Explosion Hazards in the Process Industries

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HOUSTON, TX:
Gulf Publishing Company
2 Greenway Plaza, Suite 1020
Houston, TX 77046

AUSTIN, TX:
427 Sterzing St., Suite 104
Austin, TX 78704

10 9 8 7 6 5 4 3 2 1

Library of Congress Cataloging-in-Publication Data

Eckhoff, Rolf K.

Explosion hazards in the process industries : why explosions occur and how to prevent them, with case histories / Rolf K. Eckhoff.

p. cm.

Includes bibliographical references and index.

ISBN 0-9765113-4-7 (acid-free paper)

1. Explosives–Safety measures. 2. Hazardous substances–Accidents–Prevention. 3. Explosions–Case studies. I. Title.

TP270.E25 2005

604.7–dc22

2005012133

Printed in the United States of America

Printed on acid-free paper. ∞

Production services and design by TIPS Technical Publishing, Inc.

To my dear wife Astrid for her everlasting love and support, to our five grandchildren Gunhild, Sigurd, Johanne, Elias and Maren, to the precious memory of the two, Åshild and Nora, who we so sadly lost, to the one that is on its way, and finally to those who we hope are still to come! The words in Isaiah 42.16 have continued to give me hope and courage.

Preface

My very first contact with the area of industrial explosion hazards was during my post graduate studies at King's College, London 1966–68. In the early 1970s, Norwegian process industries experienced a series of serious dust explosions, which faced me with the challenge of establishing dust explosion prevention and mitigation as a new field of research and consultancy in Norway, based at the Christian Michelsen Institute in Bergen. A few years later, when the oil and gas industry on the Norwegian continental shelf evolved at great pace, I was faced with the further challenge of starting up CMI's comprehensive research program on prevention and mitigation of accidental gas explosions.

In 1991 I approached the University of Bergen (UoB) to discuss the possibility of establishing process safety technology as a new field of teaching and research there. I was appointed part-time professor in 1992 and full-time professor in 1996, with my activity based in the Department of physics and technology. While building up my own activity at UoB, I was also taking a central role in establishing a common framework program of process technology at large within our faculty of mathematics and natural sciences. We were convinced that our candidates should build their process safety studies on a basement of general process technology (chemical engineering). The final result of a joint effort by many people is a complete five-year study program in process technology, comprised of a three-year bachelor program providing the general basis, followed by a two-year master program providing the specialities. Process safety technology is one of several options for the master studies.

During my time as part-time professor, I developed a short course on "Electrical apparatuses for hazardous areas", published as a 160-page basic textbook in Norwegian (Tapir, Trondheim, 1996). After joining UoB full time I used this book as a central element in a basic bachelor-level course on explosion hazards in the

process industries but I then had to hand out supplementary material on case histories and prevention/mitigation. I therefore decided to write a complete textbook, comprising all the material needed for this course. Since the book might also be useful outside Norway, I chose to write it in English.

Whereas the present book covers accidental gas/vapor, spray/mist, dust, and explosives/pyrotechnics/propellants explosions it does not treat various types of accidental physical explosions, BLEVEs and explosions due to run-away chemical reactions.

It is my plan to produce a Powerpoint presentation covering the material in this book, with some of the photographic material in color, and with additional illustrative material. This will primarily be for my own use, but such a presentation may also be helpful for other teachers wanting to use my book in their classes. At UoB we are also building up a set of about ten laboratory-scale experiments on initiation and propagation of gas and dust explosions, which the students shall have to carry out and report on in writing as part of the course.

I am deeply grateful to the companies, research institutions, standardization organizations, publishers etc., that granted permissions to use some of their valuable material for the book, and to the many skilled persons, not least some of my students and colleagues at UoB, for providing valuable information and advice.

—Rolf K. Eckhoff

About the Author

Rolf K. Eckhoff is professor of process safety technology in the department of physics and technology, at University of Bergen, Norway, concurrent professor at Northeastern University, Shenyang, China, and scientific adviser at Øresund Safety Advisers AB, Malmø, Sweden. He is the author of "Dust Explosions in the Process Industries" first published 1991. An extended third edition (720 pp) appeared in 2003. He is also the author and co-author of more than eighty technical and scientific publications and more than 130 research reports. He has advised industry on numerous occasions, and lectured to courses, conferences, and seminars in a number of countries in Europe and in Bahrain, Israel, China, India, Australia, Canada, and the United States.

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Chapter 1

Introduction

1.1 Process Safety—A Persistent Challenge to Educators

Right from the start of the development of the oil and natural gas industry on the Norwegian continental shelf, very high safety standards were established. These standards were a matter of both industry attitude and official national policy. Until the "Piper Alpha" catastrophe, it was felt by some that Norway was overdoing safety issues in its offshore industry. However, after this accident, the imposition of strict safety requirements in this industry has gained wide international acceptance. Moreover, it has also been pointed out that substantial benefits would result if the high standards of safety in the offshore industries could be adapted to industry onshore and to society at large.

However, high safety levels cannot be established once and for all by a single all-out effort. Deterioration results if the high level—once attained—is not actively secured by continuous maintenance and renewal. This applies both to safety technology and to human factors.

Education has a key role in the continuous maintenance and renewal process. This ranges from short practical training courses to in-depth long-term education. Universities and colleges have responded to the challenge by establishing courses of study on a wide range of safety aspects. In the case of process safety, relevant topics include reliability and risk analysis, the physics, chemistry and technology of processes and hazards, and means of accident prevention and mitigation. Much emphasis has been put

on methods of reliability and risk analysis, which are indeed very important. However, it is sometimes felt by the process industry itself that education in the "hard" aspects, i.e. the physics, chemistry, and technology of processes and process hazards has been somewhat left behind. This situation presents a special challenge to universities and colleges. Nine years ago my own university established a course of studies in process safety technology, with particular emphasis on the scientific and technological aspects.

In process safety, the prevention of fires and explosions, and the mitigation of their effects, is a central concern. Most often loss of confinement of flammable/explosive substances is the first link in the accidental chain of events. The next step is generation and ignition of flammable clouds, resulting in explosions and fires, which may in turn cause loss of life and limb and damage to the process plant, adjacent process areas, and even more remote building structures. Understanding the processes of accident escalation is an important aspect of process safety technology. Quantitative risk analysis plays an increasing role in the effort to improve offshore process safety. In the offshore oil and gas industries, a highly packed, congested process plant, with compact living quarters as a close neighbor, demands very systematic and thorough analysis of all possible risk factors, including the human elements. A concise and constructive official authority policy constitutes an important basis for ensuring the necessary high level of safety.

The purpose of this book is to provide a source of basic information on the origin, course, prevention, and mitigation of accidental explosions in the process industries. Potential readers/users of the book should include people both from a wide range of process industries, official authorities, engineering companies, and, not least, students in technical colleges and universities.

1.2 What Is an Explosion?

The concept of explosion is not unambiguous. Various encyclopedias give varying definitions that mainly fall in two categories. The first focuses on the noise or "bang" due to the sudden release of a strong pressure wave or blast wave. The origin of this pressure wave, whether a chemical or mechanical energy release, is of secondary concern. This definition of an explosion is in accordance with the basic meaning of the

word ("sudden outburst").The second category of definitions relevant in the present context is explosions caused by a sudden release of chemical energy. This includes explosions of gases and dusts and solid explosives. The emphasis is then often put on the chemical energy release itself, and explosion is defined accordingly. A possible definition could then be "An explosion is an exothermal chemical process that, when occurring at constant volume, gives rise to a sudden and significant pressure rise."

In the present book the definition of an explosion will shift pragmatically between the two alternatives, by focusing on either cause or effect, depending upon the context.

1.3 Accidental Explosions—A Real Hazard in the Process Industries

The industries or facilities in which gas, spray/mist, or dust explosions may exist include:

Oil and natural gas industries/activities

- oil and natural gas production installations on and offshore
- oil and gas refineries
- systems for transportation of oil and gas (pipelines, ships, trains, cars etc.)

Petrochemical, chemical, and metallurgical process industries

- petrochemical industries producing chemicals and polymers
- plants producing pharmaceuticals, pesticides, organic pigments etc.
- paint production plants
- pulverized metal production (aluminum, magnesium, silicon, and silicon alloys etc.)
- chemical food and feed production
- production of cellulose, paper etc. from wood

Mechanical processing

- grain and feed storage
- flour mills
- sugar refineries
- mechanical wood refining (hardboard etc.)

Special processes

- production, storage, and handling of explosives, pyrotechnics, and propellants

1.4 Basic Differences in How and Where Explosive Gas, Spray/Mist, and Dust Clouds Are Likely to Be Generated

1.4.1 Similar Ignition and Combustion Properties of the Various Clouds

Explosive gas mixtures and explosive clouds of sprays/mists and dusts, once existing, exhibit very similar ignition and combustion properties, such as

- flammability/explosibility limits
- laminar burning velocities and quenching distances
- the response of the burning velocity to cloud turbulence
- detonation phenomena
- adiabatic constant-volume explosion pressures of similar magnitudes
- well-defined minimum ignition energies, and
- minimum ignition temperatures for given experimental conditions

Recognition of these similarities may have contributed to the development of the idea that the hazards of accidental gas, spray/mist, and dust explosions can be regarded as more or less identical. As discussed in Section 1.4.2, this is a misconception. Also, there is a basic difference in the ranges of hazardous fuel concentrations between dusts, sprays/mists,

and gases. For combustible gases and sprays/mists, flame propagation is only possible when the fuel to air mixing ratios lie between the lower and the upper flammability limits. Dust flame propagation, however, is not limited only to the flammable dust concentration range of clouds. The state of settled layers and deposits constitutes an additional singular regime of flame propagation. This is because, contrary to combustible gases and liquids, settled powders/dusts will always have some air trapped in the voids between the particles, which makes it possible for sustained, although often very slow, combustion to propagate throughout the deposit.

1.4.2 Influence of Inertial Forces on the Movement of Dust Particles and Liquid Droplets

Once a combustible gas has been homogeneously mixed with air, the mixture will for most practical purposes stay homogeneous, due to random molecular motion. In clouds of dust particles and liquid droplets, however, the fuel particles are generally so much larger than the molecules of the air (often in the range 1–100 μm), that their movement within the air is controlled by inertial forces, including gravity, rather than by random molecular motion. The role of inertial forces increases systematically with increasing particle or droplet size and increasing density of the particle or droplet material. Turbulence and other convective movement of the air can prolong the time over which the particles will stay in suspension. When liquid droplets in a cloud collide, the droplets may coalesce and form one larger drop, which may require special considerations.

1.4.3 Fundamental Differences between the Ways Explosive Clouds Are Generated

There are fundamental differences in the ways and circumstances in which clouds of the three different fuel categories are generated and sustained, which have a major impact on the choice of means by which accidental explosions should be prevented and mitigated. The paramount question is whether there will be an explosive cloud in the first place. The physics of generation and sustainment of clouds of dusts, sprays/mists, and premixed gas clouds are substantially different.

In the case of *gases* and *vapors*, the fuel mixes with the air on the molecular level. Explosive clouds are generated readily when combustible gases and vapors are accidentally released into the atmosphere. Gas and vapor cloud explosions can be initiated inside process equipment, but most often the initiation is in accidental clouds generated outside process equipment, following loss of confinement due to leaks or equipment failure.

Explosive *spray* clouds will also, in most cases, be generated outside process equipment, in situations where pressurized combustible liquids are accidentally released from process equipment through narrow holes, slits, or cracks, and the liquid is broken up into fine droplets mechanically. Explosive *mists* are formed when hot mixtures of combustible vapor and air become cooled, and some of the vapor condenses out as very fine droplets. This may occur both inside and outside process equipment.

In the case of *dusts*, however, primary explosive clouds are practically exclusively found inside process equipment. Inside mills, mixers and blenders, bucket elevators, pneumatic conveying systems, silos and hoppers, cyclones, and filters the dust/powder can be kept in suspension more or less continually by rotation of the whole unit, movement of inserts, or an air flow. Therefore, explosive dust clouds may exist more or less continually in normal operation due to the basic nature of the operation.

Primary dust clouds generated outside process equipment by leaks from pressurized equipment are rare and do not play any significant role as far as primary dust explosions are concerned. The duration of the process of cloud generation is then normally very short (e.g., pouring or discharging operations, accidental bursting of sacks and bags). The dust particles will start to settle out of suspension as soon as the cloud generation process terminates, and typical total lifetimes of explosive primary dust clouds outside process equipment will be of the order of fractions of a minute. One exception would be long-duration minor dust leaks from, for example, flanges in pneumatic transport lines, but the dust cloud volumes produced in such cases would normally be quite small.

However, dust layers accumulated outside process equipment present a hazard of secondary dust clouds and secondary dust explosions if such layers can be thrown into suspension by blasts from primary explosions initiated inside process equipment.

1.4.4 Migration of Dust Particles and Liquid Droplets through Narrow Holes and Gaps in Enclosure Walls

Because dust particles and liquid droplets are so much bigger than gas molecules, they will not travel through narrow holes and slots of the order of 1 mm diameter and smaller in the same way as gas molecules will do. In principle, dust particles and liquid droplets may be carried through narrow passages by the air flow generated by a moderate pressure difference across the passage. However, both dust particles and liquid droplets will easily adhere to the area around the passage entrance and to the passage walls and eventually block the passage. In the case of liquid droplets, they will most often coalesce and form a liquid film.

Dust particles or liquid droplets that have been able to pass through narrow holes or gaps in this way will not remain suspended in the air inside the enclosure and form an explosive cloud. Instead, they will settle out of suspension and form a dust layer or a liquid film. In the case of dusts, it is also difficult to envision any mechanical process inside typical electrical apparatus enclosures that could possibly redisperse such dust layers into explosive dust clouds within the enclosure. In the case of liquid droplets, liquid films will be formed.

1.5 European Union Definition of Explosive Atmospheres

The two European Union directives "Atex 100a" (1994) and "Atex 118a" (1999) define explosive atmospheres as follows: Mixtures with air, under atmospheric conditions, of flammable substances in the form of gases, vapors, mists or dusts in which, after ignition has occurred, combustion spreads to the entire unburnt mixture.

Unfortunately the two directives are vague regarding clarification of the basic differences in the ways in which explosive clouds of gases/vapors, spray/mists, and dusts are, and can be, generated in industrial practice. Such a clarification is an essential basis for giving adequate differentiated guidance on selection of suitable means for preventing gas/vapor, spray/mist, and dust explosions. This not least applies to design of electrical apparatus (see Chapter 7).

1.6 The "Human Factor"

Proper build-up and maintenance of an integrated system for preventing and mitigating explosions in the process industries very much depends on human relations and human behavior.

A number of different personnel categories may be involved, including

- workers in the plant
- foremen in the plant
- workers from the maintenance department
- plant engineers
- safety engineers
- purchasing department officers
- safety manager
- middle management
- top management
- suppliers of equipment
- explosion experts/consultants

Adequate prevention and mitigation of accidental explosions cannot be realized unless there is meaningful communication between the various categories of personnel involved. If such communication is lacking, the result can easily become both unsatisfactory and confusing.

In general terms, meaningful communication may be defined as conveyance and proper receipt and appreciation of adequate information whenever required. However, in order to receive, appreciate, and use the information in a proper way, one must have

- adequate knowledge,
- adequate motivation, and
- adequate resources and deciding power.

Knowledge about accidental explosions can be acquired by reading, listening to lectures, talking to experts etc., although experience from actual

explosion prevention and mitigation work is perhaps the best form of knowledge.

Genuine motivation is perhaps more difficult to achieve indirectly. It seems to be a law of life that people who have themselves experienced serious explosion accidents possess the highest level of motivation, in particular if the accident caused injuries and perhaps even loss of life. This applies to workers as well as top management. However, high levels of motivation can also result from good demonstrations of real explosions, including their initiation by various ignition sources, as well as their propagation and damaging effects. Video and film can also be a good help, if used properly.

The final element, adequate resources and the authority to put the good plans into practice, is in reality controlled by the top management. Verhaegen (1989) concluded from this that the real responsibility for establishing and running a proper safety assurance system must always lie on the top management. Summarizing the experience of a large, multinational chemical company, Verhaegen suggested that the following ten essential elements be involved to ensure proper safety management:

- top management responsibility
- safety statement (explicit commitment from top management)
- objectives and goals (specification of long and short term expectations)
- stated standards (written guidelines and rules)
- safety committees (a dedicated organization for handling safety issues at all levels)
- safety audits (regular reexamination of work practices)
- accident records (written analyses of accidents. Why did they happen? How can similar future accidents be prevented?)
- safety personnel (qualified specialists essential as advisers, but responsibility remains with top management)
- motivation (by information and involvement, etc.)
- training (a continual process; courses essential; the message must get through!)

Verhaegen emphasized the problem that a good safety organization is in reality often kept active by one or two dedicated individuals. If they change position within the company, or even leave, the safety organization may suffer. Management should foresee this problem and provide a workable solution.

Chapter 2

Gas and Vapor Cloud Explosions

2.1 Combustion of Gases and Vapors

2.1.1 Diffusion Combustion and "Premixed" Combustion

Burning of combustible gases is well known from daily life. Propane (C_3H_8), for example, is often used for lighting, heating, and cooking where electricity is not easily available. When discussing the burning of combustible gases, it is important to distinguish between diffusion controlled burning and burning of premixed gas/air. Figure 2–1 illustrates the flame of a common gas lighter with butane (C_4H_{10}) as the fuel. The burning is limited to a narrow zone just outside the butane outlet, where the gas makes contact and mixes with the ambient air, and hence with oxygen. The burning occurs comparatively slowly, being controlled by the rate of diffusion of fuel and oxygen molecules into the reaction zone. The same type of burning takes place in a candle flame and in the flame of an oil lamp.

In a Bunsen burner, as illustrated in Figure 2–2, the situation is different. In this case, the combustible gas is first mixed well with air before becoming ignited. In the Bunsen burner, it is important that the mass flows of propane and air are adjusted to yield a propane concentration within the flammable range. Furthermore, the overall mass flow of the mixture must be below certain limits. If the flow is too high, the flame will be blown away upwards and extinguish; if it is too low, the flame will propagate downwards into the interior of the burner.

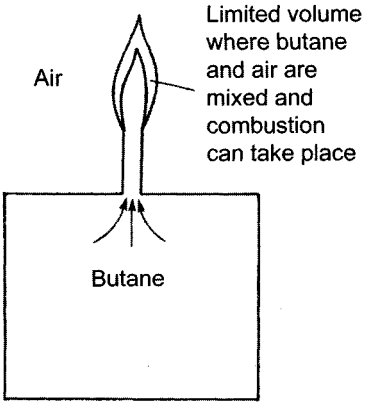


Figure 2–1 Butane lighter flame. Example of diffusion controlled burning of a combustible gas in air.

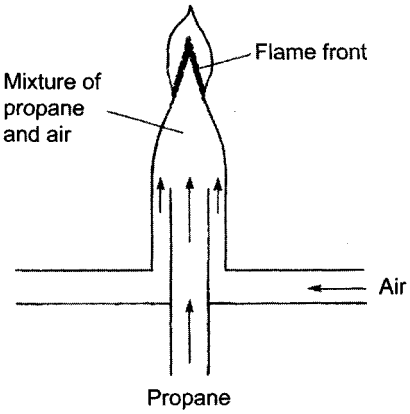


Figure 2–2 Burning of premixed propane/air in a Bunsen burner.

2.1.2 Laminar Burning of Premixed Gas/Vapor and Air

Figure 2–3 gives experimentally determined relationships between the so-called laminar burning velocity and the concentration of combustible gas in the mixture with air for four common combustible gases.

The ideal laminar burning velocity is the lowest velocity at which a flame front can propagate through a given gas mixture (given ratio of gas to air,

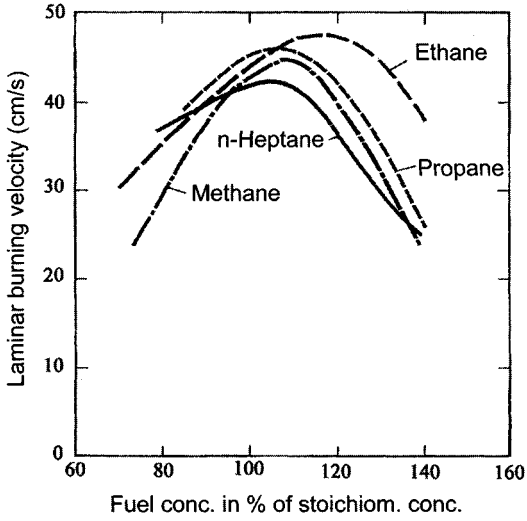


Figure 2–3 Laminar burning velocities S_u at atmospheric pressure and normal temperature for mixtures of various hydrocarbon gases and air. From Zabetakis (1965) p. 45.

given pressure and temperature). The flame zone is then a smooth, plane sheet of thickness of the order of 1 mm, traveling at constant velocity through the quiescent homogeneous gas/air mixture. The heat transfer mechanism driving the flame propagation is mainly heat conduction.

As Figure 2–3 shows, the differences between the laminar burning velocity for the various alkanes (saturated linear hydrocarbons) is relatively small, all four maximum values falling in the range 0.42–0.47 m/s. However, as illustrated by Figure 2–4, unsaturated hydrocarbons can have considerably higher maximum values.

Both Figure 2–3 and Figure 2–4 illustrate that for any combustible gas, S_u has a maximum value when the mixing ratio of fuel to air is close to the stoichiometric ratio. At the stoichiometric ratio, the ratio of fuel to oxygen is exactly what is required for all the oxygen to transform all the fuel to CO_2 and H_2O. When the fuel/air ratio moves away from the stoichiometric value, whether in the direction of leaner or richer mixtures, the value of S_u becomes systematically smaller.

Table 2–1 provides maximum experimental S_u values for some combustible gases in air.

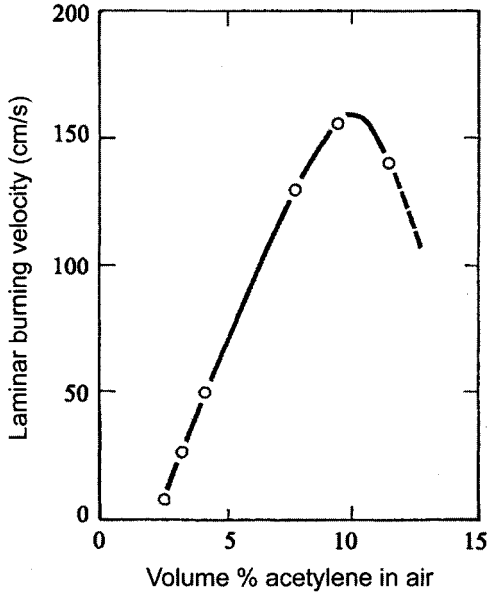


Figure 2–4 Laminar burning velocities S_u for different mixing ratios of acetylene and air, at atmospheric pressure and normal temperature. From Zabetakis (1965) p. 59.

2.1.3 Flammable Concentration Ranges for Premixed Gas/Vapor and Air

2.1.3.1 General

As the fuel to air ratio is increased or decreased sufficiently, departing from the stoichiometric value, two finite limits are eventually reached beyond which the mixture is no longer able to propagate a flame. These limits bracket the flammable (or explosive) range of a given combustible gas or vapor in air. Figure 2–5 illustrates the flammability ranges in air for some gases. Because acetylene, being a very reactive compound, is able to decompose exothermically in the absence of oxygen, its upper explosion limit is in fact 100 percent. The first two columns of Table 2–2 give numerical values for the lower and upper flammability limits of a number of combustible gases and vapors in air, including those in Figure 2–5.

Table 2-1 Maximum Laminar Burning Velocities S_u for Premixed Fuel/Air at Atmospheric Pressure and Normal Temperature for Some Fuels[a]

Fuel	Maximum S_u [cm/s]
Alkanes	40-50
Natural gas (unspecified)	40
Acetone	43
Butylene	43
Cyclohexane	44
Bensene	48
Propylene	51
Carbon monoxide	52
Methanol	57
Propylene oxide	67
Ethylene	75
Ethylene oxide	90
Acetylene	155
Hydrogen	325

a. Data from J.M. Kuchta, *Investigation of fire and explosion accidents in the chemical, mining and fuel related industries. A Manual.* US Bureau of Mines, Bulletin 680 (1985): 38.

It should be pointed out that the experimental determination of flammability limits is not straightforward, and results obtained using different methods may scatter. The method used in the United States is described in ASTM (2003a). The fuel/air mixture to be tested is placed in a five liter spherical glass flask with the spark gap providing the electric spark for ignition located close to the flask center. The flask is filled with a uniform mixture of the desired fuel to air ratio, followed by activation of the ignition source and visual observation of the extent of flame propagation. Flame propagation both upwards and sideways is regarded as sufficient for the mixture to be regarded as flammable, whereas flame propagation upwards only, is not regarded as sufficient. Downwards flame propagation is not regarded as a necessary requirement for the mixture to be classified as flammable. In a more recent comment, Ural (2004) points out that the Lower and Upper Flammability Limits (LFL, UFL) observed in the ASTM test are not necessarily identical with Lower and Upper Explosibility Limits (LEL, UEL) determined by other methods.

