer the pipe the less is the effect on the explosion pressure, but this is not always so [20].

Even as, in recent years, our understanding of the dust explosion venting process has increased considerably, this has not provided us with simple, rule-of-thumb answers to vent design

problems. On the contrary, more the new experimental evidence is generated, more we realize that dust explosion venting is an exceedingly complex process. What may happen with a given dust under one set of practical circumstances may be totally different from what may happen in others [6]. The classic dilemma in all safety-related design efforts, viz. how to balance the fear of failure with the extra cost of over-design continues to dog the field of dust explosion venting as well!

### 8.5. Expert systems

The early attempts to develop expert systems (software) for dust-explosion vent design included Dust-Expert [177,178] and ExTra [179]. Over the years newer versions of these software have been developed [180,181] as also newer systems have appeared [182,183] but they all continue to be based on  $K_{St}$  inputs.

Hesener et al. [181] and Kraus et al. [184] have developed an expert system to be used to identify hazards due to the possible occurrence of various types of electrostatic discharges in various process situations. The system, using the CEN-ELEC report R044-001 as its technical basis, covers explosive gases/vapours/mists as well as explosible dusts.

Lorenz [185] has presented an expert system for the design of explosion venting arrangement based on the VDI 3673 venting code, which is very close to the new European Union code produced by CEN TC 305. The system accounts for the inertia of vent covers and doors and assesses forces acting on these covers and doors. The extent to which debris is ejected into the surroundings by destructive explosions is also accounted for. Lorenz and Schiebler [186] have presented an expert system incorporating the four programs – SIMEK, STS, Vent, and Vessel – dealing with the design of explosion doors and lids, ejection of debris, venting assuming ‘zero mass’ vent covers, and the stability of pressurized enclosures, respectively.

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