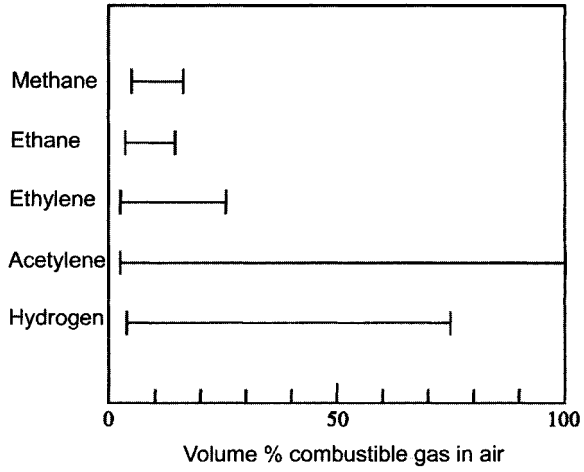


Figure 2–5 Flammable ranges for mixtures of some combustible gases and air at atmospheric pressure and normal temperature. (See also Table 2–2.)

The other data in Table 2–2 will be considered in the discussion of the respective parameters that follows.

2.1.3.2 Flash Point of a Combustible Liquid

The flash point of a combustible liquid, as defined by The Institute of Petroleum (2002), is the lowest temperature at which, in contact with air at atmospheric pressure, a flammable liquid will give off a sufficiently rich vapor for ignition to occur in the presence of an ignition source. On the condition that the ignition source is sufficiently effective, the flash point is the lowest temperature the liquid can have for the vapor/air mixture just above the liquid surface to be able to propagate a flame. There is then a direct coupling between the lower flammability limit of vapors and the flash point of their liquid phases. This is because the flash point is the temperature at which the vapor pressure of the liquid equals the partial pressure of the actual vapor at its lower flammability limit.

Flammable liquids are grouped in hazard classes on the basis of their flash points. The Institute of Petroleum (IP) classification system uses a subdivision of the main Classes II and III according to whether the condition of handling is at a temperature below or above the flash point. This is in

Table 2-2 Combustibility and Ignitability Parameters of Some Combustible Gases and Vapors in Air at Atmospheric Pressure and Normal Temperature[a]

Fuel	Flammable limits [vol. % in air]		Flash point	Min. ign. temp.	MESG	Temp. class	Gass group code
	Lower	Upper	[°C]	[°C]	[mm]		
Acetone	2.6	13.0	-19	735	1.01	Tl	IIA
Acetylene	2.5	100.0	?	305	1.01	T2	IIC
Ammonia	15.0	28.0	?	630	3.18	T1	IIA
Bensene	1.3	7.9	-11	560	0.99	T1	IIA
Butadiene	2.0	12.0	-85	430	0.79	T1	IIB
Butane	1.8	8.4	-60	370	0.98	T2	IIA
Butylalcohol	1.7	12.0	29	359	0.91	T2	IIA
Cyclohexane	1.3	7.8	-18	259	0.94	T3	IIA
1-Butene	1.6	10.0	-80	455	0.94	T1	IIA
Dekane	0.3	5.6	46	201	1.05	T3	IIA
Dietylether	1.9	36.0	-45	160	0.87	T4	IIB
Dimetylether	3.4	27.0	-42	350	0.84	T2	IIB
Etane	3.0	12.4	?	515	0.91	T1	IIA
Ethanol	3.3	19.0	12	363	0.91	T2	IIA
Ethylene	2.7	36.0	?	425	0.65	T2	IIB
Heptane	1.1	6.7	- 4	215	0.91	T3	IIA
Hexane	1.2	7.4	-21	233	0.93	T3	IIA
Hydrogen	4.0	75.0	?	560	0.28	T1	IIC
Hydrogen sulphide	4.0	44.0	?	270	0.90	T3	IIA
Carbon disulphide	1.3	50.0	-30	102	0.20	T5	IIC
Carbon monoxide	12.5	74.5	?	605	0.84	T1	IIB
Metane	5.0	15.0	?	540	1.14	Tl	I/IIA*
Nafta	0.0	6.0	<-18	290	?	T3	IIA
Propane	2.1	9.5	-104	493	0.92	T1	IIA
Styrene	1.1	6.1	30	490	?	T1	IIA
Toluene	1.2	7.1	4	535	?	T1	IIA

* For methane Code I applies to coal mines only, whereas Code IIA applies to all other situations.

a. Data from International Electrotechnical Commission *Data for flammable gases and vapours relating to the use of electrical apparatus.* Committee draft 31 (SEC.) 16Y' made for IEC TC31 by the working group WG4. Central Office of International Electrotechnical Commisssion, Geneva, Switzerland. (1992)

agreement with the conception that an explosive gas (vapor) atmosphere generally cannot exist if the flash point is significantly above the maximum temperature of that fluid.

However, as will be discussed in Chapter 3, if a pressurized combustible liquid at a temperature below its flash points is released to the atmosphere as a spray of small droplets in air, the spray cloud can become ignited and propagate a flame just like a pre-mixed homogeneous explosive gas cloud.

The third column of Table 2–2 gives flash points for some combustible liquids in air. The other four parameters included in Table 2–2 will be discussed later in this chapter in their relevant contexts.

2.1.3.3 Classification of Flammable Fluids According to Their Flash Points

The classification system for flammable fluids recommended by The Institute of Petroleum (2002) is given in Table 2–3. The main classification conforms to the flash point classification systems of the European Directives. The subdivision of Classes II and III has been standardized by The Institute of Petroleum (2002) to aid the handling of petroleum materials throughout bulk storage, loading, conveyance, discharge, and distribution. It should, however, be recognized that there is an upper temperature limit, though unlikely to be encountered, in Classes I, II(2), and III(2) classification applications above which they should not be extended; this upper limit is the temperature at which the liquid would boil.

Table 2–3 Classification of Flammable Fluids According to Their Flash Points[a]

Fluid class	Description
Class 0	Liquefied petroleum gases (LPG)
Class I	Liquids that have flash points below 21 °C
Class II(1)	Liquids that have flash points from 21 °C up to and including 55 °C, handled below flash point
Class II(2)	Liquids that have flash points from 21 °C up to and including 55 °C, handled at or above flash point
Class III(1)	Liquids that have flash points above 55 °C up to and including 100 °C, handled below flash point
Class III(2)	Liquids that have flash points above 55 °C up to and including 100 °C, handled at or above flash point
Unclassified	Liquids that have flash points above 100 °C

a. *Source*: Data from The Institute of Petroleum *Area classification code for installations handling flammable fluids*, 2d ed. August 2002, The Institute of Petroleum, London.

Typical examples of petroleum fluids are shown in Table 2–4.

Where the flammable fluid is a liquid, its volatility in relation to the conditions of temperature and pressure at which a potential release might take place is an essential factor, since it will determine the extent of rapid vapor formation from that release. In many of the commonly encountered

Table 2–4 Classification of Commonly Encountered Combustible Petroleum Materials[a]

Class of petroleum based on flash point	Examples of petroleum materials	Flash point (°C)	Boiling range (°C)
Class 0	LPGs, ethylene, propylene	Not applicable	Propane - 42 Butane -1 Isobutane -12
Class I			
<21 °C	Gasoline (petrol)	-45	ca. 20-205
	Stabilised crude oil	<21	ca. - 1 to 380+
	Avtage wide cut jet fuel (JP4; Jet B)	-25	ca. 0-220
	Benzene	-11	80
	Toluene	4	110
	Naphtha	-2 to 10	30-177
	Methanol	11	65
Class II			
21-55 °C	Avtur/Jet A Turbofuel Kerosine	38 (min.)	150-240
	(a) premium grade	43 (min.)	160-280
	(b) regular	38 (min.)	150-280
Class III			
55-100 °C	Gas oil/distillate heating oil	55+	250-360
	Automotive diesel fuel	55+	180-360
Unclassified	Atmosphere residues		
>100 °C	Heavy fuel oils	>100	>350

a. *Source*: Data from The Institute of Petroleum *Area classification code for installations handling flammable fluids*, 2d ed. August 2002, The Institute of Petroleum, London.

types of process plant, the flammable fluid can be classified satisfactorily by its flash point, e.g. in accordance with the procedure of The Institute of Petroleum (2002). Further sub-division of the main classes (II and III) into sub-classes (1) and (2) is done according to whether the liquids are handled at temperatures above or below the flash point.

2.1.4 Maximum Pressures Generated from Constant-Volume Adiabatic Combustion of Premixed Gas/Vapor and Air

If the combustion occurs at constant volume, a pressure rise will result. The adiabatic temperature rise at constant volume can be expressed as

$$T_2 - T_1 = \Delta E/C_v \tag{2.1}$$

where ΔE is the liberated combustion heat per mole at constant volume, C_v is the mean specific heat of the gaseous system at constant volume, T_1 is the

initial gas mixture temperature (prior to combustion), and T_2 is the final temperature of the combustion products and the nitrogen (from the air) after complete combustion. Ideal T_2 values computed using Equation (2.1) are most often somewhat higher than experimental values, due to inevitable heat losses and non-ideal chemical conversion in real experiments. However, assuming that an adequate T_2 value can be obtained (see Table 2–6), the corresponding constant-volume adiabatic pressure rise can be estimated, using the equation of state (assuming ideal gases)

$$P_2/P_1 = n_2T_2/n_1T_1 \tag{2.2}$$

where P is pressure (absolute), T is temperature in Kelvins and n is total number of moles of gas in the constant volume considered. A more detailed analysis is given by Kuchta (1985). Table 2–5 and Table 2–6 give some experimental P_2 values. The data in the two tables originate from different sources and deviate somewhat. Different fuel to air ratios in the experiments may be one reason for the discrepancies.

Note that initial temperature T_1 and pressure P_1, deviating from normal atmospheric conditions, will give correspondingly different P_2 values. As

Table 2–5 Highest Maximum Explosion Pressure (Gauge) for Combustion of Premixed Fuel Gas/Air in a Closed 5 Liter Test Vessel[a] [b]

Fuel	bar(g)	Fuel	bar(g)
Acetylene	10.3	i-Butylalcohol	7.6
Ethylene oxide	9.9	Ethyl alcohol	7.5
Bensene	9.0	n-Butylalcohol	7.5
Acetone	8.9	n-Decane	7.5
Ethylene	8.9	Hydrogen	7.4
n-Hexane	8.7	Carbonmonoxide	7.3
n-Pentane	8.7	Methane	7.2
n-Butane	8.6	Toluene	6.8
Cyclohexane	8.6	Vinylchloride	6.8
n-Heptane	8.6	Metylbutadiene	6.6
Propane	8.6	Styrene	6.6
Propylene	8.6	Ammonia	6.0
i-Octane	8.1	Acetic acid	5.4
Carbon disulphide	7.8	Hydrogen sulphide	5.4

a. The fuel gas/air ratio was varied until the highest explosion pressure had been found for each fuel. Normal atmospheric initial conditions.

b. *Source:* Data from H.H. Freytag, *Handbuch der Raumexplosionen,* Verlag Chemie, Weinheim, Germany. (1965): 152.

Table 2–6 Combustion Properties of Some Fuel Gas/Air Mixtures. Normal Atmospheric Initial Conditions[a]

Fuel	Vol.% fuel in air	Constant volume	Constant pressure	Adiabatic flame temperature	
		Max. pressure [bar(g)]	Expansion ratio E [-]	Constant volume [K]	Constant pressure [K]
Acetylene	7.75	8.80	8.38	2925	2542
Hydrogen	29.60	7.02	6.88	2755	2384
Ethylene	6.53	8.32	7.48	2740	2372
Butane	3.13	8.43	8.06	2639	2272
Propane	4.0	8.34	7.98	2633	2266
Ethane	5.67	8.21	7.87	2627	2261
Methane	9.51	7.60	7.25	2591	2227

a. *Source*: Data from J.H.S. Lee, *Physics of explosions*, lecture notes, MeGill University, Montreal. (1984): 249.

a rule, the maximum adiabatic explosion pressure (absolute) produced by a given fuel/air mixture at constant volume, is proportional to the initial pressure (absolute) prior to combustion. Hence, a gas mixture giving 8 bar (absolute) at atmospheric initial pressure, will yield 16 bar (absolute) at 2 bar (absolute) initial pressure (1 bar overpressure).

In the case of premixed gas explosions in a system of closed volumes in series, coupled via comparatively narrow ducts or pipes, an explosion in one part of the system can give rise to considerable pressure rise in the still unburned gas mixture in other parts (pressure piling). When the flame front reaches and ignites such precompressed gas volumes, very high pressure transients can result, in spite of the fact that the entire system was at atmospheric pressure prior to the first ignition.

2.1.5 The "Expansion Ratio" for Combustion of Premixed Gas/Vapor and Air

Figure 2–6 illustrates idealized adiabatic (no heat loss to the tube, no buoyancy, no interference of the wall with the gas flow) planar, laminar combustion of premixed quiescent explosive gas/air in a one-end open straight tube/duct, at constant pressure.

If the gas mixture is ignited in a plane across the open end of the tube (Figure 2–6a), the combustion products will expand freely into the ambient atmosphere, whereas the still unburned gas further into the tube will remain

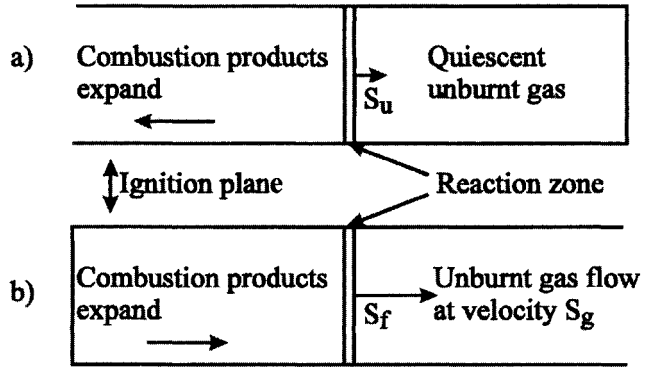


Figure 2–6 Premixed fuel gas/air burning at constant pressure with a plane, laminar flame in a one-end-open tube. a) ignition at open tube end. b) ignition at closed tube end.

quiescent. The observed flame speed in relation to the tube wall will then be identical with the laminar burning velocity S_u of the actual gas/air mixture. If ignition occurs at the closed tube end, however, (Figure 2–6b) the expansion of the combustion products is forced to occur in the same direction as that of flame propagation. Therefore the unburned gas mixture ahead of the flame will be pushed towards the open tube end. In this case the observed flame speed S_f in relation to the tube wall is the sum of S_u and the flow velocity S_g of the unburned gas. Hence, Equation (2.3) and Equation (2.4) describe how the liberated combustion heat, in the case of constant pressure, gives rise to expansion.

$$S_f = S_u + S_g \tag{2.3}$$

or

$$S_f = S_u \cdot E \tag{2.4}$$

Here E is the "expansion ratio", simply defined by Equation (2.4) as S_f/S_u, or by using Equation (2.3), as $(1 + S_g/S_u)$.

In other words, E is the ratio of the volume that a given quantity of premixed gas has attained after complete, ideal adiabatic combustion at constant pressure, and its volume before combustion. For a given fuel, E

varies with the fuel/air ratio and has its maximum around the stoichiometric ratio or somewhat higher. Maximum E values for most hydrocarbons in air are in the range 7.5–8.0. Some values are given in Table 2–6. The reason for this substantial expansion is the large amount of heat liberated when the fuel and oxygen combine and form combustion products (CO_2 and H_2O for complete combustion of hydrocarbons). Note that the term *burning velocity* is by definition the relative linear velocity S_u by which the combustion reaction is "eating" itself into the unburned gas/air mixture, whereas *flame speed* denotes the linear velocity S_f of the flame front in relation to a stagnant observer or structure or apparatus. If the combustion occurs under conditions where expansion cannot occur freely, the liberated heat may give rise to a substantial pressure rise, which, in turn, may cause substantial damage to industrial equipment, plants, buildings etc. (see Table 2–5 and Table 2–6).

Figure 2–7 illustrates ideal laminar spherical combustion of a premixed gas/air cloud, following ignition at a point in the cloud. In this case, assuming a very thin flame and negligible buoyancy, the unburned gas/air will always be pushed in the direction of flame propagation, and Equation (2.3) and Equation (2.4), with the same numerical values as for the linear case in Figure 2–6b, apply even in this case.

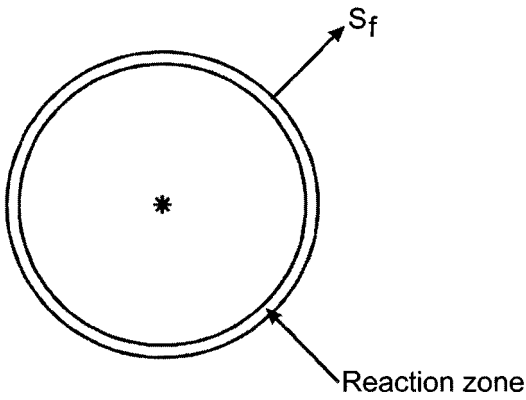


Figure 2–7 Ideal laminar spherical burning of quiescent premixed fuel gas/air, following ignition at a point.

2.1.6 Turbulent Combustion of Premixed Gas/Vapor and Air

In real premixed accidental explosions of gas clouds in an industrial plant, the cloud will not be quiescent but turbulent. Flame propagation in turbulent gas is illustrated in Figure 2–8.

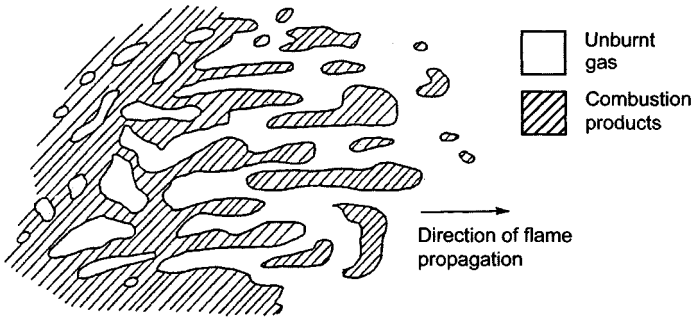


Figure 2–8 Illustration of flame front structure in turbulent premixed gas.

In this case the flame front is no longer a continuous, plane sheet, but folded and torn, with an effective surface area that can be substantially larger than the nominal projected area. This in turn gives rise to nominal turbulent burning velocities that can be considerably higher than the laminar value for the gas mixture in question. Figure 2–9 gives some results from measurements in a burner similar to the one illustrated in Figure 2–2.

When discussing accidental gas explosions one distinguishes between two kinds of turbulence, initial or pre-explosion turbulence, and turbulence generated by the explosion itself. The latter is caused by the transient flow of the unburned gas ahead of the flame, generated by the expansion of the combustion products behind the flame. The production of this kind of turbulence is particularly exaggerated in areas packed with process equipment, pipes and other kinds of obstructions to the flow, as illustrated in Figure 2–69.

Figure 2–10 illustrates the dramatic effects that can result from gas explosions in obstructed geometries.

These experiments were performed in a one-end-open tube of diameter 2.5 m, length 10 m and volume 50 m^3. Ring-shaped turbulence-generating steel baffles could be mounted at various positions in the tube. Because the tube was fully open at one end, the rate of combustion had to be very

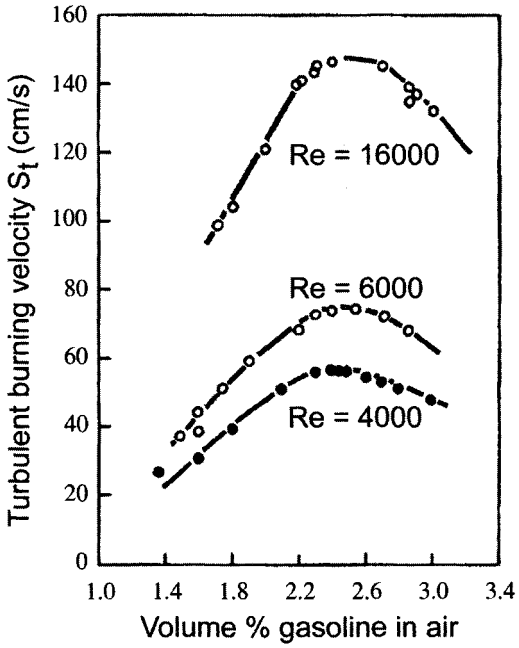
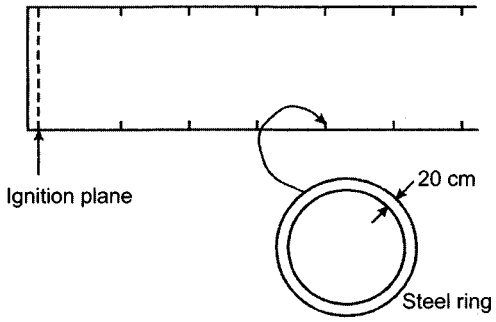


Figure 2–9 Turbulent burning velocity as a function of Reynolds number Re. Premixed gasoline/air in steady-state burner of diameter 16 mm. From Freytag (1965) p. 70.

high for high explosion overpressures to be generated inside the tube. As the table in Figure 2–10 shows, the influence of introducing the flow obstructing rings on the maximum explosion overpressure in the tube was dramatic, no rings giving 0.15 bar(g), 1 ring 0.6 bar(g), 3 rings 3.5 bar(g), and 6 rings 8.0 bar(g). In other words, the variations in overpressure were by more than a factor of 50! These experiments urged the development of the computer code FLACS that can predict the explosion overpressures to be expected from a given fuel gas/air system, as a function of geometry of the explosion environment, and the location of the ignition point (see Section 2.4.5.7).

2.1.7 Detonation of Premixed Gas/Vapor and Air

Detonation is an extreme singular mode of flame propagation in premixed gas. In this mode the combustion zone follows immediately behind, and



Max explosion pressures in tube for 9.5 vol. % methane in air	
Ring configuration	[bar(g)]
No ring in tube	0.15
1 ring 2 m from ign.plane	0.60
3 rings	3.50
6 rings (as shown in fig.)	8.00

Figure 2–10 Experiment performed in Norway to determine the influence of explosion-generated turbulence on the violence of gas explosions, using turbulence-generating rings (baffles). From Moen et al. (1982).

drives, a leading shock front moving into the quiescent, unburned gas mixture at supersonic velocity. The mechanism of flame propagation is then no longer thermal diffusion, but extremely rapid compression, whereby the temperature in the compression zone is raised to such a high level that ignition occurs almost instantaneously. The necessary condition for self-sustained detonation propagation is that the shock wave is sufficiently strong for the volume inside it to become ignited and react chemically before the shock wave has traveled a significant distance away. In this way the shock wave and the chemical reaction zone remain closely coupled, and the shock wave speed and strength is maintained. As shown in Table 2–7, typical detonation velocities in premixed hydrocarbon gas/air, at normal initial pressure and temperature and optimum fuel concentrations, are 1800–1900 m/s, i.e. of the order of five times the velocity of sound in the unburned, uncompressed premixed gas/air. For this reason, the unburned mixture ahead of the detonation front cannot receive any gas dynamic signal from the approaching detonation front until being caught

Table 2–7 Detonation Properties of Some Fuel Gases Premixed with Air at Atmospheric Pressure and Temperature—Stoichiometric Mixtures[a]

Fuel	Detonation velocity [m/s]	Overpressure in detonation front [bar(g)]
Methane	1800	16.1
Ethane	1800	17.0
Propane	1800	17.3
Butane	1800	17.4
Ethylene	1860	17.7
Acetylene	1870	18.0
Hydrogen	1960	14.6

a. *Source*: Data from J.H.S Lee, *Physics of explosions.* Lecture notes, MeGill University, Montreal. (1984): 249.

by the front itself. Therefore, reducing the maximum explosion pressure of a detonation by explosion venting is impossible.

Detonations can be initiated directly by means of a sufficiently strong explosive charge or electric spark. In accidental explosions in industry, Deflagration-to-Detonation Transition (DDT) is the common initiation mechanism. This can occur in long ducts, when turbulence and burning rates increase to such high levels that the compressed unburned gas, trapped between the flame front and the leading shock, ignites spontaneously. Extremely high transient pressure peaks can then be generated before stable detonation conditions are attained.

In addition to typical detonation front velocities, Table 2–7 also gives corresponding detonation peak pressures. Hydrogen deviates from the hydrocarbons because the density of 30% hydrogen in air is only 18/29 of the density of air, and the velocity of sound therefore 418 m/s as opposed to 330 m/s (C_1 in Equation (2.5)).

Figure 2–11 shows the result of an experiment in which stoichiometric premixed acetylene/air detonated in a 46 m long tube of diameter 0.6 m.

Some simple relations are useful. The first is

$$P_{CJ}/P_1 = [1 + \gamma_1(V^*/C_1)^2]/[1 + \gamma_2] \qquad \textbf{(2.5)}$$

where P_{CJ} is the Chapman-Jouguet detonation front pressure, P_1 the initial pre-ignition pressure, γ_1 and γ_2 the C_p/C_v ratio (specific heats at constant pressure and volume) for the pre-ignition gas mixture and the combustion

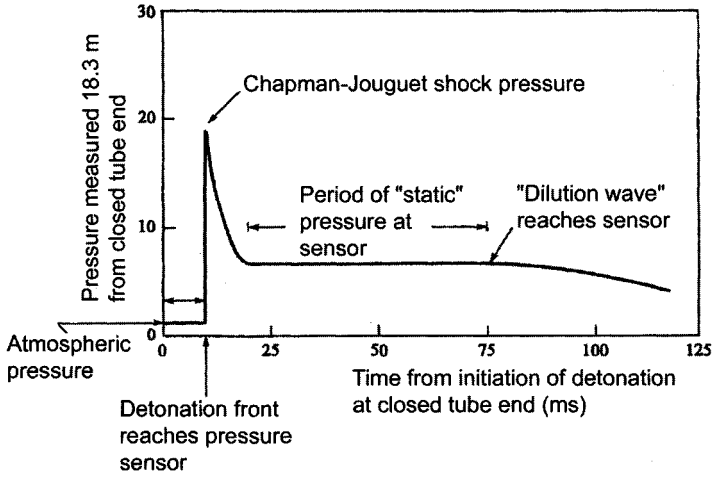


Figure 2–11 Detonation of a stoichiometric mixture of acetylene and air in a one-end-open tube of diameter 0.6 m. The detonation was initiated at the closed tube end by means of an explosive charge. The pressure sensor was located 18.3 m from the closed tube end. From Kuchta (1985) p. 18.

products respectively, V* the detonation front velocity in the unburned mixture, and C_1 the velocity of sound in the pre-ignition mixture. V^*/C_1 is the Mach-number, which is >>1 for detonation fronts in hydrocarbon/air mixtures.

The following rule of thumb gives useful guidance:

$$P_{CJ} \approx 2P_2 \tag{2.6}$$

where P_2 is the maximum adiabatic explosion pressure at constant volume (Table 2–5 and Table 2–6).

For any given ideal gas the velocity of sound C is independent of pressure, but dependent on temperature. The following relation applies:

$$C = [(\gamma P_o T)/(273\ \rho_o)]^2 \tag{2.7}$$

where γ is C_p/C_v ($\approx$ 1.4 for 2-atomic gases), P_o and ρ_o the gas pressure and density at the same reference temperature, and T is the absolute temperature of the gas. In other words, at any given T, the velocity of sound C is inversely proportional to $M^{0.5}$, where M is the mole weight of the gas.

For this reason, C at any given temperature is considerably higher in hydrogen than in air.

Figure 2–12 illustrates the dramatic span (logarithmic scale) of possible burning velocities of a given fuel/air mixture, using propane/air as an example. The concentration ranges for detonation are somewhat narrower than for turbulent combustion, which are in turn somewhat narrower than that for laminar burning. Table 2–8 gives some figures illustrating this.

Table 2–8 Comparison of Composition Limits for Laminar Burning and Detonation for Two Combustible Gases Mixed with Air[a]

Fuel	Limits for laminar flames [vol.% in air]		Detonability limits* [vol.% in air]	
	Lower	Upper	Lower	Upper
Acetylene	2.5	100	4.2	50
Hydrogen	4.0	75	18.3	59

a. *Source*: Data from Table 2–2 and H. H. Freytag, *Handbuch der Raumexplosionen.* Verlag Chemie, Weinheim, Germany. (1965): 74.

2.2 Ignition of Premixed Gas/Vapor and Air

2.2.1 Introduction

This section concentrates mainly on initial cloud conditions of normal atmospheric pressure and temperature, and on ignition of pre-mixed hydrocarbon gases and air. Experimental evidence elucidating the basic features of various ignition processes and practical guidelines for prevention of ignition in industrial process plants are considered.

The phrase *ignition source* is used as a general term embracing all categories of heat source that may, in principle, give rise to ignition. The phrase *effective ignition source* indicates that a particular source will actually cause ignition if brought into contact with the specific explosive atmosphere of concern.

Numerous national and international standards and guidelines are currently in place for preventing accidental ignition of explosive atmospheres. In Europe, a comprehensive standard was drafted by CEN

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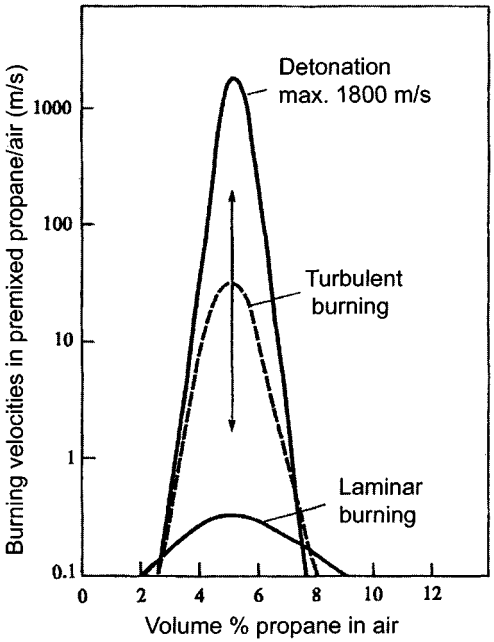


Figure 2–12 Burning velocities in premixed propane/air at atmospheric pressure and normal temperature, for different combustion modes (laminar, turbulent, detonation).

(1993). The document also addresses ignition prevention, and in this context refers to and builds on a series of European and fully international standards on various specific aspects of the ignition prevention problem. The guidelines for industrial practice given in the various sections that follow are mainly based on this document. In addition to Eckhoff (1996), a main source for this section is the paper by Eckhoff and Thomassen (1994).

2.2.2 What Is Ignition? The Basic Theory of "Thermal Runaway"

Explosive gas mixtures can be ignited by a variety of ignition sources including:

- open flames (matches, welding and soldering flames etc.)
- glowing or smoldering materials

- hot solid surfaces
- burning metal particles and "thermite" flashes from impacts, grinding etc.
- electrical and electrostatic sparks, arcs, and other discharge forms
- jets of hot combustion gases
- adiabatic compression
- light radiation, e.g. light conveyed through optical fibers or cables

The question then arises whether ignition by all these quite different sources can be described qualitatively by one common global conceptual model. The answer is yes, and the model is the classical thermal explosion theory formulated by Frank-Kamenetskii (1955). In the following the essence of this theory will be outlined qualitatively with reference to Figure 2–13. In principle the theory applies to any combustible system exposed to a potential ignition source or process. In the context of the present book it applies with equal validity to explosive clouds of combustible gases, liquid sprays/mists, or dusts, and to combustible dust layers/deposits.

Consider an explosive gas mixture, e.g. 4 vol.% propane in air, initially at room temperature. Then consider a small "ignition volume" inside the bulk of this mixture. Assume that the ignition volume can be heated to any desired temperature by a small heating source, e.g. an electrically heated resistance wire that has been placed inside it. Then make the simplifying assumptions that the temperature throughout the ignition volume is uniform at any time during the heating-up process, with a very sharp temperature drop to ambient temperature at the boundary, and that the ignition volume does not expand during heating. Figure 2–13 then illustrates the increase of the heat generation rate inside the ignition volume due to the chemical reaction of the gas mixture there, as well as the increase of the conductive heat loss rate from this volume to the ambient atmosphere, with increasing temperature in the ignition volume.

First, consider the solid lines representing an ignition volume V. The curved line then represents the rate of heat generation G(T) within the ignition volume due to the exothermal reaction between the fuel and the air, as a function of the temperature in this volume. According to the classical Arrhenius theory, this relationship is exponential for a zero-order chemical reaction. The straight solid line represents the rate of heat loss L(T), which increases linearly with the temperature drop between the ignition volume and the ambient gas. Figure 2–13 illustrates that at modest temperatures T the rates of heat generation by combustion are normally substantially

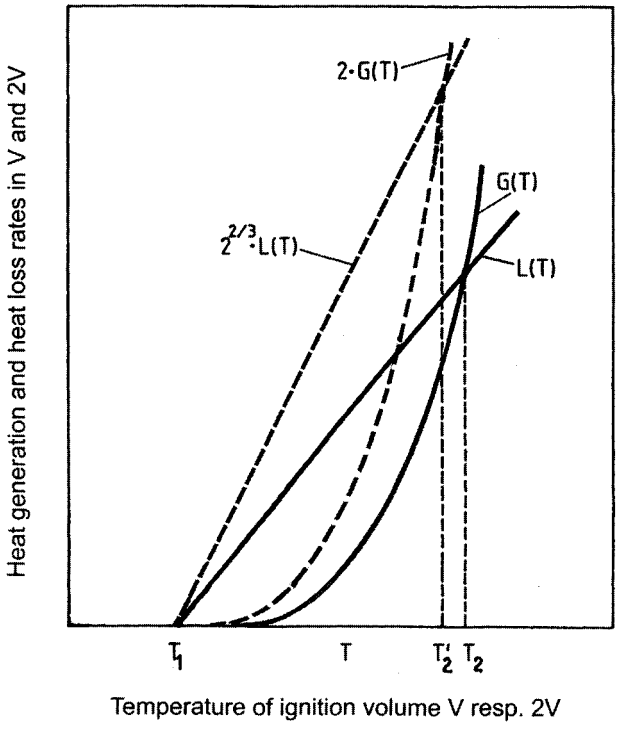


Figure 2–13 Simplified schematic illustration of the basic idea of the "thermal explosion theory" by Frank-Kamenetskii (1955).

lower than the heat loss rates, and it is thus impossible for the temperature in the ignition volume to rise above the ambient temperature T_1 by the slow heat-producing chemical reaction only.

However, if T is raised by activating the external heating source, G(T) will start to raise exponentially. If the external heating is brought to an end while the temperature of the ignition volume is still below the critical temperature for ignition T_2, the rate of heat loss L(T) will still exceed the heat generation rate by chemical reaction G(T), and the temperature in the ignition volume will drop back to ambient temperature again. If, however, T is raised further to a point where G(T) exceeds L(T), a positive feed back loop is established by which further temperature rise is accomplished by the combustion reaction itself. This process will eventually lead to ignition. The critical temperature T_2 for ignition is the one at which:

$$G(T) = L(T) \tag{2.8}$$

and

$$dG(T)/dT > dL(T)/dT \tag{2.9}$$

Now consider the dotted lines in Figure 2–13, representing an ignition volume of 2V. In this case the value of G(T) will be twice that for the ignition volume V for any value of T, simply because the amount of reactive mixture has been doubled. However, because the heat loss rate from the ignition volume increases proportionally with the surface area of this volume, and not with the volume itself, L(T) at any T has only increased by a factor of $2^{2/3}$ compared with the factor of 2 for the ignition volume V. Because of this, the critical temperature T_2' at the crossing point between G(T) and L(T), will be lower in the case of 2V than in the case of V.

This indicates that the minimum temperature that an ignition source must have to cause ignition will decrease with increasing size of the ignition volume to be associated with that source. In reality the shape of the ignition volume will also play an important role.

2.2.3 Ignition by Open Flames and Hot Gases

Open flames are gaseous combustion reactions at temperatures of at least 1000°C. Flames and their hot gaseous reaction products, even at very small volumes, are among the most effective ignition sources for explosive gas clouds.

2.2.4 Ignition by Hot Surfaces

2.2.4.1 Overview

The minimum hot surface temperature for igniting a given mixture of combustible gas in air has sometimes been regarded as a fundamental constant for that mixture. However, this is a false perception. In general terms, ignition is a dynamic process where chemical heat generation and physical heat loss compete in a complex manner in the potential ignition region, and where the former eventually overtakes the latter. This also applies to hot surface ignition.

The size of the hot surface and the relative movement between the explosive gas mixture and the hot surface are two key parameters controlling the minimum ignition temperature, T_{min} of a given gas mixture. Classical investigations of these effects are illustrated in Figure 2–14 and Figure 2–15. Both these experiments should be compulsory background in any study of hot surface ignition processes. The results by Silver (1937) and Paterson (1939, 1940) are shown in Figure 2–14, from which two systematic trends can be extracted

- T_{min} decreases with increasing sphere diameter
- T_{min} decreases with decreasing sphere velocity

There was no significant difference between spheres of quartz and aged platinum.

A similar set of classical results obtained by Mullen et al. (1949) is shown in Figure 2–15.

In this case, an explosive pentane/air mixture was flowing past a stationary hot metal rod at comparatively high velocities. The minimum rod temperature for ignition was recorded as a function of rod diameter and gas velocity. Figure 2–15 reveals the same main trends as Figure 2–14. T_{min} decreases systematically with increasing size of the hot surface and with decreasing relative velocity between the hot surface and the gas. It should be noted that the range of relative velocities is considerably higher in Figure 2–15 than in Figure 2–14, by a factor of about 10.

This strong dependence of T_{min} on actual experimental circumstances is exposed further when comparing values for identical gas mixtures determined in different laboratory test apparatus. Müffling (1946) performed such a comparison and grouped the apparatuses in eight categories as follows:

(a) The explosive gas mixture is passed through a tube of known internal wall temperature.

(b) The mixture is admitted to a vessel of known internal wall temperature.

(c) The mixture is compressed adiabatically and T_{min} is calculated from the lowest compression ratio that gives ignition.

(d) The combustible gas and the air are preheated separately to the desired test temperature and subsequently mixed.

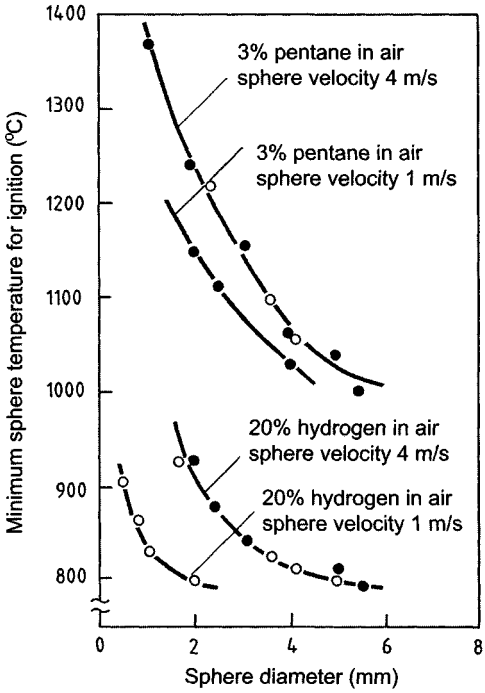


Figure 2–14 Ignition of explosive gas mixtures by hot solid spheres injected into the gas, showing the influence of sphere diameter and velocity on minimum sphere temperature for ignition: (•) quartz spheres; (°) spheres of aged platinum. From Silver (1937) and Paterson (1939, 1940).

(e) A jet of the combustible gas is injected into a vessel containing air preheated to the desired test temperature.

(f) Cool explosive mixture is admitted to a soap bubble surrounding a hot platinum wire of known surface temperature.

(g) A hot solid body of known surface temperature is dropped or ejected into cool explosive mixture.

(h) A hot solid rod of known surface temperature is inserted into cool mixture.

Müffling (1946) reported some results for 7 vol.% hexane in air, showing that method (c) gave a T_{min} of only 300°C, whereas method (d) gave a

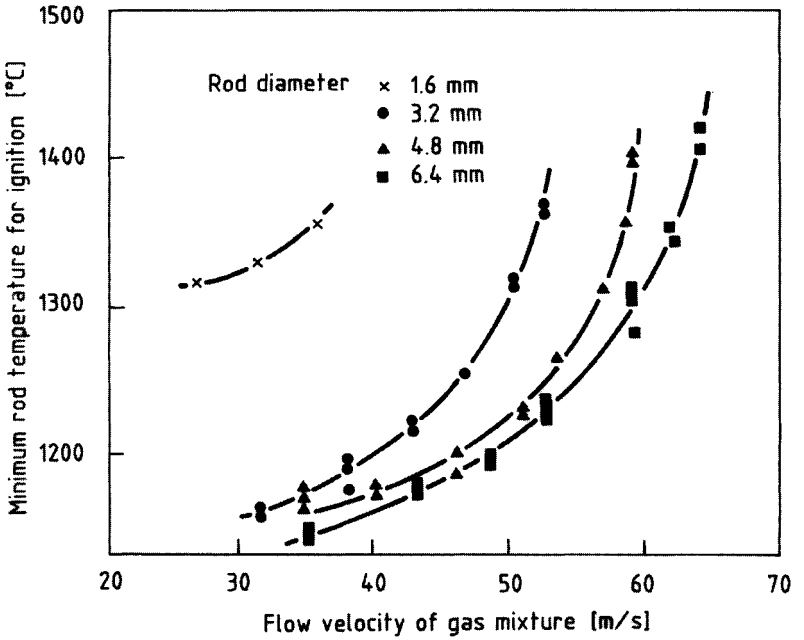


Figure 2–15 Ignition of flowing stoichiometric pre-mixed pentane/air by hot stationary metal rods inserted into a gas flow, showing the influence of the rod diameter, and the gas velocity in relation to the rod, on the minimum rod temperature for ignition. From Mullen et al. (1949).

value of 630°C. For 4–5 vol.% heptane in air the corresponding values were 280°C and 580°C, respectively. The difference of about 300°C in both cases is substantial.

Laurendeau (1982) summarized some more recent published experimental and theoretical work on hot surface ignition of various fuel/air mixtures, with the main emphasis on methane/air. The general validity of the overall trends exhibited by Figure 2–14 and Figure 2–15 was confirmed.

In a subsequent investigation, Alfert and Fuhre (1988) ignited propane/air mixtures in the apparatus shown in Figure 2–16.

After covering the top of the box by a sheet of aluminum foil, making provision for two gas outlet tubes, the box was flushed gently with a propane/air mixture of the desired composition, entering the box through the inlet in the bottom of the box. The rise of fuel concentration in the box

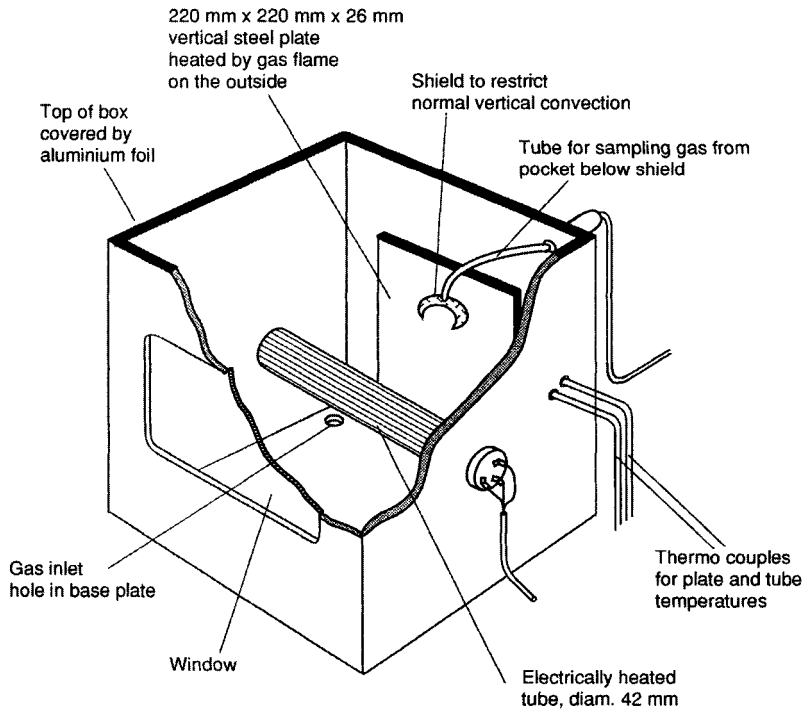


Figure 2–16 Sketch of 50 liter explosion box used for determining minimum ignition temperatures for propane/air under various conditions. In addition to the gas inlet, holes for gas concentration and gas temperature measurement probes were also provided in the base plate. From Alfert and Fuhre (1988).

was monitored continuously. The hot surface was generated when the overall fuel concentration approached the inlet concentration.

When the hot surface temperature reached 600°C, the heating rate was reduced to generate a surface temperature increase of a few degrees per minute. Corresponding values of the hot surface temperature and the propane concentration in the box were recorded at the moment of ignition, the latter being in the range of 4.5–5.0 vol.% in most of the experiments.

The results are summarized in Figure 2–17 and compared with data obtained using the two standard test methods ASTM (2003) and IEC (1975), and a closed bomb method described by Kong et al. (1995).

The results from the experiments in the apparatus shown in Figure 2–16 were from 300°C to 500°C higher than those obtained in the three conservative flask/bomb tests, including the current standard ASTM (2003) and IEC (1975) tests for minimum ignition temperature. This discrepancy calls for a reconsideration of the general applicability of results from current methods of standardized testing for T_{min}.

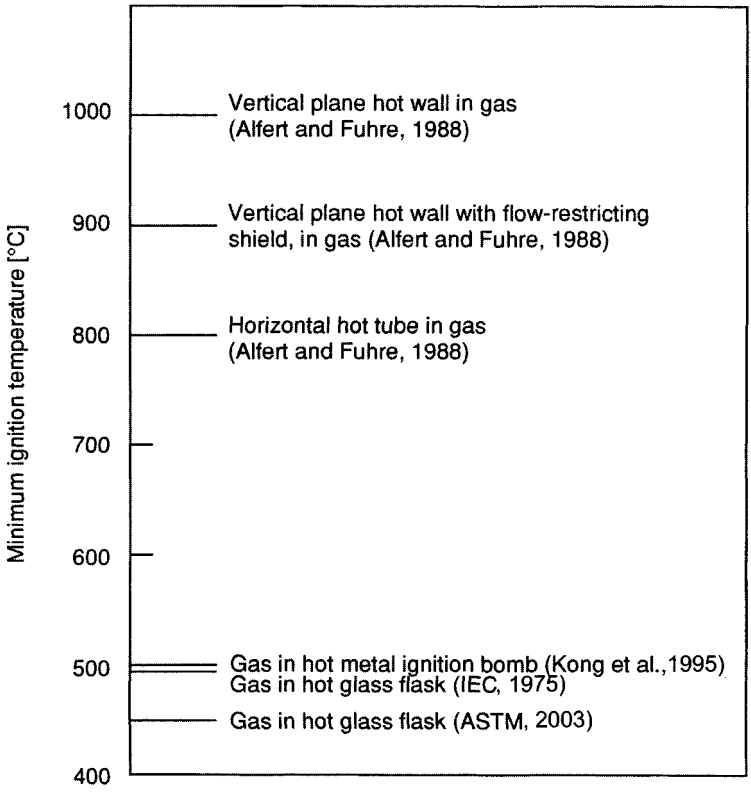


Figure 2–17 Minimum hot surface ignition temperatures for a propane/air mixture at atmospheric pressure, determined by different methods.

2.2.4.2 Minimum Ignition Temperatures of Multi-Component Fuels in Air

In offshore oil and gas production, minimum ignition temperatures of mixtures of multi-component fuels and air are often of interest. For example, natural gas is a multi-component fuel, containing higher hydrocarbons in addition to the methane. For mixtures of methane, propane and

air Kong et al. (1995) investigated the dependence of the ratio of propane/methane on T_{min} of the mixtures. Figure 2–18 shows a significant non-linear decrease of T_{min} with increasing propane/methane ratio. In the hot-bomb method used, a very high equivalence ratio Φ of 2.3 gave a significantly lower T_{min} than for $\Phi = 1$, i.e., for stoichiometric mixtures.

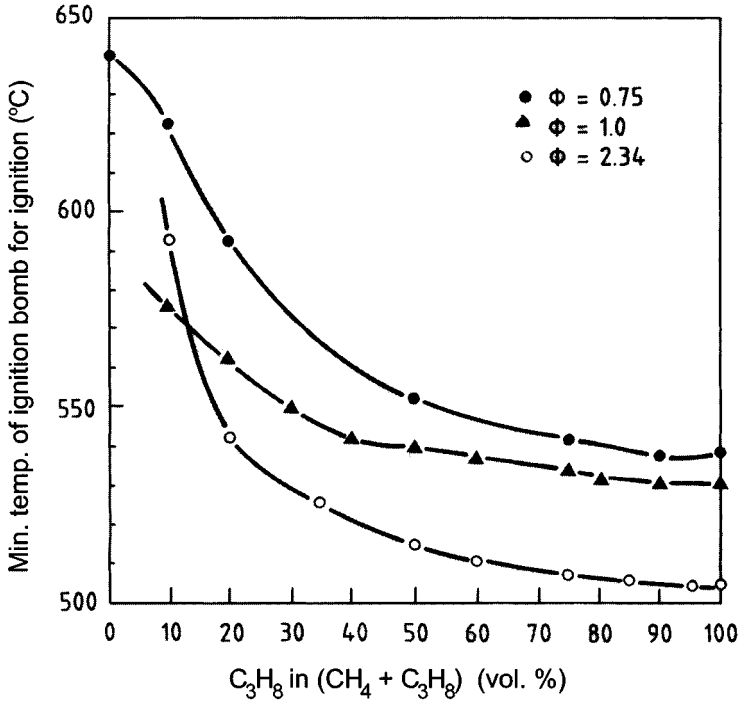


Figure 2–18 Influence of fuel composition on the minimum ignition temperature of CH_4/C_3H_8/air mixtures, for three different overall equivalence ratios Φ, where $\Phi = 1$ for a stoichiometric fuel/air mixture, and $\Phi < 1$ and $\Phi > 1$ represents lean and rich mixtures respectively. From Kong, Eckhoff, and Alfert (1995).

2.2.4.3 Solution for the Future: Dynamic Computer Simulation Models of Hot Surface Ignition

There is a need to reconsider current practices for assessing T_{min} of explosive gaseous fuel/air mixtures for the purpose of process design. A differentiated approach should be developed, which allows critical ignition

conditions to be predicted for both mono- and multi-component fuels, for different geometric hot surface configurations, and for the conditions of heat and gas flow that actually occur in the industrial situation of concern.

The likely approach for the future will be development of comprehensive numerical models containing sub-models of both chemical kinetics and transport processes. Because of the rapid development of fast computers and advanced methods of measurements, reaction kinetics modeling is progressing at great pace. Simulation of fuel oxidation by considering elementary reaction steps has shown good agreement with experimental results.

Integration of a full chemical kinetics package into a mathematical model for predicting T_{min} in practical process situations may be too ambitious at the outset. Using semi-empirical approximations of the influence of chemistry could be a more realistic point of departure. From this perspective, the work of Oberhagemann (1989) is interesting. He developed a semi-empirical model by which T_{min} for single organic components in air can be computed on the basis of the molecular structure of the fuel. Using this model, he computed T_{min} for 380 different organic fuels and correlated with experimental values obtained by the standard IEC (1975) method. The coefficient of correlation was 0.97, corresponding to a mean deviation between experiment and theory of about 10°C. Oberhagemann extended his model to mixtures of two, three, and four different fuels and air, and obtained good correlation with experimental values even in these cases.

2.2.4.4 Standard Test Methods for T_{min}

The current international standard test apparatus used for determining the T_{min} value of explosive gas mixtures for practical explosion prevention is shown in Figure 2–19. However, this is a quite conservative method, as illustrated in Figure 2–17.

In spite of the two standard methods IEC (1975) and ASTM (2003) being very conservative, they may not produce the absolute minimum values that can be found for any given fuel/air mixture. This is because T_{min} determined by closed, isothermal vessel methods of this type decrease somewhat with increasing vessel size. Hence, tests in significantly larger isothermal vessels than the comparatively small vessels used in the two standard methods may yield somewhat lower T_{min} than those obtained with the standard methods.

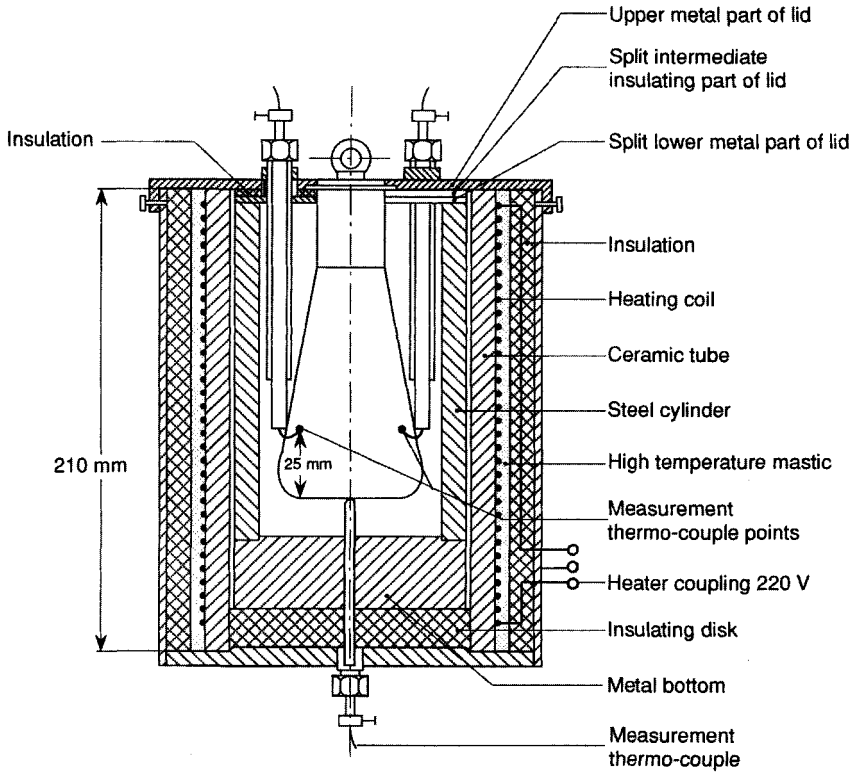


Figure 2–19 The IEC standard 200 cm^3 glass flask apparatus for determining the T_{min} values of explosive mixtures of combustible gases and vapors with air.

2.2.5 Ignition by Burning Metal Particles, "Thermite" Reactions, and Transient Hot Spots

2.2.5.1 Introductory Overview

In the past, the initiation of accidental gas explosions was sometimes attributed to "friction sparks" without any further explanation. However, this imprecise term covers a multifarious category of potential ignition sources, comprising mechanical hot-work operations such as grinding and cutting, repeated impacts on one spot, and single impacts. However, in all these cases, the ignition source is generated by transformation of mechanical energy (impact and friction) into heat.

Existing knowledge suggests that the probability of ignition of a combustible gas-in-air mixture by a mechanical impact is determined by a set of basic parameters characterizing the impact and a set of basic parameters characterizing the gas mixture. Key parameters of the impact include

- chemical composition of colliding bodies
- surface topography of colliding bodies at areas of contact
- contact area
- contact pressure
- sliding velocity parallel to the contact surface
- sliding distance (or contact time)

Key parameters of the gas include

- chemical composition of the fuel gas
- concentrations of fuel, oxygen and inert gas
- dynamic state of the gas (turbulence and systematic velocity components, which are also influenced by the impact process itself)
- temperature
- pressure

Depending on the circumstances, the actual ignition source produced by the impact/friction process is generally either the small hot particles released during the process or the hot spot generated on one or both of the colliding bodies during the impact. In special situations where metals such as aluminum or titanium are involved in the impact together with rust, the impact can generate very incendiary "thermite" flashes due to exothermic transfer of oxygen from the rust to the aluminum or titanium (see Section 2.2.5.3).

2.2.5.2 Ignition by Small Burning Metal Particles from Single Impacts

Small flying hot particles from single impacts between chemically inert materials (rock and other inert minerals) cannot ignite explosive gas mixtures. This is because the mechanically generated temperature rise will not reach the high levels required for such small particles to become an

ignition source. However, if metals are involved in the impact, small metal particles may be torn or cut away from the main bulk and heated mechanically to such an extent that they start to burn spontaneously while flying through the air. Then the particle temperature will increase substantially due to the release of chemical energy. Such small burning metal particles are the genuine "friction sparks".

Much work has been carried out to assess the possibility of igniting methane/air ("firedamp") in coal mines by coal-picking equipment impacting on rock. This is essentially a single-impact process. Research over many years has revealed that, if ignition occurs, the source is not the burning metal sparks (steel/hard metal), but the transient hot spot generated on the pick after multiple impacts on the rock. In his extensive review paper, Powell (1984) concluded that small burning metal particles from mechanical impacts are not capable of igniting methane/air, and possibly not even higher alkanes/air, unless the particle temperature exceeds 2000°C. This means that steel sparks from single impacts are unlikely to ignite natural gas/air. However, burning particles of titanium, zirconium, magnesium and aluminum can ignite such gases. According to Powell (1984) the low probability of igniting methane/air and group IIA gases/air (see Table 2–2) with steel sparks from single impacts is in accordance with American and British recommendations saying that "non-sparking" tools are superfluous in areas where such gases may be present.

Pedersen and Eckhoff (1986) studied the ignition of propane/air and acetylene/air by heat generated in tangential impacts between tips of different steel qualities or of titanium and a rusty or sandblasted steel plate. Tangential impacts against the steel plate were generated by a rigid spring-loaded arm carrying the test tip. The apparatus is illustrated in Figure 2–20.

The strength of the impact (net impact energy) was expressed in terms of the loss of kinetic energy of the impacting arm during the impact. Some results are shown in Table 2–9.

Within the experimental range of net impact energies up to 20 J, it was not possible to ignite 4.6 vol.% propane/air with sparks or hot spots from single impacts between different steel qualities and rusty or sandblasted steel. This agrees with the general conclusion of Powell (1984) on ignition of groups I and IIA gases, mentioned above.

Table 2–9 Results from Impact Ignition Experiments Using Different Tip Materials Impacting on a Target of Naturally Rusted Steel in an Explosive Mixture of 4.6 vol.% Propane in Air[a]

Tip material	Number of visible sparks	Net impact energy (J)	Ignition
St 37 steel	20–50	8–10	No
Chrome–vanadium steel	10–30	6	No
Unbraco screw	~20	8	No
Acid-resistant screw	~5	8	No
Non-sparking tool	0	13–14	No
Titanium	10–1000	~10–15	Yes

a. *Source*: Data from G. H. Pedersen and R. K. Eckhoff, *Initiation of Gas Explosions by Heat Generated During Single Impact Between Solid Bodies,* CMI Report No. 863302-1, December 1986, Christian Michelsen Institute (now GexCon AS), Bergen. Norway (1986).

However, as Table 2–9 shows, titanium sparks were able to ignite propane/air. In about half of the experiments with titanium that gave ignition, one or several specific flying sparks could be identified by high-speed video as the ignition source(s). Such ignition mostly took place 50–90 ms after the impact. By this time, the velocity of the sparks had dropped to 2–5 m/s, which presumably allowed a sufficient residence time in a given gas volume for ignition to occur. In some cases, single-spark ignition took place after the spark had collided with the wall of the explosion chamber and lost most of its velocity. Similar observations were made by Ritter (1984).

Figure 2–21 shows some more detailed results from impacts with titanium against rust in 4.6 vol.% propane in air. Observe that there seems to be a worst-case range of tangential impact velocities that favors ignition. At lower velocities fewer burning metal particles are produced, and ignition is less probable. At higher velocities more burning metal particles may be produced, but this is more than compensated for by the violent disturbance of the local gas cloud by the movement of the impact arm, which makes ignition less probable.

Some results from ignition of acetylene/air by burning metal particles are given in Table 2–10, showing that highly sensitive gas mixtures can be ignited even if the particle material is stainless steel.

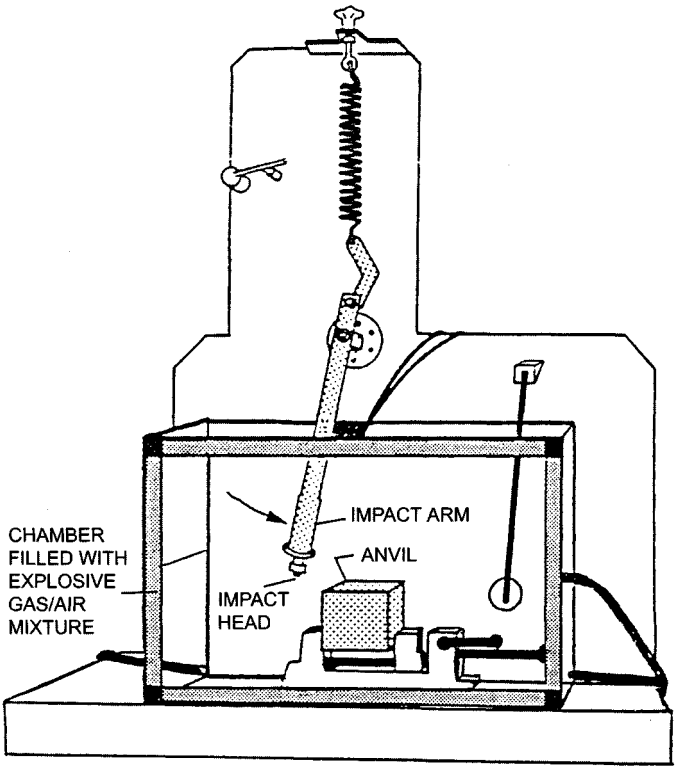


Figure 2–20 Impact spark ignition apparatus used by Pedersen and Eckhoff (1986) for studying the ability of burning metal particles and "thermite flashes" from single impacts to ignite explosive gas/air mixtures.

As can be seen, the acetylene/air mixture was readily ignited by sparks from steel qualities that are known to be less able to release incendiary sparks than standard construction steels. High-speed video recordings revealed that ignition was caused by single flying steel sparks in all the cases covered by Table 2–10. Ignition occurred after the sparks had traveled sufficiently far for the spark velocity to become < 5 m/s. Ignition of acetylene/air by the hot spot generated at the point of impact on the anvil plate was not observed in these experiments.

When using a non-sparking material as the test object, only one single visible spark was observed altogether. This spark became visible 45 ms after the impact, and remained visible for 90 ms. During this period, the spark velocity decreased from about 1 m/s to 0.4 m/s. It is not clear whether this weak spark originated from the non-sparking material or

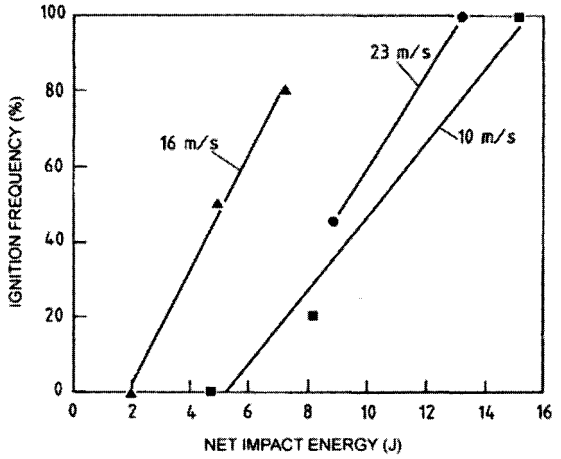


Figure 2–21 Results from tangential impact experiments with titanium against rust in a mixture of 4.6 vol.% propane in air, using the apparatus illustrated in Figure 2–20. From Pedersen and Eckhoff (1986).

Table 2–10 Results from Impact Ignition Experiments Using Different Tip Materials Impacting on a Target of Naturally Rusted Steel in an Explosive Mixture of 7.7 Vol.% Acetylene in Air[a]

Tip material	Number of visible sparks	Net impact energy (J)	Ignition
Chrome–vanadium steel	~10	~6	Yes
Unbraco screw (stainless)	~20	~8	Yes
Acid-resistant steel	3–5	~8	No
Non-sparking tool alloy	0–1	8–10	No

a. *Source*: Data from G. H. Pedersen and R. K. Eckhoff, *Initiation of Gas Explosions by Heat Generated During Single Impact Between Solid Bodies,* CMI Report No. 863302-1, December 1986, Christian Michelsen Institute (now GexCon AS), Bergen. Norway (1986).

from the anvil material, which was rusty steel. Although the spark observed was very weak, its occurrence indicates that impacts involving so-called non-sparking materials as one of the colliding partners may in fact generate visible sparks. However, in view of what has been said above, it seems highly unlikely that such sparks can ignite alkane/air mix-

tures. It is not clear whether impacts between a non-sparking material and light metals, e.g. titanium, may produce incendiary light-metal sparks.

2.2.5.3 Ignition by Thermite Flashes

The experiments in propane/air with impacts of titanium on rusty St37 steel (see Table 2–9) revealed two different modes of ignition. In about half of the tests where ignition occurred, it was observed to take place at or very close to the point of impact on the anvil plate, immediately after or even during impact. In such cases, an extremely luminous hemispherical volume was observed in the region of impact. The onset of flame propagation could neither be referred specifically to one single metal spark nor to the hot spots generated at the points of impact. It is not fully clear what this luminous hemisphere consisted of, but the very high numbers of sparks (of the order of 1,000) that were observed in these cases were located within this volume. The luminous hemisphere was only observed when the peripheral velocity of the tip holder exceeded about 15 m/s. It is believed that the luminous volume was a "thermite flash" corresponding to the overall formal exothermic reaction

$$3Ti + 2Fe_2O_3 \rightarrow 3TiO_2 + 4Fe + \text{heat}$$

Similar reactions would be expected between rust and other metals having greater affinities for oxygen than iron. In the case of aluminum, the softness of the metal may prevent sufficient heat generation during impact to produce a thermite flash. However, according to Kornai et al. (1994), impacts involving harder aluminum alloys may generate thermite flashes, as well as incendiary metal particles. Also, impacts on rusty surfaces covered by aluminum paint of high pigment content have generated thermite flashes. Gibson et al. (1968) demonstrated that smears of aluminum on rusty steel produced thermite flashes capable of igniting methane/air when struck by strikers made from almost any metal, including steel, brass, bronze, and even copper-beryllium.

2.2.5.4 Ignition by Transient Hot Spots

As already pointed out, impacts not only create burning metal particles or thermite flashes but also create transient hot surfaces (hot spots) on the two colliding bodies. Powell and Quince (1972) applied the classical theory of frictional impacts to calculate the maximum hot spot temperatures generated in such impacts. Eckhoff and Pedersen (1988) discussed this theory in relation to impact ignition hazards on offshore oil and gas installations. Cutler (1974, 1978) conducted experiments where methane/air and propane/air were ignited by artificial transient hot spots generated electrically on tungsten strips. Figure 2–22 shows how the values of T_{min} were influenced by the strip width.

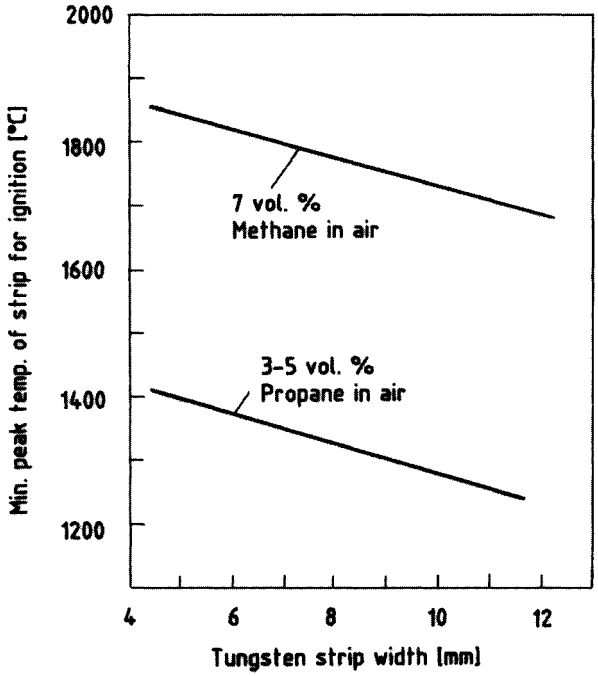


Figure 2–22 Influence of the width of a transiently heated tungsten strip on the minimum peak temperature of the strip for igniting methane/air and propane/air. Data from Cutler (1974, 1978).

As pointed out above, experiments with net impact energies of up to 20 J in acetylene/air revealed that ignition never occurred at the hot spot on the anvil plate or on the impacting metal tip. Therefore, net impact energies that are substantially higher than 20 J would be required for generation of hot spots by single impacts, which would be capable of igniting natural gas/air.

2.2.6 Ignition by Electric Sparks and Arcs and Electrostatic Discharges

2.2.6.1 Electric Sparks between Two Conducting Electrodes

Electric sparks are produced when the strength of the electric field in the gap between two conducting electrodes exceeds what the dielectric medium in the gap can resist. For a given dielectric medium, e.g. air at atmospheric conditions, the relationship between the gap distance and the critical gap voltage for gap breakdown depends on the electrode shape and the electrode material.

Figure 2–23 gives some results obtained in two different laboratories from measurement of electrical breakdown voltages in air at atmospheric conditions using various electrode configurations. All data confirm that, for one specific electrode system, the electrical breakdown voltage increases systematically with the length of the gap between the electrodes. Also, the data confirm that for a given gap length the electrical breakdown voltage is considerably lower with needle point electrodes than with rounded electrodes having radii of curvature of at least a few mm. The discrepancies between the absolute breakdown voltages obtained in the two laboratories for apparently similar conditions illustrate that the actual breakdown voltage also depends on other experimental conditions than just global electrode shape and gap distance.

Extrapolation of the various curves in Figure 2–23 into the region of voltages of the order of 100 V and less indicate that spark-over will then be possible only at very small gap distances $<<$ 1 mm. This in turn means that any ignition of an explosive gas mixture under such circumstances occurs under partly quenched conditions (see Figure 2–26). This means that the required spark energy for ignition can be considerably higher than the minimum ignition energies obtained under unquenched conditions.

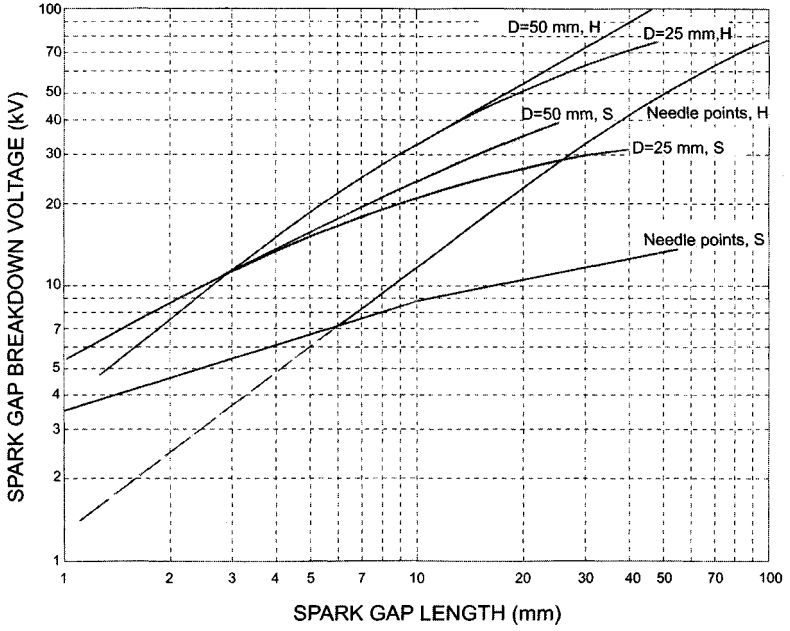


Figure 2–23 Spark gap breakdown voltages in air at normal atmospheric conditions as functions of spark gap length and electrode shape. "H" indicates data from Handbook of Chemistry and Physics (1959/60) and "S" indicates data from Smithsonian Physical Tables (1959). Details of electrode materials and electrode surface structures are not known.

Electric sparks can be capacitive, inductive, or resistive. Capacitive sparks, associated with the discharge of a capacitor across a given spark gap, can occur in electric circuits, or they can be caused by tribo-electrically generated electrostatic charges. The theoretical spark energy, neglecting losses, is $\frac{1}{2}CV^2$, where C is the capacitance and V is the voltage across the spark gap just prior to gap breakdown. Inductive sparks are associated with discharge of inductive energy across gaps that are formed when live circuits are broken. In this case, the theoretical spark energy is $\frac{1}{2}Li^2$ where L is the inductance and i is the current just before formation of the break. Any given explosive mixture of a combustible gas and air is associated with a given minimum value of electric spark energies E_{min} that can ignite the mixture. Figure 2–24 illustrates this for the alkanes, e.g. which constitute the main components in natural gas from the North Sea. It is seen that the values of the equivalence ratio Φ at which the minimum ignition energy for each gas occurs increase systematically

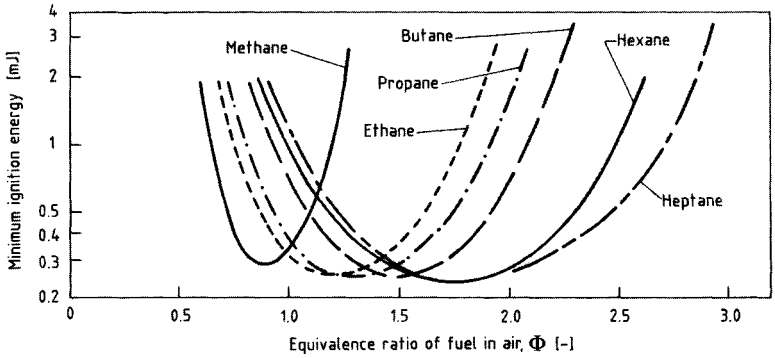


Figure 2–24 Minimum capactive electrical spark ignition energies of mixtures of various gaseous alkanes and air as a function of the volumetric ratio of fuel and air. For $\Phi = 1$ the ratio of fuel to oxygen is stoichiometric. For $\Phi < 1$ the mixture is lean and for $\Phi > 1$ it is rich.

with increasing molecular weight of the fuel gas, from 0.85 for methane (Mw = 16) to 1.80 for heptane (Mw = 100). It is also seen that the standard E_{min} value of methane (0.28 mJ) is slightly higher than those of the higher alkanes (0.23–0.25 mJ).

Figure 2–25 shows curves corresponding to those in Figure 2–24, for a range of different explosive gas/air mixtures. As can be seen, E_{min} for the most ignition sensitive gases hydrogen, acetylene and carbon di-sulphide, are about two orders of magnitude lower than the values for the alkanes in Figure 2–24.

All the E_{min} values in Figure 2–24 and Figure 2–25 were determined by capacitive sparks under unquenched conditions, which implies that the fixed spark gaps used in the test were slightly larger than the quenching distances QD for the gas mixtures tested. This quenching distance concept is illustrated in Figure 2–26. True E_{min} values are obtained only if the length of the electrode gap exceeds QD. As soon as the gap length becomes < QD the minimum spark energy for ignition can increase quite significantly. The increase will be more or less abrupt, depending on the physical dimensions of the electrodes, as illustrated in Figure 2–26.

Figure 2–27 illustrates the close correlation between E_{min} and QD.

It is generally assumed that the minimum ignition energies obtained for capacitive sparks by optimizing spark gap length and fuel/air ratio are the true minima. However, this may not be entirely true. Figure 2–28 gives

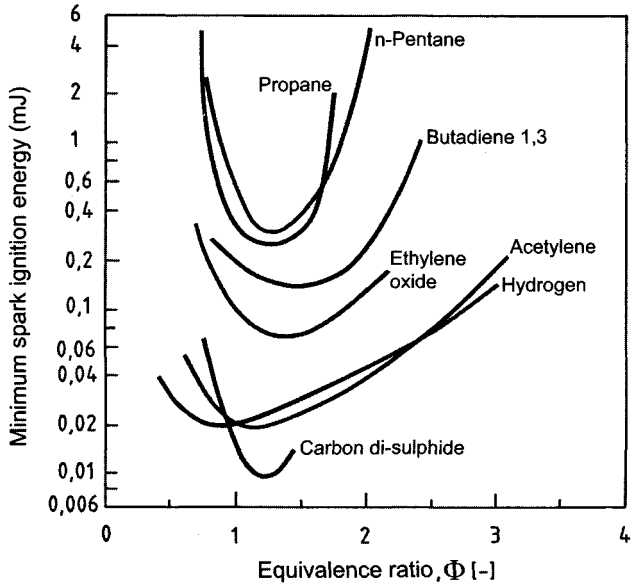


Figure 2–25 Minimum capacitive electrical spark ignition energies of mixtures of various gaseous fuels and air as a function of the volumetric ratio of fuel and air. For $\Phi = 1$ the ratio of fuel to oxygen is stoichiometric. For $\Phi < 1$ the mixture is lean and for $\Phi > 1$ it is rich.

some results from a study by Parker (1985), showing that minimum ignition energies may also depend on the basic features of the spark discharge. Using constant-power sparks of mutually independent energy and duration, Parker found that the minimum ignition energy of a lean propane/air mixture increased from about 0.2 mJ for 0.1 μs discharge duration to about 2 mJ for 100 μs duration. The reason for this quite pronounced effect may be found in the fundamental process of ignition, where the production of chemically active radicals in the spark, which is enhanced by the high spark temperatures generated at short discharge times, plays a central role.

Various electrical apparatus can give rise to electric sparks and arcs that can cause ignition of explosive gas mixtures. Electric sparks can be generated when electric circuits are opened and closed, or by stray currents. The use of low-voltage equipment (maximum = 50 V) for protecting personnel against electric shock does not eliminate the explosion hazard. Even lower voltages than this can produce energies which can ignite explosive gases. It has been found useful, therefore, to introduce an

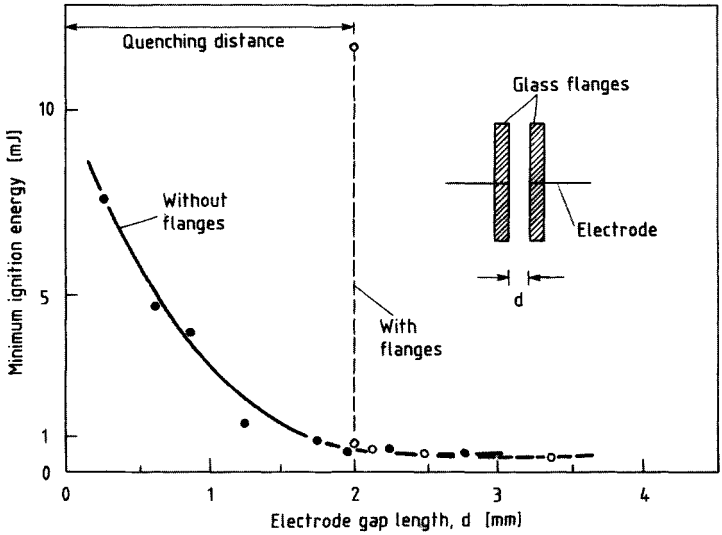


Figure 2–26 Minimum capacitive electric spark energies for ignition of a stoichiometric mixture of natural gas and air, as a function of length of gap between electrodes, with and without glass flanges at the electrode tips. Data from Freytag (1965).

international standard method for experimental determination of the combinations of electrical circuit parameters that are critical for producing incendiary sparks/arcs from electrical circuitry. The IEC standard apparatus is illustrated in Figure 2–29.

Figure 2–30 shows experimental ignition curves (solid lines) for propane/air (IIA), ethylene/air (IIB) and hydrogen/air (IIC) determined in the apparatus illustrated in Figure 2–29. Ignition occurred at capacitor voltages even below 10 V, which is only possible with minute spark gap lengths of << 1mm. This in turn means that the sparks were delivered to the explosive mixture under quenched conditions (see Figure 2–26). It is not surprising, therefore, that the dotted lines, representing the minimum $\frac{1}{2}CU^2$ for igniting the three different gas mixtures with spark gaps > the quenching distances, falls significantly below the experimental solid lines.

As shown in Figure 2–31, a corresponding effect is found for inductive circuits, where the lines for $\frac{1}{2}Li^2 = E_{min}$ are also significantly below the corresponding minimum break flash ignition curves (minimum ignition current as a function of circuit inductance).

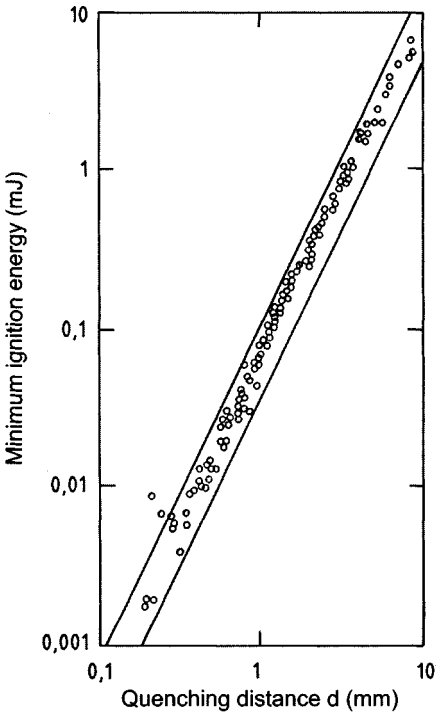


Figure 2–27 Correlation of minimum electric spark ignition energies for unquenched ignitions, and corresponding quenching distances. From Kuchta (1985).

Besides sparks from electrical apparatuses, spark discharges between two conducting electrodes can also arise from tribo-electric charging of non-earthed electrically conducting items. Table 2–11 indicates the levels of capacitance C and voltage V that may be associated with electrostatic charging of non-earthed electrical components in industry. The resulting stored energies $\frac{1}{2}CV^2$ represent the maximum spark energies that can be generated from discharge of the various capacitive components to earth.

Table 2–11 also includes the charging of a human being. For example, charging can occur whenever a person wearing electrically insulating shoes is walking across a floor. Charge transfer occurs every time the shoes are lifted or separated from the floor. Figure 2–32 illustrates the discharge of an electrostatically charged person to earth. In electrical terms, the human body can be regarded as a capacitor of the order of 100–300 pF, with a given internal ohmic resistance. It is commonly

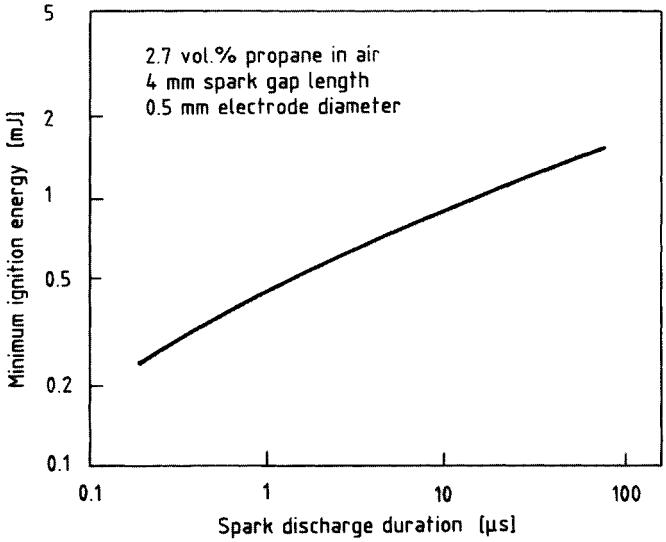


Figure 2–28 Influence of spark discharge duration on minimum ignition energy of a lean (2.7 vol.%) mixture of propane/air at normal temperature and pressure. Data from Parker (1985).

assumed that during discharge about half the energy $\frac{1}{2}CV^2$ stored in the capacitor is dissipated in the spark and the other half in the internal body resistance. As Table 2–11 shows, spark energies well above E_{min} for most gases and vapors can be generated when a charged person is discharged to earth. In many process operations it is therefore necessary to take active measures to make sure that electrostatic charging of persons cannot occur, or to ensure that any charged person is reliably discharged before entering an area where explosive gas atmospheres may occur.

The most important measure for preventing electrostatic spark discharges is earthing of all conductive parts that could become dangerously charged (see Table 2–11). However, this protective measure is insufficient if non-conductive materials are present and become electrostatically charged to hazardous levels. In this case electrostatic one-electrode discharges can occur which may also ignite explosive gas/air mixtures. Then non-conductive materials must be avoided. This may imply that only earthed metals and earthed anti-static non-metals are permissible in e.g. process equipment.

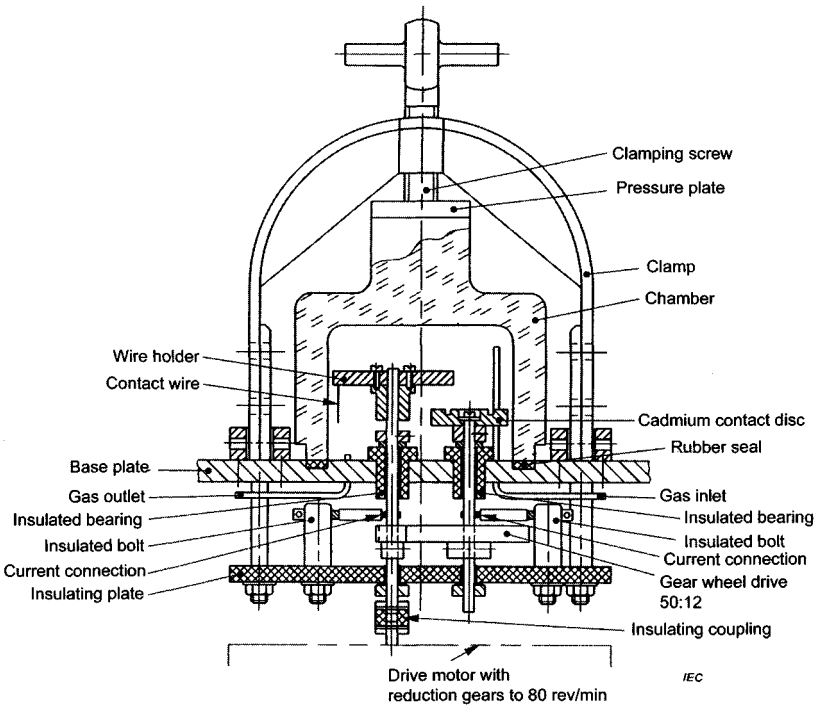


Figure 2–29 Standardized IEC apparatus for experimental determination of ignition curves, and for testing of the ability of electrical circuits to produce sparks that can ignite gas mixtures. From IEC (1999).

2.2.6.2 Various Forms of One-Electrode Electrostatic Discharges from Charged Non-Conductors: Concept of "Equivalent Energy"

A one-electrode discharge occurs when electrostatic charge accumulated on the surface of a non-conductor by a charge separation process, is drained to earth via a conducting electrode approaching the charged non-conductor. It is common to distinguish between three types of one-electrode discharges

(1) corona discharge

(2) brush discharge

(3) propagating brush discharge

Lüttgens and Glor (1989) and Lüttgens and Wilson (1997) have discussed the nature of the three types of discharges in detail and have illustrated their possible appearance in practice by examples.

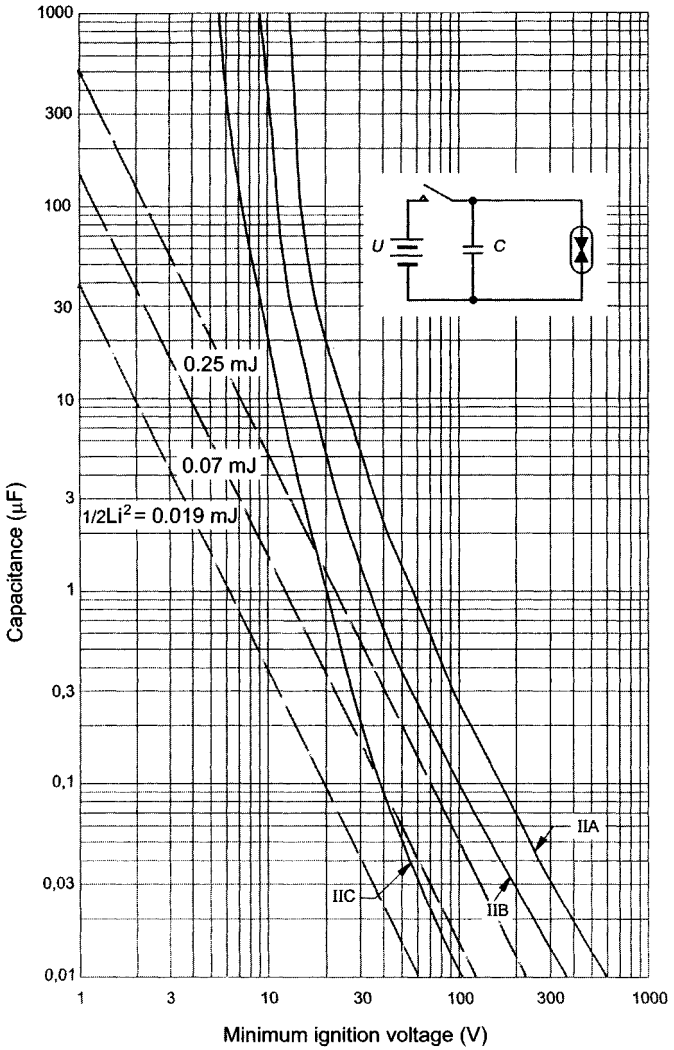


Figure 2–30 Experimental standard ignition curves for propane/air (IIA), ethylene/air (IIB) and hydrogen/air (IIC) for capacitive spark discharges, determined by the standard IEC break flash apparatus shown in Figure 2–29. The dotted straight lines represent all combinations of voltage and capacitance that yield ½ CU^2 values equal to the experimental minimum ignition energies for the three respective gas mixtures. The standard ignition curves are taken from IEC (1999).

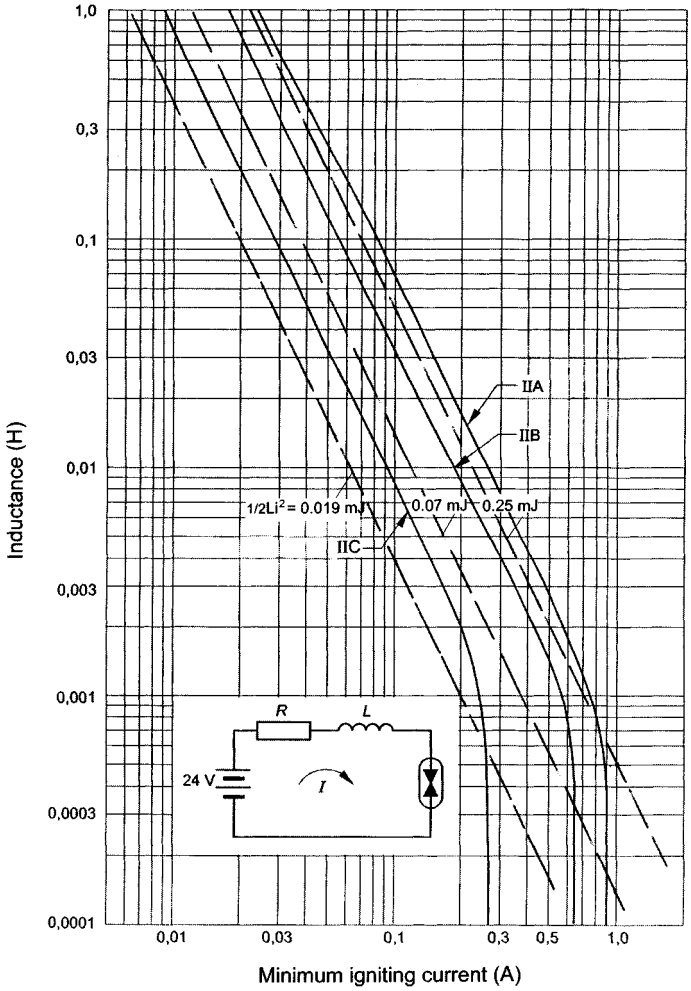


Figure 2–31 Experimental standard ignition curves for propane/air (IIA), ethylene/air (IIB) and hydrogen/air (IIC) for inductive spark discharges, determined by the standard IEC break flash apparatus shown in Figure 2–29. The dotted straight lines represent all combinations of current and inductance that yield ½ Li^2 values equal to the experimental minimum ignition energies for the three respective gas mixtures. The standard ignition curves are taken from IEC (1999).

Corona discharges, illustrated in Figure 2–33, occur when the tip of a pointed conducting electrode (radius of curvature < 1 mm) approaches a charged non-conductor. This type of discharge does not occur abruptly

Table 2-11 Examples of Combinations of Capacitances and Voltages and Resulting Spark Energies in Industrial Practice

Charged object	Capacitance (pF)	Potential (kV)	Energy (mJ)*
Single screw	1	5	0.01
Flange, nominal width = 100 mm	10	10	0.5
Shovel	20	15	2
Small container (~50 litres)	50	8	2
Funnel	50	15	6
Person	300	10	15
Drum (200 litres)	200	20	40
Road tanker	1000	15	100

Data are taken from Luttgens and Glor (1989)
*Approximate values

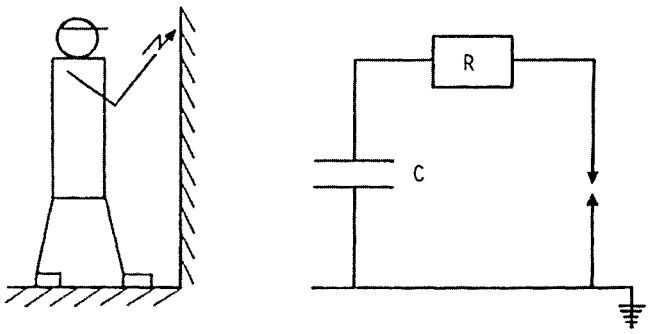


Figure 2–32 Discharge of an electrostatically charged person to earth. Approximate equivalent electrical circuit of a human being, with internal resistance and capacitance in series. From Glor (1988).

within a short time interval as a spark, but rather as a continuous charge leakage, and only gas mixtures of very low minimum ignition energies can be ignited by corona discharges.

However, when the radius of curvature of the conducting electrode tip exceeds a few min, the discharge will occur more abruptly, and a stem of high energy density will appear close to the electrode. Because of this appearance, this type of discharge, illustrated in Figure 2–34, is called a *brush discharge.*

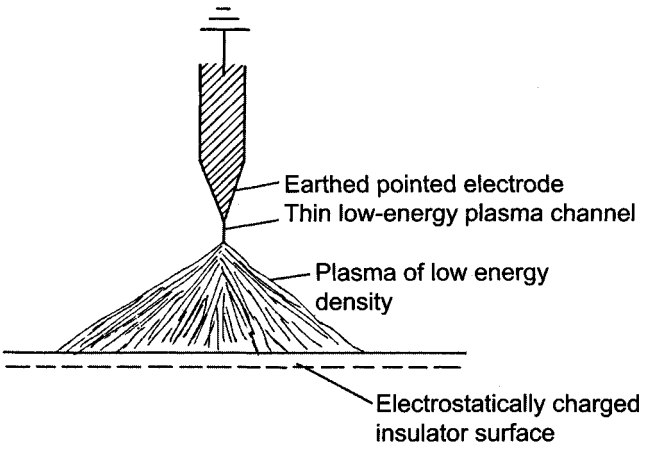


Figure 2–33 Illustration of a corona discharge from a charged insulator surface to earth.

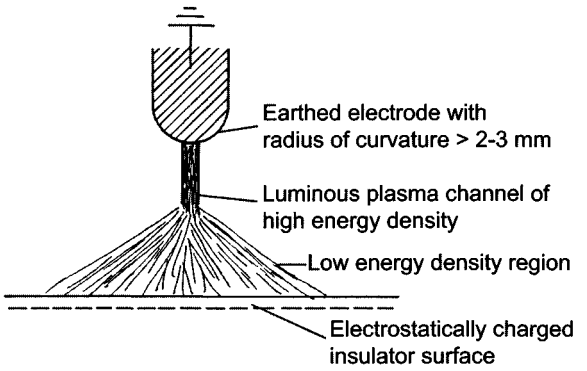


Figure 2–34 Illustration of a brush discharge from a charged insulator surface to earth.

Substantially stronger and more incendiary *propagating brush discharges* than ordinary brush discharges can be obtained if a charged double layer of opposite polarities is generated rather than just a single charge layer, as illustrated in Figure 2–35 and Figure 2–36. This can occur if an insulating film is charged tribo-electrically with opposite charge polarity on both sides, or if the side of the film which is not charged directly can be charged by induction from a distant or close earthed object.

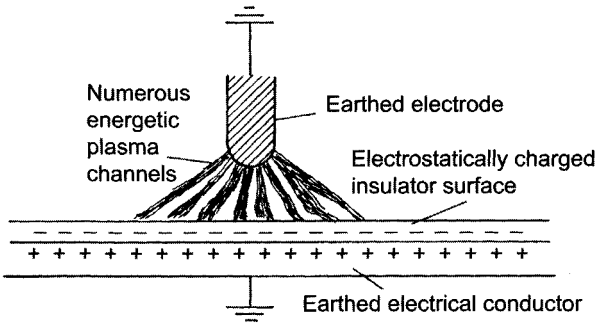


Figure 2–35 Illustration of a propagating brush discharge occurring when an earthed electrode is brought close to the charged insulator surface that is in contact with an earthed conductor.

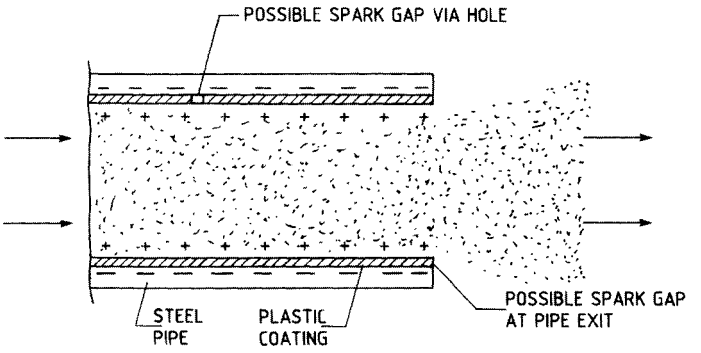


Figure 2–36 Illustration of a propagating brush discharge occurring in a steel pipe lined with an insulating film, when the strong electric field breaks down the charged insulating film and drains the charge to earth through the breakdown channel. Alternatively the charge can occur across a gap at the pipe exit.

For basic reasons, the energy of the incendiary part of a corona, brush or propagating brush discharge is not easy to measure directly. The concept of *equivalent energy* is useful in partly overcoming this problem. This concept, which was first introduced by Gibson and Lloyd (1965), is defined as follows: An electric discharge has the equivalent energy W if it is just capable of igniting an explosive mixture of $E_{min} = W$. The equivalent energy of any reproducible but not quantifiable electric discharge is determined in two steps. The first is to determine the composition of a pre-mixed fuel/air/nitrogen mixture, which can just be ignited by the unknown discharge. The second step is to determine E_{min} of that particular mixture, which will per definition be equal to W. The equivalent energies for the three types of one-electrode discharges discussed above are indicated in Table 2–12.

Table 2–12 Equivalent Energies of Three Types of Electrostatic One-Electrode Discharges

Type of discharge	Equivalent energy (mJ)
Corona	$\leqslant 2 \times 10^{-2}$
Brush	$\leqslant 3$
Propagating brush	$\leqslant 10^5$

Data from Lüttgens and Glor (1989)

The minimum ignition energy for alkanes in air is of the order of 0.25 mJ. Therefore, it is unlikely that such mixtures, including natural gas in air, can be ignited by corona discharges. However, brush discharges may cause ignition, and propagating brush discharges can generate energies that are substantially higher by orders of magnitude than those required for ignition.

2.2.7 Ignition by a Jet of Hot Combustion Products

2.2.7.1 The Basic Process

The situation is illustrated in Figure 2–37.

The central question is whether the hot jet will ignite an explosive gas cloud outside the enclosure of the same composition as the one exploding inside. Systematic research has been carried out to investigate the influence on the probability of ignition of hole diameters/slot width, the length

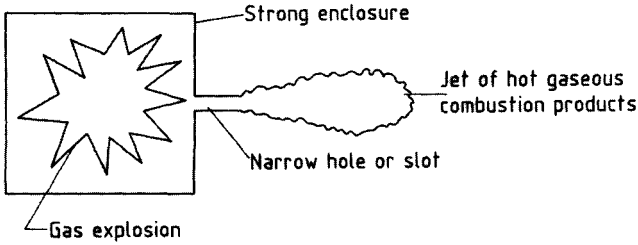


Figure 2–37 Illustration of the ejection of a jet of hot gaseous combustion products through a narrow hole or slit in the wall of a strong enclosure in which a gas explosion takes place.

of hole/slot, the volume of the enclosure in which the primary explosion takes place, the volume of the external explosion chamber, the location of the ignition point inside the primary enclosure, and of course the composition of the gas mixture.

2.2.7.2 Grouping of Ignition Sensitivity of Premixed Gas/Air According to MESG

Various standard test apparatuses (see Figure 2–44) have been designed to determine the maximum experimental safe gap (MESG) for a given explosive gas mixture. Experience has shown that there is a correlation between MESG values obtained in such tests and the quenching distance (QD) discussed above (Figure 2–26). A rough rule of thumb is that QD $\approx 2 \cdot$ MESG. Keeping in mind that there is a strong positive correlation between the quenching distance of a gas mixture and its minimum ignition energy (Figure 2–26), one would expect MESG to vary with gas type and fuel/air ratio in a manner similar to that in Figure 2–24 for the minimum ignition energy. As shown in Figure 2–38, this is indeed the case. Table 2–2 gives some MESG values from standards tests (Figure 2–44) and the resulting grouping of the gases tested in gas groups.

Figure 2–39 shows the apparatus used in some gas explosion transmission experiments conducted by Larsen and Eckhoff (2000). The vertical cylindrical system consisted of two concentric chambers, one primary and one secondary, separated by a disc with the cylindrical transmission hole at its center. Before an experiment both chambers were flushed with premixed propane/air of the desired composition, until this concentration was obtained throughout the system. The electric spark ignition source in the

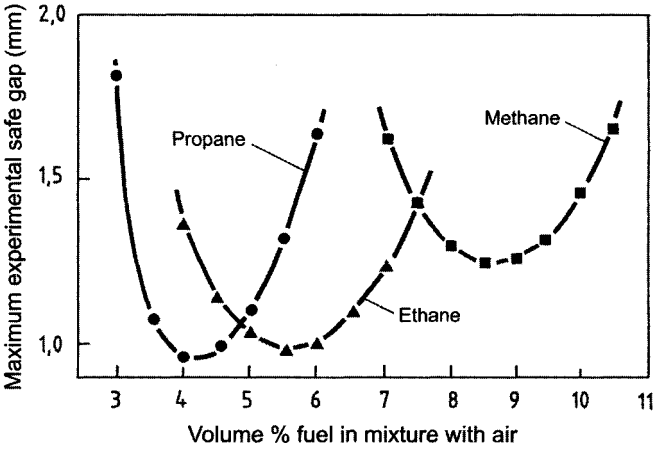


Figure 2–38 Influence of fuel concentration in mixtures with air on MESG for the three lower alkanes. Data from Alfert (1985).

primary chamber was positioned at the chamber axis, and the distance X_i from the ignition point to the entrance of the transmission hole was varied.

Figure 2–40 gives a set of results showing the critical distance X_i for explosion transmission as a function of the hole diameter D. As can be seen, the critical hole diameter for flame transmission with $X_i = 0$ was about 3 mm. In this case the conditions were probably close to laminar, and what was measured was the critical laminar quenching tube diameter.

However, as X_i was gradually increased, a significant amount of gas had to burn before the flame front reached the hole entrance. This means that at the time of flame front arrival at the hole entrance the pressure in the primary chamber was significantly higher than the ambient pressure downstream of the transmission hole. Hence, when the flame front arrived at the hole, hot combustion gases were pushed through the hole, producing a jet into the unburned explosive gas in the secondary chamber. The question was then no longer whether a laminar flame front would be able to travel through the transmission hole, but rather whether re-ignition would occur downstream of the hole, where the hot combustion gases ejected from the hole mixed with the unburned mixture. As Figure 2–40 shows, it appeared that the minimum critical hole diameter for such re-

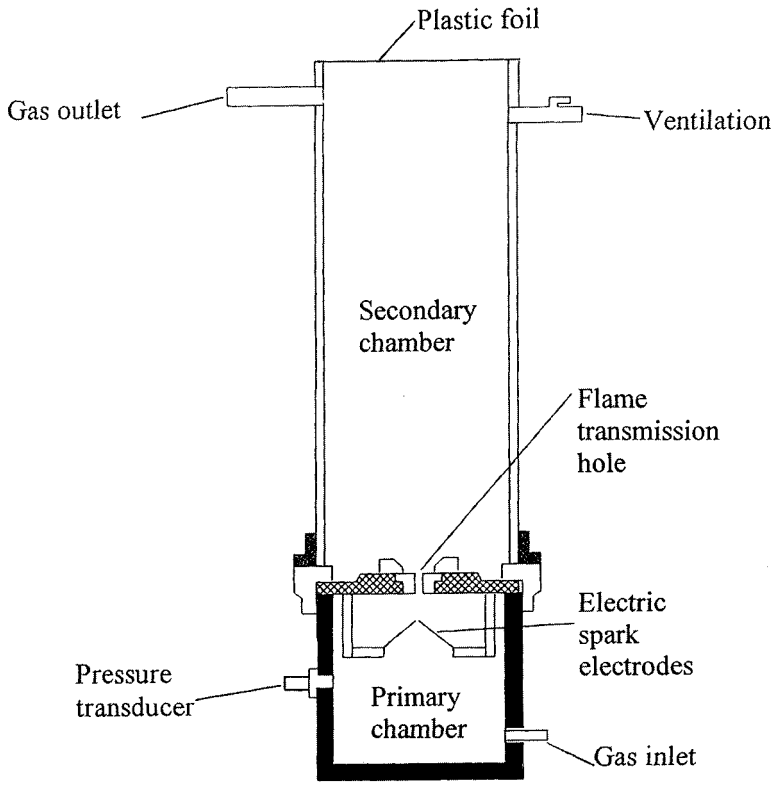


Figure 2–39 Cross section of apparatus used by Larsen and Eckhoff (2000) for experimental determination of critical hole diameters for transmission of gas explosions. The volume of the primary chamber was 1 liter.

ignition was in fact about half the critical diameter in laminar flame transmission. However, with a further increase of X_i the critical hole diameter for re-ignition increased far beyond the laminar flame transmission value. This is because, at large X_i values, the overpressures in the primary chamber reach high values at the moment of flame arrival at the hole entrance, and give rise to correspondingly high jet velocities. This in turn causes very fast mixing of the hot combustion products and the cold unburned mixture, and the whole system cools down before the combustion chemistry gets under way.

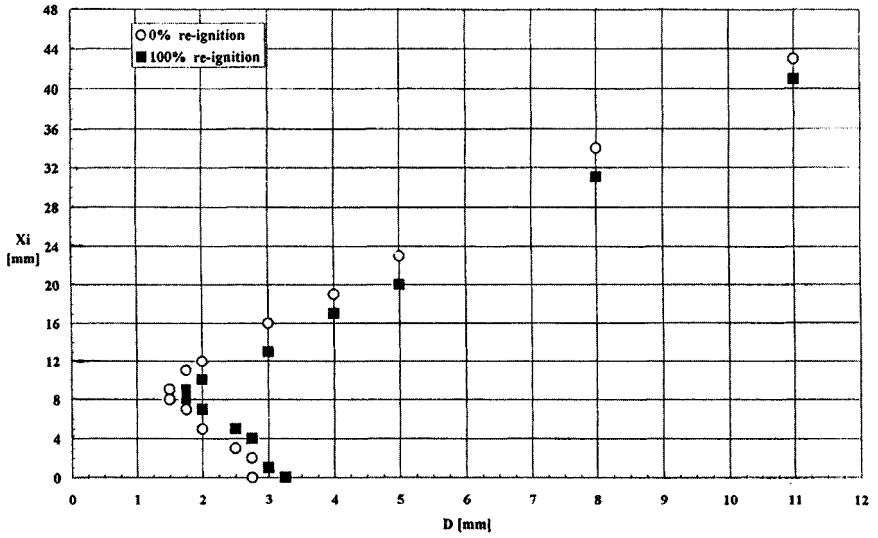
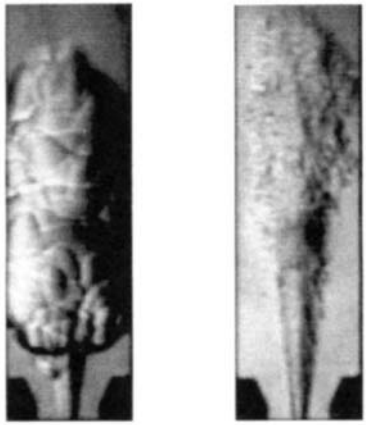


Figure 2–40 Results from experiments with a cylindrical 1 liter primary chamber (Figure 2–39). X_i is the distance from the electric spark ignition source on the chamber axis to the entrance of the axial cylindrical transmission hole of diameter D and length 12.5 mm. Explosive gas mixture: 4.2 vol.% propane in air. From Larsen and Eckhoff (2000).

Figure 2–41 shows two high-speed video pictures at critical conditions for re-ignition, using a smaller primary chamber than that illustrated in Figure 2–39. With X_i = 8 mm re-ignition occurred, with X_i = 9 mm there was no re-ignition. The characteristic difference between the two cases is the flame ball that appears in the downstream region of the jet in the case of re-ignition.

Results from some earlier similar studies by Wolfhard and Bruszak (1960) are given in Figure 2–42. In this case, the gas mixtures in the primary and secondary chambers were different. The gas in the primary chamber was stoichiometric mixtures of methane, oxygen, and nitrogen in all the experiments. The ratio of oxygen to nitrogen was varied and characterized by the oxygen index (OI) = (vol.% O_2)/(vol.O_2 + N_2). The purpose of the experiments was to determine the critical OI in the primary chamber for re-ignition in the secondary chamber. The three fuel gases used in the secondary chamber were methane, ethane and a mixture of 90 vol.% CO and 10 vol.% H_2, and they were mixed with air to stoichiometric compositions. The gap for the electric spark used for igniting the



Re-ignition No re-ignition

Figure 2–41 High speed video Schlieren pictures of hot-gas jets expelled from a 0.021 liter primary chamber into a larger secondary chamber via a straight cylindrical hole of diameter 4 mm and length 12.5 mm. 4.2 vol.% propane in air in both primary and secondary chamber. Distance from ignition point in primary chamber to hole entrance was 8 mm for re-ignition and 9 mm for no re-ignition. Pictures taken 6.25 ms after first sign of hot gas jet out of hole. From Larsen (1998).

gas in the primary chamber was located close to the entrance of the tube connecting the two chambers.

As Figure 2–42 shows, the OI required for re-ignition increased systematically with increasing tube length for tube lengths greater than 5–10 mm. The likely reason for this is cooling of the hot combustion gases by the tube wall, which will increase with increasing tube length. For very short tube lengths of a few mm, the critical OI started to rise again. It is also seen that the critical OI for ignition increases with decreasing tube diameter, as would be expected, and with the type of fuel, with methane being the most difficult and CO + H_2 the easiest to re-ignite.

Various apparatuses have been used for determining standardized MESG (maximum experimental safe gap) values of combustible gases and vapors mixed with air (see data in Table 2–2 and Figure 2–44). Normally the primary chamber is spherical, whereas the secondary one is an annulus surrounding the sphere. The connection between the two chambers is a flange gap of length 25 mm and adjustable width. The aim of the test is to determine the critical maximum gap width that does not give re-ignition in the second chamber (MESG), for the actual combustible gas or vapor being tested. The highest values are normally obtained with the

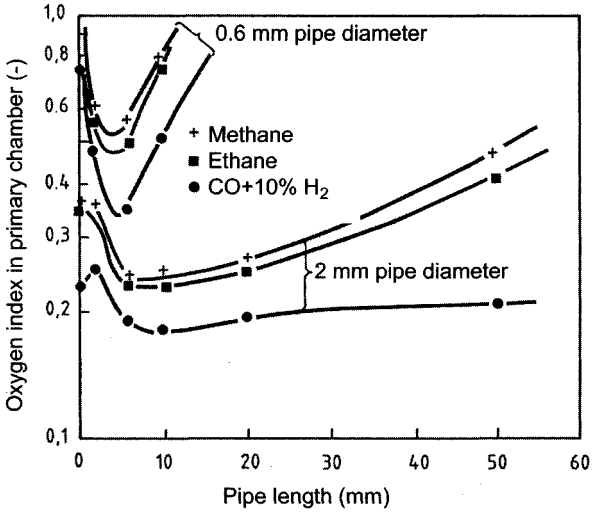


Figure 2–42 Critical oxygen index for ignition of an explosive gas mixture in a secondary chamber by a hot jet of combustion gases from a gas explosion in a primary chamber, as a function of the length of the cylindrical tube connecting the two chambers. Volume of primary chamber: 80 cm^3. Volume of secondary chamber: 3.8 liters. From Wolfhard and Bruszak (1960).

ignition source located fairly close to the gap entrance, rather than at the center of the primary chamber.

Experiments have confirmed that the volume of the primary chamber does not influence the measured MESG as long as this volume is at least 20 cm^3, and the volume of the secondary chamber is sufficiently large to prevent pre-compression of the gas there before the appearance of the hot jet. This is illustrated in Figure 2–43.

As can be seen no significant effect of the volume of the primary chamber was found over the range 0.020–8.0 liters. For this reason it was decided to adopt a primary chamber of 0.020 liter volume in the standard IEC (1971) test method. Figure 2–44 shows the standard apparatus developed by IEC (1975a). Prior to an experiment the width of the 25 mm long flanged gap is adjusted to the desired value by means of a spring-loaded micrometer screw. The ignition source in the internal chamber is a 3 mm spark gap located perpendicularly to the gap plane, 14 mm away from the gap entrance. The chambers are flushed with the gas/air mixture to be tested until homogeneous composition is obtained throughout. Whether re-ignition occurs is observed visually through the glass window in the

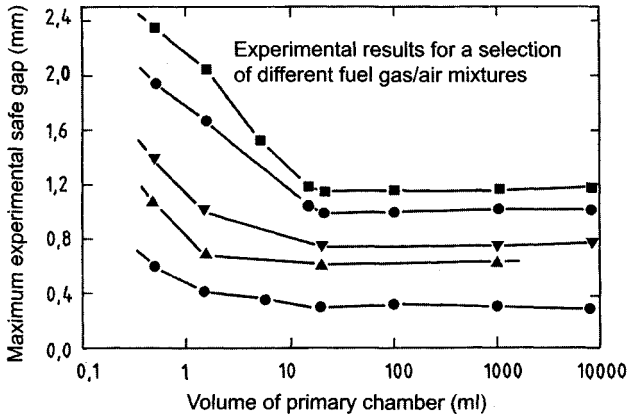


Figure 2–43 Influence of the volume of the primary chamber on MESG, determined experimentally by PTB, Germany. From Phillips (1987).

wall of the external chamber. The MESG data in Table 2–2 were determined by this method.

Figure 2–45 gives some experimental results from Maskow (1950) showing the effect on MESG of varying both the length of the gap from the standard 25 mm and downwards, and the fuel/air ratio.

This figure shows that the critical gap width for flame transmission (re-ignition downstream of gap) increased systematically with the gap length L, apart from a sharp edge (L = 0) being slightly less efficient than a short finite gap of length 1 mm. It is also seen that the fuel/air ratios that gave flame transmission most readily were close to the stoichiometric ratio.

2.2.8 Ignition by Rapid Adiabatic Compression

In the case of rapid adiabatic compression, much work has been performed in the context of internal combustion and diesel engines. Some useful information is given in Freytag (1965). If an explosive gas mixture is compressed rapidly adiabatically, or in a shock wave, the temperature will rise rapidly and the cloud may ignite. The temperature increase mainly depends on the ratio between the pressure after compression and the initial pressure. Shock waves may be generated during sudden relief

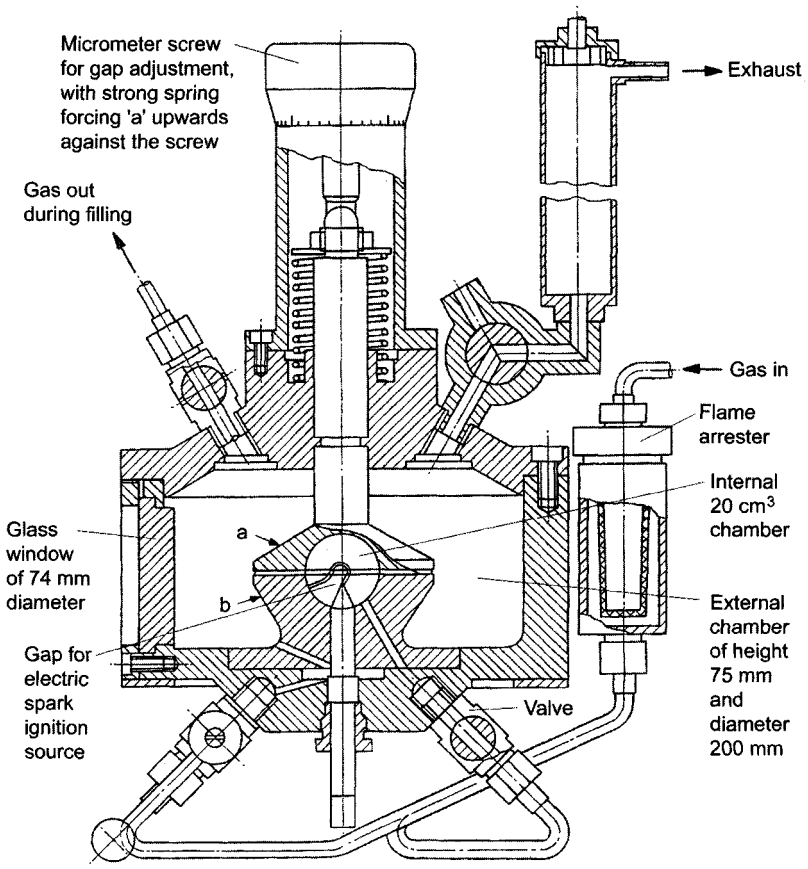


Figure 2–44 Illustration of the standard IEC apparatus used for determination of MESG for combustible gases and vapors mixed with air. Volume of primary spherical chamber is 20 cm^3. Standard gap length 25 mm. From IEC (1975a).

of high-pressure gases into pipelines. The shock wave then propagates into regions of lower pressure faster than the speed of sound. Very high peak pressures and temperatures can occur if the wave is diffracted or reflected by pipe bends, constrictions, connection flanges, closed valves etc.

Figure 2–46 shows pressure as a function of time during rapid compression of a mixture of hydrogen and air in a cylinder with a piston. The induction time between completion of compression and onset of ignition is a characteristic feature of this kind of process. Figure 2–47 and

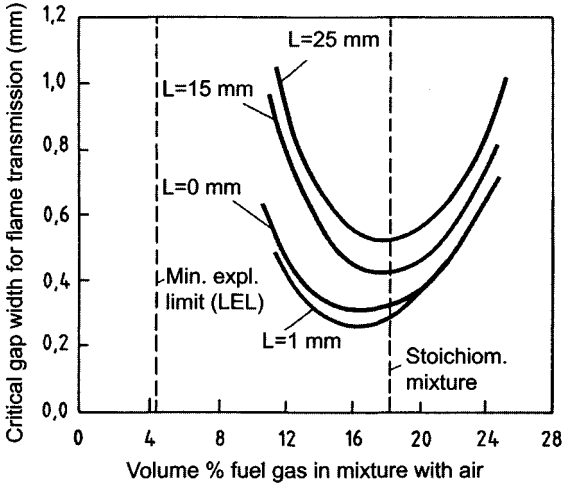


Figure 2–45 Maximum experimental safe gap, determined in a standard MESG test apparatus, for different gap lengths L and different fuel/air ratios. Fuel: 30 vol.% methane + 70 vol.% hydrogen. From Maskow (1950).

Figure 2–48 show experimental results for some hydrocarbons. The induction time depends both on the fuel type and the fuel/air ratio, and on the degree of compression (temperature of gas mixture just after compression has been completed). Fuel/air ratios of about half the stoichiometric composition seem to ignite most easily in this ignition mode.

A more recent, detailed study of ignition following rapid adiabatic compression of heptane/air mixtures, was undertaken by Minetti et al. (1995).

2.2.9 Ignition by Light Radiation

Some informative investigations were conducted by Welzel et al. (2000) and Thowle (2000). Welzel et al. studied ignition of a wide range of explosive gases and vapors mixed with air, by continuous optical radiation conveyed through an optical fiber and absorbed by a solid black iron/manganese oxide target mounted at the end of the fiber. The combustible gases and vapors covered a wide range of auto-ignition temperatures The experimental apparatus used is illustrated in Figure 2–49.

The ignition chamber containing the explosive gas mixture was a vertical glass tube of length 50 cm and diameter 15 cm, i.e. a volume of about 9 liters. The iron-oxide-covered end of the optical fiber was positioned in

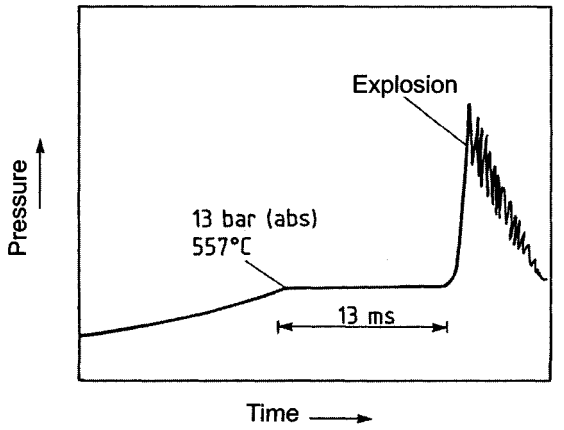


Figure 2–46 Pressure as a function of time during rapid compression of a mixture of hydrogen and air in a cylinder with piston. State of gas mixture at end of compression: 13 bar (abs) and 557°C. Induction time 13 ms. From Freytag (1965).

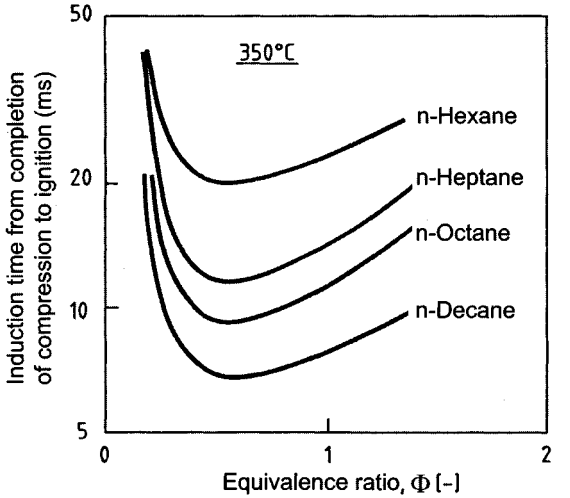


Figure 2–47 Induction time as a function of fuel/air ratio for some alkane/air mixtures compressed adiabatically to 350°C. (Φ = 1 corresponds to the stoichiometric ratio). From Freytag (1965).

the lower part of the chamber. The light source was a laser of wavelength 1064 nm. Figure 2–50 shows a selection of experimental results. The Figure also indicates conservative limits embracing all the data, viz.

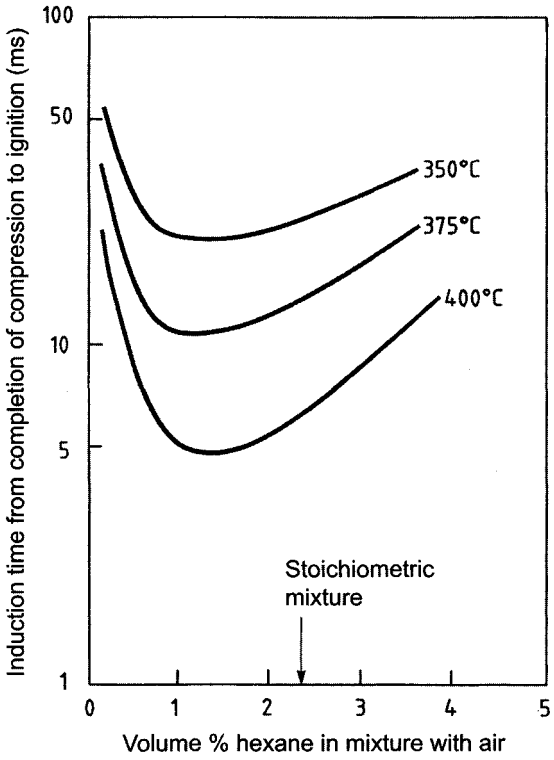


Figure 2–48 Induction time as a function of fuel/air ratio for hexane/air for various degrees of adiabatic compression. From Freytag (1965).

35 mW for target areas smaller than 4 mm^2, and 5 mW/mm^2 for larger target areas.

Figure 2–51 illustrates how the minimum radiant power required for ignition drops with increasing temperature of the explosive mixture.

Figure 2–52, from Thowle (2000), summarizes data from several independent investigations, including that of Welzel et al. (2000), and the data confirm the validity of the conservative asymptotes suggested by Welzel et al (2000). Note that the very low values in the lower right corner of Figure 2–52 were obtained by targets heated electrically, not by light radiation.

Correspondence of the absorption bands of the combustible gas with the laser or radioactive radiation wavelength may cause ignition by direct excitation and ionization of the gas molecules. Radioactive radiation is

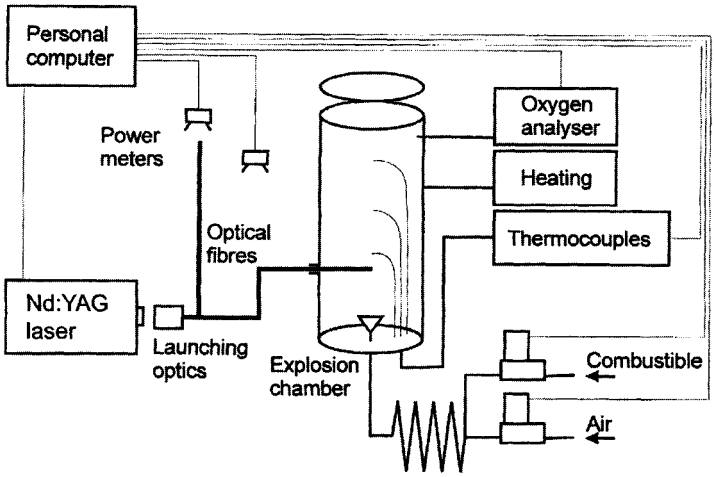


Figure 2–49 Experimental apparatus used for determining critical data for ignition of a wide range of explosive gases and vapors mixed with air by optical radiation absorbed by a solid iron/manganse oxide target attached to the end of an optical fiber. From Welzel et al. (2000).

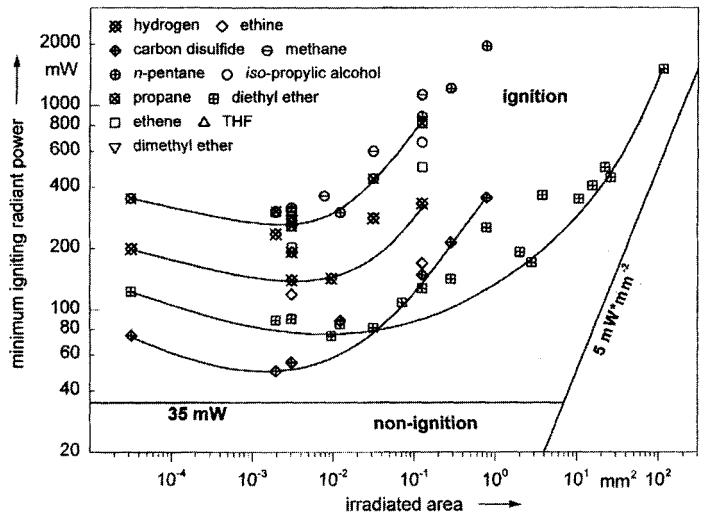


Figure 2–50 Minimum radiant power (mW) for ignition as a function of the surface area of the iron/manganese target (mm^2) for a range of explosive mixtures of combustible gases/vapors and air. From Welzel et al. (2000).

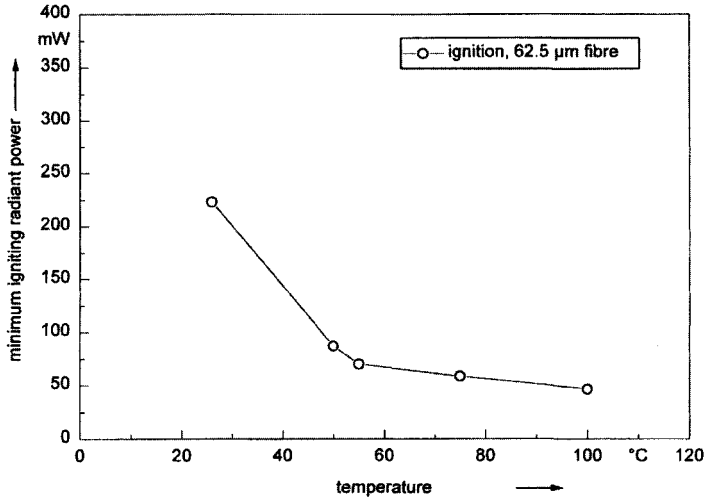


Figure 2–51 Minimum radiant power required for igniting a mixture of 12 vol.% diethyl ether in air, as a function of the temperature of the explosive mixture. The heated target was a thin layer of black iron/manganese oxide attached to an optical fiber of diameter 62.5 µm. From Welzel et al. (2000).

generated by X-ray tubes and radioactive substances. Radioactive radiation can heat up a solid surface, owing to internal absorption of radiation energy, to such an extent that the minimum ignition temperature of the surrounding explosive gas mixture is exceeded. Such radiation can also cause direct chemical decomposition of the gas or other chemical reactions which can generate highly reactive radicals or unstable chemical compounds which can lead to ignition.

2.2.10 Concluding Remark

Critical ignition parameters, e.g. the minimum ignition temperature and the minimum ignition energy, of a given explosive gas mixture can vary substantially with the actual ignition source characteristics, and the dynamics, pressure and temperature of the gas mixture. This means that there is a need for methods that permit differentiation when specifying critical parameters for various practical situations. For example, it is essential that the minimum ignition temperature corresponds as closely as possible to the particular industrial situation of concern. The application

Next Page

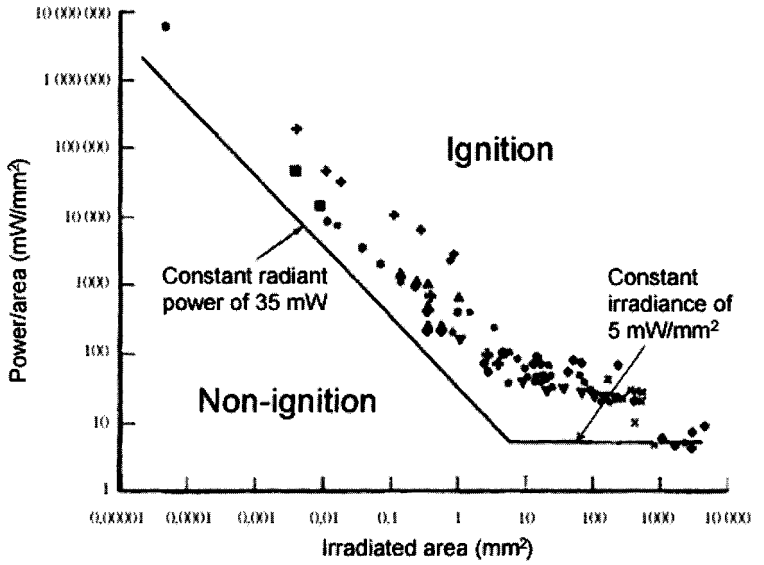


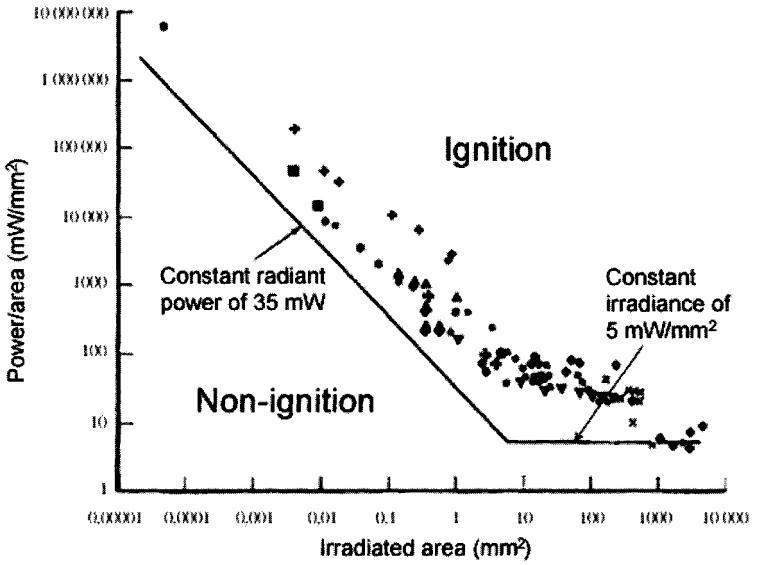
Figure 2–52 Summary of experimental data of minimum radiated power per unit target area (mW/mm²) for igniting a range of explosive mixtures of combustible gases/vapors and air, as a function of the irradiated area (mm²). From Thowle (2000).

of just one value for a given explosive gas, based on a highly conservative laboratory test, irrespective of the actual industrial situation, may in some cases put industry to considerable unnecessary expense.

2.3 Case Histories of Accidental Gas/Vapor Cloud Explosions

2.3.1 Motivation for Section

Experience has shown that "learning by doing" is an effective way of acquiring new knowledge. Unfortunately, this also applies to learning about explosions, which can give rise to much human suffering and grief, as well as material damage and loss of profit. People who have experienced an explosion accident, whether as workers or management in industrial plants, or elsewhere, have a profound appreciation of the realism of this hazard, beyond the reach of those who have only heard or



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read about accidental explosions in general terms. Real understanding in turn produces the proper motivation for minimizing the probability of occurrence of such events in the future. Clearly, accidental explosions are highly undesirable in any situation, and one therefore seeks less dramatic means of transferring knowledge and motivation. One way is the use of case histories, i.e., detailed accounts of explosions that have actually occurred.

This section describes a number of gas explosion accidents in different types of industry and equipment, with emphasis on ignition sources, propagation mechanisms, and destructive effects. Most of the material is taken from Eckhoff (1996a).

2.3.2 Historical Perspective: Methane Explosions in Coal Mines

As coal mining developed in Europe during the 17th and 18th centuries, severe mine explosions became common. It was soon discovered that the origin of the explosions was the ignition of mixtures of flammable gas, or *firedamp* and air, which accumulated in the mines. Firedamp is essentially methane liberated from coal when pressure is released. In early coal mining in Europe, testing of mines for possible explosive gas was undertaken by volunteers creeping into the mine galleries wrapped in wet blankets and carrying an open flame on a stick. This is illustrated in Figure 2–53.

In 1815 the UK's Sunderland Society for Preventing Accidents in Coal Mines asked Sir Humphrey Davy to carry out systematic research on the causes of ignition of and flame propagation in firedamp/air mixtures. As Davy recorded:

> The great object was to find a light, which at the same time that it enabled the miners to work with security in explosive atmospheres, should likewise consume the firedamp.

As pointed out by McQuaid (1990), one of the striking features of this investigation was the speed with which Davy moved from receiving the problem to producing a successful solution. However, when the task of finding a safe source of light in coal mines was first put to Davy, he was not optimistic. It appeared to him that there was very little hope of finding an effective solution. Earlier attempts at finding a safe source of light had

Figure 2–53 A fire fighter igniting methane under the roof of a mine gallery prior to the miners being admitted to the gallery. From McQuaid (1990).

not met with much success. Designs of safety lamps in existence relied either on complete confinement so lamps could burn only for a short time, or else the air supply was pumped through water seals so lamps were not practicable and the flame could easily be extinguished.

Despite his initial pessimism, Davy embarked on his research. He first investigated the chemical composition of the firedamp and then carried out numerous experiments on the circumstances under which it combusts. He established the limits of flammability of the firedamp in air and also established that it was much less easy to burn than other flammable gases. Under the test conditions used, it could neither be ignited by red hot charcoal nor by red hot iron. The heat produced by it in combustion was less than that of most other flammable gases.

Davy found that in glass tubes one quarter of an inch in diameter and one foot long, more than a second was required before the flame had traveled from one end to the other. When the tubes were one seventh of an inch in diameter, flammable mixtures could not propagate a flame to the open end. Metallic tubes prevented flame propagation better than glass tubes.

He now reasoned that the effect of the surfaces of small tubes in preventing combustion depended on their cooling powers, that is, upon their lowering the temperature of the burning mixture so much that it was no longer sufficient for continued combustion. He proceeded to conclude that a metallic tissue or gauze would offer the perfect barrier to flame propagation. He experimented until he found the geometry of the gauze and its material which was safe in all possible varieties of flammable mixtures, not only of firedamp but also of coal gas. His final design was produced in January 1816 and was immediately adopted in the coal mines. Figure 2–54 shows an early version of the Davy lamp.

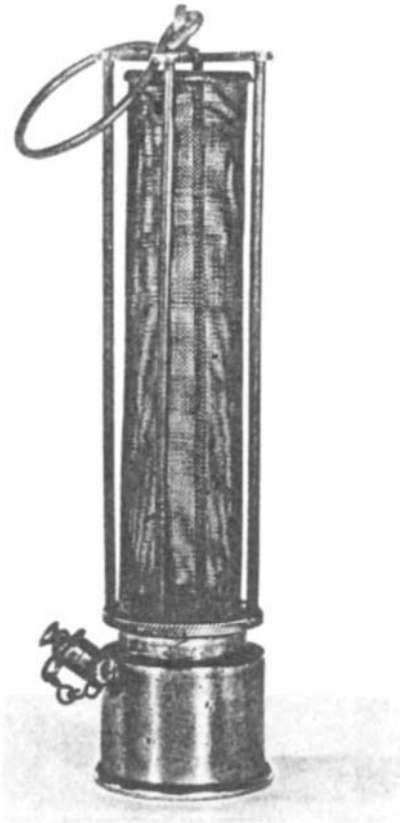


Figure 2–54 Early version of the coal mine lamp developed by Sir Humphrey Davy about 1815.

Davy had achieved his objective less than six months after receiving the letter from the Sunderland Society for Preventing Accidents in Coal Mines. And he did this without any of our modern analytical facilities. However, it seems reasonable to assume that this highly efficient and successful research was not only due to Davy himself. His young and extraordinarily talented assistant, whose name was Michael Faraday, must also have played a major role. Much later, in 1844, he was the first researcher to realize that also coal dust played a crucial part in many of the severe coal mine explosions that Europe experienced in the growing coal mining industry at that time.

The excellent performance of the Davy lamp was confirmed in a letter from Mr. George Morris, Plas Issa, Wales, dated 27 January 1817, to his mine owner (from McQuaid, 1990):

> Sir: You will be pleased to recollect that some time in the month of June last, I applied to you with a request you would send me immediately some of Sir Humphrey Davy's safety lamps, in consequence of an explosion of the firedamp taking place in one of your coal mines, by which several of the men were dreadfully burned and bruised. On the arrival of the safely lamps no accurate account of their use accompanied them. But I at length obtained (I think) the Edinburgh Review, in which was a detail of some experiments. This I read to the colliers, which gave them some confidence in the lamps, prior to which they secretly treated them with silent contempt; and I found, notwithstanding these interesting details, that a great doubt existed in their minds.
>
> The men had no sooner descended than the enemy (methane) was discovered (by ignition and combustion inside the Davy lamp), which they say very much alarmed them. They would have retreated if they could, but finding that impossible, took courage, and soon found they had destroyed the enemy so far; advancing a little farther, they found him again, and again destroyed him, and so on through the whole work. Thus the first alarm was got over, when all the knowing men in the neighbourhood were got collected together to hear the result, all of which were astonished and amazed, that so simple looking an instrument should destroy and defy an enemy, heretofore unconquerable. The same precaution and use of the lamp, was gone through the second day, and when the firedamp was destroyed, we began working and continued to work in this way for some weeks.

However, in spite of this and other great achievements, methane and coal dust explosions continued to represent a severe hazard in coal mines.

Cybulski (1975) summarized the major mine disasters that occurred world wide during 1900–1970. In 135 major explosions worldwide, more than 20,000 miners lost their lives; that is an average of about 150 in each of the disasters. Cybulski also described much of the extensive research that had been conducted in many countries up to about 1970, in order to reduce the explosion hazard in coal mines.

2.3.3 Previously Published Reviews of Major Accidental Gas/ Vapor Cloud Explosions

Gugan (1979) published a comprehensive review of one hundred major gas/vapor cloud explosions in the process industries, up to the late 1970s. One of the main sources of information was the preceding account by Davenport (1977). Gugan concluded that because of incomplete records, only 29 of the 100 incidents could be analyzed in any depth. Of these, only eight of the incidents that gave significant blast effects were documented sufficiently well to permit reasonably comprehensive analyses. Gugan (1979) gave a fairly detailed discussion of five of the eight explosions.

In his book, Marshall (1987) presented 11 reviews and collections of case histories of major gas/vapor cloud explosions up to 1984, including the book by Gugan (1979) and the updated report by Davenport (1984). Marshall gave quite detailed case histories of the major di-methylether explosion at Ludwigshafen, Germany in 1948; the propane explosion at Port Hudson, Missouri, USA in 1970; the cyclo-hexane explosion in Flixborough, UK in 1974; and the iso-butane explosion at Decatur, Illinois, USA in 1974.

Gow (1991) provided brief abstracts of published records of vapor/gas cloud explosions, including the explosions with propane (Sweden, 1981); vinyl chloride (USA, 1971); acetylenic alcohol (USA, 1967); ethanol (USA, 1972);1-pentol (USA, 1967); hydrogen-rich treat gas (UK, 1983); methanol synthesis gas (USA, 1983); natural gas (USA, 1986); hydrogen (USA, 1989); a mixture of ethane, propane, ethylene, and propylene (USA, 1967); ethylene oxide (USA, 1968); butadiene (USA, 1971); propylene (USA, 1976); propylene (USA, 1977); and hydrogen (Norway, 1989).

2.3.4 The Flixborough Explosion, UK (1974)

2.3.4.1 Summary

Some central data of this catastrophic event are given in Table 2–13.

Table 2–13 The Flixborough Explosion

Place:	Flixborough, Humberside, England
Date:	June 1, 1974
Fuel:	Cyclohexane
Type of plant:	Congested process area of cyclohexane oxidation section of caprolactam production plant (basic raw material for Nylon 6)
Fatalities:	28
Injuries:	89
Material Loss:	Entire plant demolished. 1821 houses and 167 shops in nearby residential areas suffered various degrees of damage

The Flixborough disaster is probably the most well documented major accidental industrial vapor cloud explosion world wide. The comprehensive public report by Parker et al. (1975) contains references to about sixty detailed reports on various aspects that were available by early spring of 1975. A further considerable number of articles, scientific papers and reports were published after the public report had been issued. Examples are the paper by Warner (1975) and the papers by Sadee et al. (1976/1977), Roberts and Pritchard (1982), and Berg (1985). Extensive reviews and analyses of the Flixborough disaster, containing further references to papers and reports, were given by Gugan (1979) (12 pages); Lees (1980) (20 pages); and Marshall (1987) (20 pages). Marshall includes an interesting analysis of blast damage of dwelling houses in the surroundings of the industrial plant, as a function of the assumed explosion epicenter.

Roberts and Pritchard (1982) estimated the maximum explosion loads at various locations in the Flixborough explosion, on the basis of the degree of deformation and damage on process equipment, lamp posts and other mechanical structures.

Lees (1994) provided a detailed account of the various aspects of the Flixborough disaster, a summary of which follows.

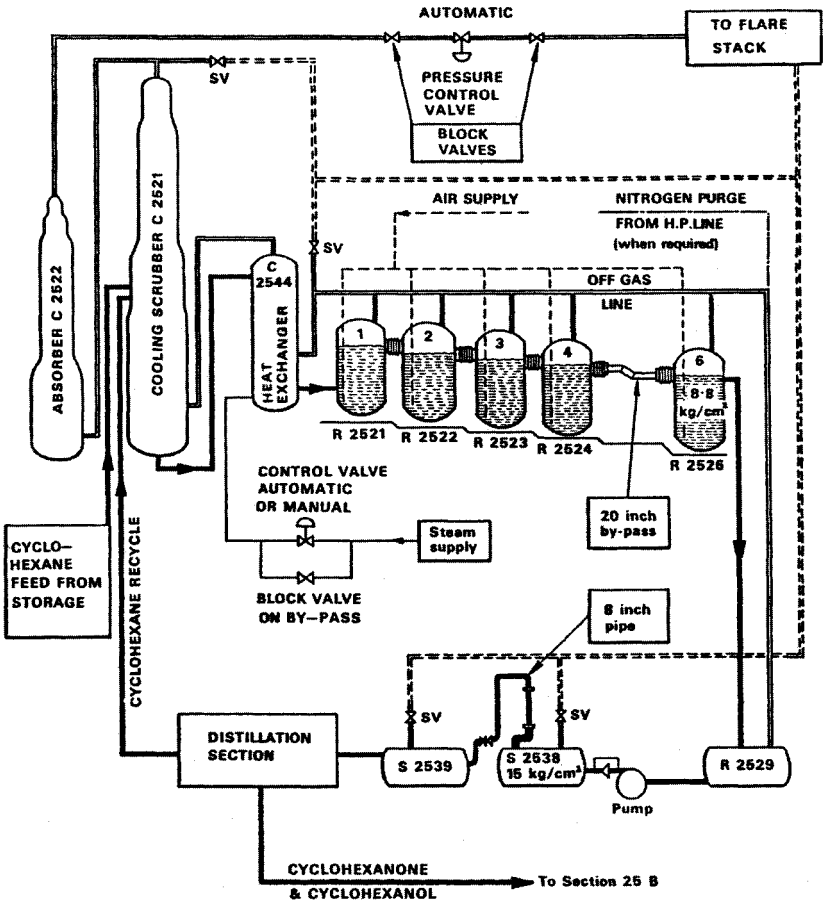
2.3.4.2 The Process and the Plant

The cyclo-hexane oxidation plant shown in Figure 2–55 consisted of a train of six reactors in series in which cyclo-hexane was oxidized to cyclo-hexanone and cyclo-hexanol by air injection in the presence of a catalyst. The reactions are exothermic. The feed to the reactors was a mixture of fresh cyclo-hexane and recycled material. The product from the reactors still contained approximately 94 percent of cyclo-hexane. The liquid reactants flowed from one reactor to the next by gravity. In subsequent stages, the reaction product was distilled to separate the unreacted cyclo-hexane, which was recycled to the reactors, and the cyclo-hexanone and cyclo-hexanol, which were converted to caprolactam. The operating conditions in the reactors according to the design were 8.8 bar(g) and 155°C.

2.3.4.3 Events Prior to the Explosion

On the evening of 27 March 1974, it was discovered that Reactor No. 5 was leaking cyclo-hexane. The following morning, the decision was taken to remove Reactor No. 5 and to install a bypass assembly to connect Reactors No. 4 and 6 so that the plant could continue in production. The diameter of the openings to be connected on these reactors were 28", with bellows on the nozzle stubs, but the largest pipe which was available on site, which was considered to be suitable for making the bypass, had a diameter of 20". Furthermore, the two flanges were at different heights so that the connection had to take the form of a dog-leg of three lengths of 20" pipe welded together with flanges at each end, which were bolted to the existing flanges on the stub pipes on the reactors. The bypass assembly is indicated in Figure 2–55. Calculations were done to check that the pipe was large enough for the required flow and that it was capable of withstanding the same pressure as if it were a straight pipe.

However, as pointed out by Lees (1994), no drawing of the bypass pipe was made other than in chalk on the workshop floor. The existing stub pipes were connected to the reactors by bellows, as indicated in Figure 2–55. No calculations were done to check whether the bellows would withstand the forces caused by the dog-leg pipe. The bypass assembly was supported by a scaffolding structure, which was intended to support the pipe and to avoid straining of the bellows during construction of the bypass. It was not suitable as a permanent support for the bypass assembly during normal operation. No pressure testing was carried out



CYCLOHEXANE OXIDATION
PLANT SECTION 25 A
Simplified flow diagram (not to scale)

Cyclohexane loop	SV	=	Safety valve
Off gas	R	=	Reactor
Safety valve vent	S	=	Separator
Nitrogen (when required)	C	=	Column
Air (not operating at time of disaster).			

Figure 2–55 Simplified illustration of section 25 A of the cyclo-hexane oxidation plant of the caprolactam production plant at Flixborough. From Parker at al. (1975).

either on the pipe or on the complete assembly before it was fitted. A pressure test was performed on the plant, however, after installation of the bypass. The equipment was tested to a pressure of 9 bar(g), but not up to the safety valve pressure of 11 bar(g). The test was pneumatic, not hydraulic.

Following these modifications, the plant was started up again. The bypass assembly gave no trouble. There did appear, however, to be an unusually large usage of nitrogen on the plant, and this was being investigated at the time of the explosion. On 29 May, the bottom isolating valve on a sight glass on one of the reactors was found to be leaking. It was decided to shut the plant down to repair the leak. On the morning of 1 June, start up began. The precise sequence of events is complex and uncertain. The crucial feature, however, is that the reactors were subjected to a pressure somewhat greater than the normal operating pressure of 8.8 bar(g). A sudden rise in pressure occurred early in the morning when the temperature in Reactor No. 1 was still only 110 °C and that in the other reactors was less. Later in the morning, when the temperature in the reactors was closer to the normal operating value, the pressure reached 9.1–9.2 bar(g). The control of pressure in the reactors could normally be performed by venting the off-gas, but this procedure involved the loss of considerable quantities of nitrogen. Shortly after warm-up began, it was found that there was insufficient nitrogen to begin oxidation and that further supplies would not arrive until after midnight. Under these circumstances, the need to conserve nitrogen probably inhibited an attempt to reduce pressure by venting.

2.3.4.4 The Explosion

During the late afternoon of the day of the explosion, the provisional 20" bypass system ruptured, without or with a contribution from a fire on a nearby 8" pipe. This resulted in the escape of a very large quantity of cyclo-hexane. The cyclo-hexane formed a vapor cloud (and possibly also a liquid-droplet spray cloud, see Section 3.5.1), and the flammable part of the cloud found a source of ignition. At about 4.53 p.m. there was a huge, catastrophic vapor cloud explosion.

The explosion caused extensive damage and started numerous fires. The blast and the fires destroyed not only the cyclo-hexane plant but several other plants also. Many of the fires were in the tank farm. The blast of the explosion shattered the windows of the control room and caused the

control room roof to collapse. Of the twenty-eight people who died in the explosion, eighteen were in the control room. Some of the bodies had suffered severe injuries from flying glass. Other victims were crushed by the roof. No one escaped from the control room. The main office block was also demolished by the blast of the explosion. However, since the accident occurred on a Saturday afternoon, the offices were not occupied. If they had been, the death toll would have been much higher. The fires on the site burned for many days. Even after ten days, the fires were hindering rescue work on the site.

2.3.4.5 The Investigation

Some of the possible causes of failure of the bypass assembly were outlined early in the inquiry. They included failure of the 20" diameter pipe due to a small pressure rise in the system. Several possible causes for a pressure rise in the 20" diameter pipe were suggested, viz. entry of high pressure nitrogen into the system due to instrument malfunction, entry of water into the system, temperature rise in the system due to excessive heating by steam in re-boiler of the heat exchanger C2544, leakage of steam from a tube in C2544, explosion of peroxides formed in the process, and explosion due to air in the system.

The overall conclusion of the final inquiry was that the disaster was due to "ignition and rapid acceleration of the resulting deflagration, possibly to the point of detonation, of a massive vapor cloud formed by the escape of cyclo-hexane under a pressure of at least 8.8 bar(g) and a temperature of 155 °C, and that the escape was from Section 25A of the cyclo-hexane plant." There was no dispute that the main part of the cyclo-hexane came from the 20" bypass assembly, but there was some uncertainly as to whether the mechanical failure of the assembly was the primary failure or whether it was a secondary failure caused by a preceding one.

2.3.5 The Beek Explosion, the Netherlands (1975)

2.3.5.1 Summary

Some central data of this event are given in Table 2–14.

Table 2-14 The Beek Explosion

Place:	Beek, The Netherlands
Date:	November 7, 1975
Fuel:	C3-C4 hydrocarbon fractions
Type of Plant:	Congested process area of naphta cracker installation
Fatalities:	14
Injuries:	104
Material Loss:	Entire plant and a nearby tank farm destroyed. Window breakage in surroundings up to 2.5 km from explosion centre.

Reviews of this accident were given by Gugan (1979) and Lees (1994). A comprehensive two-part report of investigation in the Dutch language (Anonym, 1976) was issued by the Directorate of Labour of the Ministry of Social Affairs, in the Netherlands. This report contains quite detailed maps and photographs of the damaged plant, including a map indicating the contours of the explosive cloud prior to ignition.

2.3.5.2 Process and Plant

According to Wingerden et al. (1995), the explosion occurred in a naphtha cracker installation, covering an area of approximately 160 m × 80 m. Naphtha was cracked in furnaces, resulting in hydrogen, ethylene, and other hydrocarbons, ethylene being the main product. Separation of the products was accomplished by distillation, compression, and cooling. The plant was subdivided in seven process sections, viz. the furnaces for cracking, various separation and distillation processes, a compression section, and cooling sections. Transport of raw materials, and products between the various parts of equipment within sections, and between sections, occurred in pipes gathered in pipe-racks. Most process sections were densely packed with equipment. In some, the equipment was located on elevated grated, or concrete floors.

2.3.5.3 Events Prior to the Explosion

In the morning on the day of the explosion, a leakage of a mixture of C3 and C4 hydrocarbons occurred from one of the process sections. The most likely cause was low-temperature embrittlement at a weld in a feed drum.

The prevailing wind caused the flammable vapor cloud to drift into the congested process area. After approximately two minutes from the start of the leak, the visible part of the cloud was approximately 2 m high. It was estimated that in total approximately 5,500 kg of gas had escaped. At this point, the cloud was ignited by one of the furnaces near one corner of the process area.

2.3.5.4 The Explosion

The resulting explosion caused severe damage inside the process area. Overpressures of up to 1 bar must have existed to account for the damage produced. Pipelines and beams were bent over and several pipelines had ruptured, resulting in fires. Vessels had become dented and pushed away from their original locations. Part of a concrete platform had fallen down and another had tilted. Severe damage occurred to the concrete walls of the control room. The engineers' room was totally demolished. Fire broke out in a tank farm north of the process installation. Outside the plant area, damage was limited to window breakage up to a distance of 4.5 km (5 mbar). The fourteen fatalities occurred in the control room and engineers' room.

2.3.5.5 Computer Simulation

Twenty years after the accident, Wingerden et al. (1995) were able to perform a series of unique computer simulations of the Beek explosion disaster, using the comprehensive CFD-based simulation code FLACS, developed in Norway from 1980. A comparison of local peak pressures estimated from the damage picture generated by the actual vapor cloud explosion in 1975, and the corresponding local peak pressures produced by the FLACS simulation, showed similar trends in the peak pressure distribution across the process area. The simulations indicated that the explosion occurred with a fuel that was more reactive than pure propylene. Significant amounts of ethylene or butadiene must also have been present. The simulations also showed that for land-based process installations in the open, where partial confinement is mainly due to process equipment, the reactivity of the fuel is important with respect to the violence of an accidental explosion. The simulations further revealed that in an exploding heavy (flat) vapor cloud, a stoichiometric mixture of fuel and air is not necessarily the worst case. Over-stoichiometric mixtures can be

even more dangerous due to the rapid mixing of air above the cloud with the gas mixture in the cloud during the explosion process.

2.3.6 The Arendal Explosion, Gothenburg, Sweden (1981)

2.3.6.1 Summary

Some central data of this event are given in Table 2–15

2.3.6.2 The Site of the Event

Details of this accident have been obtained from the Special Working Group (1981), Brandsjö (1988), and Nilsson (1991). The site of this explosion is not, strictly speaking, a process plant, but the event provides valuable information that is relevant even in the present context. Figure 2–56 gives a view of the explosion site and indications of the local damage caused by the explosion. The site was a large obstacle field in the open characterized by pipe tracks, bridges, buildings, cars, and trees, acting as turbulence-generating obstacles.

Table 2–15 The Arendal Explosion

Place:	Arendal, Gothenburg, Sweden
Date:	May 8, 1981
Fuel:	Propane
Type of Plant:	Pipe track through urban area. Rupture of pipeline
Fatalities:	1
Injuries:	1
Material Loss:	15 mill. SEK

2.3.6.3 Leak and Vapor Cloud Formation

The gas leak giving rise to a large explosive gas cloud was probably caused by sabotage. A 5 km long 6 inch diameter pipeline, transporting liquid propane at 8-9 bar(g), became perforated, presumably by an explosive charge placed and detonated intentionally on the pipeline wall. About 95 m^3 of the liquid propane was flashed to the atmosphere through the

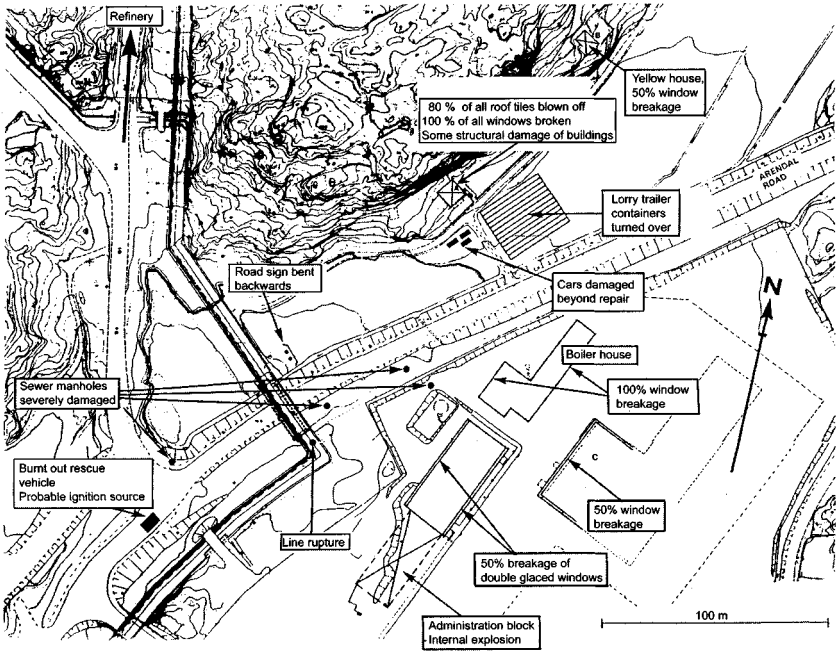


Figure 2–56 Plan of site and consequences of propane vapor cloud explosion at Arendal, Gothenburg, Sweden, in 1981. From Nilsson, 1991.

resulting 90 cm^2 hole in the pipe wall. The gaseous propane mixed with the ambient air, and a major explosive cloud was generated in the surroundings. The report of the Special Working Group (1981) provides estimates not only of the leak rate of propane from the perforated pipeline, but also of the total amount that escaped to the atmosphere. Visual observations of the size and shape of the vapor cloud prior to ignition are also included. However, Nilsson (1991) pointed out that the visible part of the cloud was due to water droplets generated by condensation due to cooling of the air by the propane evaporation, and that the extension of the real propane gas cloud could not be observed visually.

2.3.6.4 Ignition and Explosion

Ignition of the cloud occurred about fifty-five minutes after the pipeline had been perforated. The ignition source was probably a fire appliance

that was driven into the explosive cloud. There was a series of distinguishable explosions, extending over about thirty seconds. The major fireball had a diameter of 300 m. Maximum explosion overpressures at various locations, e.g. in the sewer manholes indicated in Figure 2–56, were estimated from damage analyses.

2.3.7 Methane Explosion in 17,000 m^3 Coal Silo at Elkford, British Columbia, Canada (1982)

Storage of coal can present a gas explosion risk, due to spontaneous release of methane from some types of coal. An account of such an explosion in British Columbia, Canada, in 1982 was given by Stokes (1986).

2.3.7.1 Plant and Process

The silo that exploded, of height 48 m, diameter 21 m, and capacity of 15,000 tons, was used for storage and load-out of cleaned, dried metallurgical coal. Prior to the explosion, a methane detector had been installed in the roof of the silo. The detector activated a warning light in the silo control room when a methane concentration of 1% was detected, and an alarm light was activated when detecting 2% methane. A wet scrubber was located in the silo head house to remove dust from the dust-laden air in the silo during silo loading. A natural ventilation methane stack was also located in the silo roof to vent any build-up of methane gas from the silo.

The silo was full of coal twenty-four hours prior to the explosion. However, during the evening before the explosion, 10,000 tons of coal was discharged, at the same time as conveying of deep-seem coal into the silo had been started. This continued until the explosion occurred. At the moment of the explosion, there was approximately 12,300 tons of coal in the silo, of which 7,600 tons was deep-seam coal. Testing had shown that this quality of coal had a high methane emission rate and produced a low volatile coal dust. Clouds in air of this dust could not be ignited unless the air was mixed with methane.

2.3.7.2 The Explosion and Its Consequences

The explosion occurred early one morning in May. Witnesses stated that a flash was noticed in the vicinity of the head house, followed seconds later by an explosion displacing the silo top structures. This was followed by an orange-colored fireball that rolled down the silo walls and extinguished prior to reaching the base of the silo. Fortunately, neither injury nor death resulted, and damage to surrounding structures was minimal, although large blocks of concrete and reinforcing steel had been thrown several hundred meters from the silo. However, the plant itself had suffered substantial damage, including a devastated silo roof, head house, and conveyor handling system.

2.3.7.3 Possible Ignition Source

The ignition source was not identified, but three possible sources were considered: spontaneous combustion of the stored coal, an electrical or mechanical source, and hot coal from the thermal dryer. Spontaneous combustion had never presented a problem during ten years of operation, with coal being stored in different environments for varying lengths of time. Consequently spontaneous combustion was not considered to be a probable source of ignition. During demolition of the damaged silo, all electrical and mechanical components were recovered and inspected and did not show any evidence of having been the ignition source. Therefore, Stokes (1986) concluded that the remaining possibility, i.e. hot coal from the thermal dryer, was the most probable source of ignition.

2.3.8 The "West Vanguard" Explosion, The North Sea (1985)

2.3.8.1 Summary

Some central data of this event are given in Table 2–16.

2.3.8.2 Site of Explosion

A detailed discussion of this accident is given in the report by the Public Commission of Investigation (1986). The explosion occurred on board

Table 2-16 The "West Vanguard" Explosion

Place:	Haltenbanken, Norwegian continental shelf
Date:	October 6, 1985
Fuel:	Natural gas
Type of Plant:	Mobile drilling platform
Fatalities:	1
Injuries:	0
Material Loss:	Several hundred mill. NOK

the comparatively new, well equipped and well maintained mobile oil drilling platform "West Vanguard", while performing test drilling on Haltenbanken on the Norwegian continental shelf in the North Sea. The main platform structure is illustrated in Figure 2–57.

2.3.8.3 Events Leading to the Explosion

An uncontrolled blow-out of natural gas occurred during the early stages of a normal drilling operation. The blow-out was a so-called "shallow-gas kick" and happened before the blow-out preventing valve had been mounted. Shallow gas pockets were expected in this area, but at greater depths than the 263 m below the sea bed that had been reached when the blow-out occurred. According to the Public Commission of Investigation (1986), a series of unfortunate circumstances led to the uncontrolled blow-out, including inadequate reporting, partial lack of confidence in the process monitoring instruments, deficiencies in the well program, and reluctance to increase the mud weight. The gas pressure in the shallow pocket was of the order of 50 bar, and the velocity of the uncontrolled upwards gas flow was very high. The gas diverter equipment was unable to withstand the heavy erosion by the sand and other solid particles in the flowing gas. Consequently, after a few minutes, the gas flow had perforated the equipment and started to flood the entire platform structure. The gas was effectively distributed to most parts of the platform via the ventilation system. Eventually it reached an ignition source and got ignited.

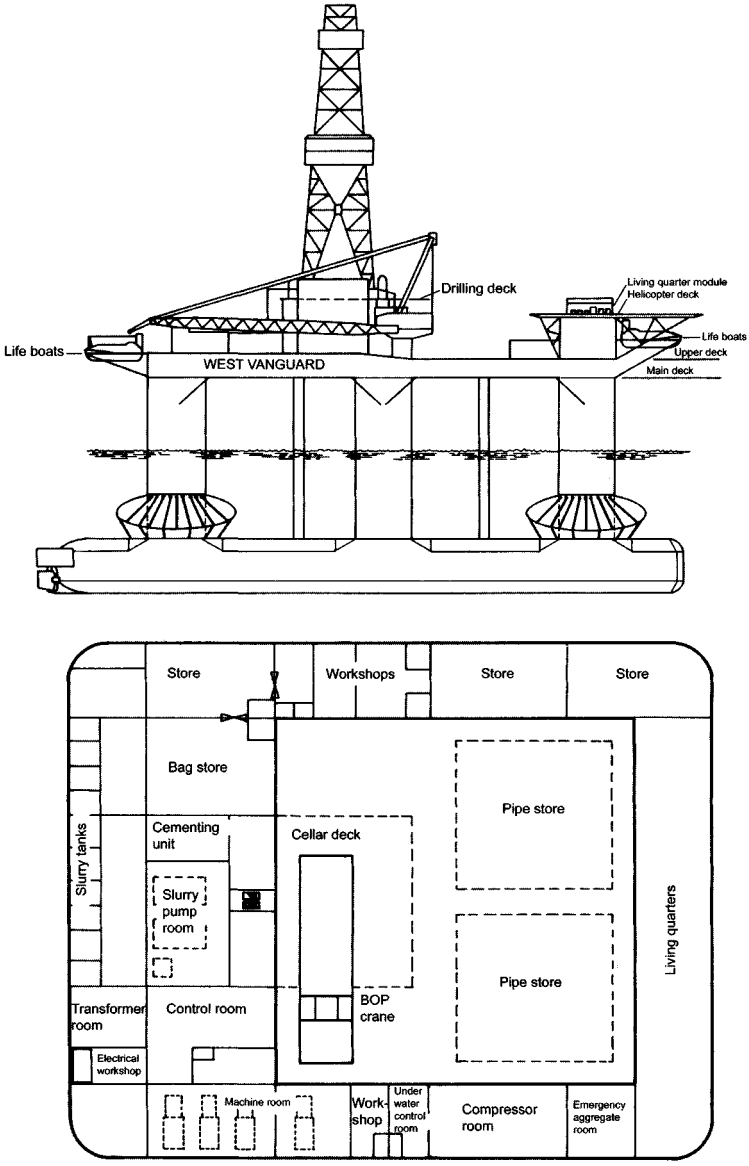


Figure 2–57 Illustration of the mobile drilling platform 'West Vanguard', with a side view of the entire platform, and a top view of the main deck. The main deck is 85 m long and 75 m wide. The platform was kept floating by means of two 100 m long pontoons. From Public Commission of Investigation (1986).

2.3.8.4 Ignition Sources

The ignition sources were probably metal sparks or a hot surface generated in the cellar deck area during the intense erosion of metal during the blow-out, and some electrical or other ignition source in the engine room of the platform. The former source was considered the most likely possibility for the first, strong explosion.

2.3.8.5 The Explosions and Their Consequences

At least two distinct explosions were heard. The first and most powerful produced a large flameball in the area of the drilling tower and across the pipe deck. According to most witnesses, this explosion probably occurred in the cellar deck. The second major explosion probably occurred in the engine room about one minute after the first explosion. Fortunately, as soon as the blow-out got out of control, and only a few minutes before the first explosion, the rescue operation had started, and 79 of the 80 persons on board were able to get into the lifeboats from which they were picked up by the standby ship before the explosion occurred. Large amounts of natural gas were bubbling through the sea to the sea surface, and from the stand-by ship it was observed that natural gas was burning across the sea surface in some areas around the platform, but fortunately not in the area from which the people were rescued. The platform suffered severe damage from the explosions and subsequent fires, mainly in the after part, including the area around the drilling tower and the adjacent areas and rooms. Roofs, bulkheads and bolted steel doors were blown up or heavily deformed by the explosion pressures.

2.3.8.6 Computer Simulations of Gas Explosions

As part the investigations after this accident, numerical simulations of explosions in various parts of the platform, were performed by means of the FLACS code developed in Norway. This made it possible to compare the observed actual structural damage with that to be expected from various ignition scenarios. This work was discussed in more detail by Bjorkhaug et al. (1985).

2.3.9 Catastrophic Gas Explosion in Taegu, South Korea, April 1995

2.3.9.1 Overview

This gas explosion catastrophe did not occur in the process industry, but in the middle of the busy city Taegu in South Korea (see Figure 2–58). However, the lessons to be learned are also indeed relevant for the process industries. The following provisional summary is taken from the brief report by Spaeth (1995) published shortly after the accident:

> At 0750 hours on April 28, 1995, an immense fireball tore through the heart of Taegu, South Korea's third-largest city, of a population of 2.3 million. Cars were hurled aloft, chunks of steel beams landed atop buildings and 250-kg steel plates that had covered a subway excavation, rained down on vehicles and pedestrians. A bus from the Youngnam Middle School pitched into the hole, killing all the students aboard.

Figure 2–58 Site of the LPG explosion disaster in Taegu, South Korea, April 28, 1995, which killed 100 people, and injured 117. Ten buildings were completely demolished and 48 others partly damaged, while 80 vehicles were destroyed or damaged. From Fire Investigation Report (1995).

The descriptions of the cause of the catastrophe in magazines and newspapers shortly after the accident were somewhat diverging. According to Spaeth (1995):

> The explosion occurred when a drill operator working at a department store building site near the subway construction pit mistakenly drilled into a fuel pipeline. The liquid (LPG) poured out into the pit for about 20 minutes, vaporized and the resulting vapor cloud was ignited, probably by a workers blowtorch.

Korea Times, a few days after the explosion, reported that:

> A 100 mm diameter gas pipeline was ruptured during digging near a department store adjacent to the explosion site. The pipeline was buried 1.3 meters deep under the road. Gas leaking from the ruptured pipe flew into the subway construction site through a sewage line and got ignited by a spark from nearby welding work. Subsequent investigations disclosed a 7 cm hole in the gas pipe.

However, the comprehensive Fire Investigation Report (1995) concludes that the exact cause of the explosion disaster could not be found. The following is a brief summary of the findings given in this report:

2.3.9.2 Circumstances before the Explosion

The residents living near the construction site had, since the night of 27 April, smelled gas several times and reported this to the City Gas Corporation and nearby police station, but the responsible persons had not responded to the reports. As late as 20 minutes before the explosion in the morning of 28 April, workers on site again reported to the Corporation about the gas leakage, but no immediate action was taken. It later turned out that after receiving similar messages from the workers at the site about a gas leaks both in January and March, the City Gas Corporation had not taken any significant measures. Fortunately, in these two cases there were no explosions.

2.3.9.3 Possible Causes of Explosion

The exact causes of the explosion were not resolved, but three scenarios were suggested.

2.3.9.3.1 Scenario 1

Gas leaked from a loosened joint or a crack on one of the LPG pipelines passing through the construction site, and the gas cloud produced became ignited and exploded. The construction company in charge insisted that, before the explosion occurred, they had reported several times to the City Gas Corporation about the smell of gas on the site, without any actions being taken. There were indications of poor quality of the work done when the pipelines were first laid.

2.3.9.3.2 Scenario 2

Generation of the explosive gas cloud was due to workers on the site having damaged the underground gas pipelines during their works. The City Gas Corporation received a report from a worker on the site saying that gas was leaking due to their mistakes during excavation work just before the explosion. In this scenario, the gas cloud was probably ignited by sparks or embers from welding operations. However, the company in charge of the whole subway construction strongly rebutted this hypothesis by claiming that the explosion had been initiated early in the morning before the work on the site had started. Furthermore, the gas pipelines were hanging about 15 m above the base of the working site, just below the perforated steel plates acting as a provisional road cover across the site. Therefore, it would not have been easy for the workers to touch the pipelines accidentally.

2.3.9.3.3 Scenario 3

Another possible cause of the gas leak might be formation of one or more cracks in the pipelines due to the continual vibrations induced by the heavy traffic across the perforated steel plates acting as the provisional road across the construction cavity.

2.3.9.4 The Extent of the Explosion and Its Consequences

Besides the violent noise of the explosion, a huge flameball extending 50 m above the road level was observed. Numerous perforated steel plates used as provisional road cover for a road length of about 300 m

were flung into the air. One of these, weighing 280 kg (270 cm × 70 cm × 2 cm), was found hanging on a church cross on top of a three-story building near the explosion site. The explosion severely deformed and displaced the heavy duty H-beams supporting the provisional road cover within 1 km of the explosion centre. This caused numerous other steel plates to collapse under the load of all the vehicles, and thirty cars and buses were plunged into the 18 m deep construction cavity. Other cars soared to 10 m above road level. In all, eighty vehicles were damaged by the explosion. Ten buildings were completely demolished and forty-eight others partly damaged. Gas pipelines, water pipes, and underground cable pits were also destroyed, and the whole area was flooded, and the electrical power was cut off in thousands of homes. Rescuers found 97 people killed, three people missing, and 117 people injured.

2.3.9.5 Rescue Operations

A total of thirty fire trucks, ten cranes, and about 2,500 rescue personnel were dispatched to the explosion site. However, because of the possibility of further explosions of remaining unburned gas in the underground, rescue personnel could not be admitted to the site immediately after the explosion. Flooding due to the broken water pipes delayed the rescue operations even more. Rescue operations could not start fully until about twelve hours after the explosion.

2.3.9.6 Lessons to be Learned

The overall cause of this catastrophe was an inadequate governmental bid system for construction work, including inadequate quality assurance of technical and managerial safety matters. The workers on the site did not have adequate understanding of the technical standards applying to the work they were doing, and the responsible authorities had not given sufficient guidance. The company in charge of the construction work had not checked the underground pipelines drawings prior to starting the excavation works. Furthermore, the construction site was not provided with any gas alarm system. The investigations also disclosed lack of updated maps of underground gas pipelines, electric cables etc.

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2.3.9.7 The "Human Factor"

The following story, also from Spaeth (1995), illustrates the important part played by human behavior, not only in the chain of events leading to accidents, but indeed also in the subsequent investigation to disclose what actually happened:

> Five days after the explosion a local fire fighter told reporters that the gas leak had been reported to the local fire station by a street sweeper four hours before the explosion. According to the fire fighter, the complaint was entered in the station's log, but destroyed after the accident had occurred. When journalists found the sweeper, he said he reported a gas smell to a police box at 2100 hours in the evening before the blast and again at 0400 hours the following morning. After the tragedy he was confronted by officials who threatened to have him fired from his job unless he repudiated his reports. He did so, and the retraction was videotaped. When his story was made public, however, the prosecutor general in Seoul ordered an investigation.

2.4 Means of Preventing and Mitigating Gas/Vapor Explosions in the Process Industries

2.4.1 Introduction

Explosions and fires involving combustible gases and vapors constitute a major hazard in process industries and other environments where such materials are produced, used and handled. Therefore, the efforts to minimize the risk of explosions and fires in these industries continue nationally as well as internationally, and much work is spent on preventing and mitigating accidental gas and vapor cloud explosions.

Explosion risk is often defined as the product of the probability of an explosion and its expected consequence. Therefore the basic principle of gas explosion risk management is to minimize the explosion probability as well as the expected explosion loads, which are in turn related to the explosion consequences. Reduced gas explosion consequences can be obtained by active as well as passive measures. The most important means of prevention and mitigation/control are summarized in Figure 2–59.

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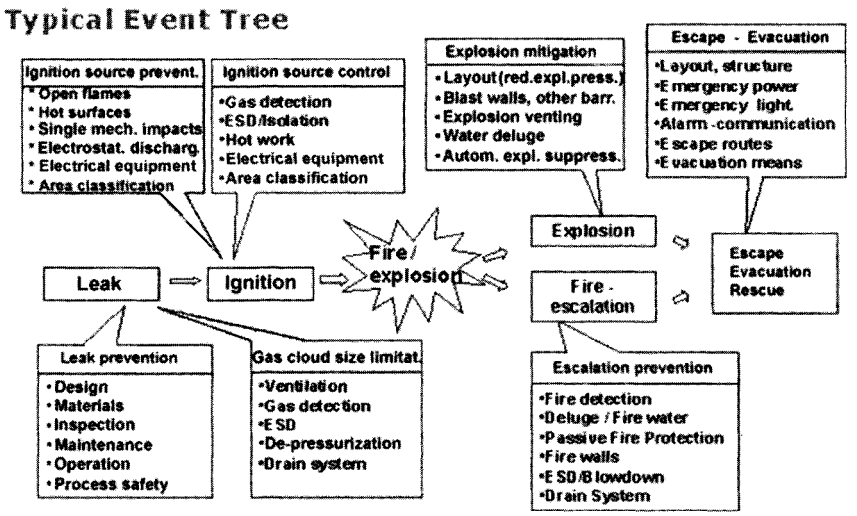


Figure 2–59 Event tree illustrating the various means for preventing and mitigating/ controlling gas cloud explosions in the process industries.

The following overall aims have high priority in gas explosion risk management:

- prevention and control of potential leaks of combustible gases
- good ventilation to minimize extent and duration of explosive atmosphere due to gas relases (reduces ignition probability)
- prevention and control of possible ignition sources
- minimizing high equipment congestion/blockage to ease explosion venting, and to minimize flow-induced turbulence during gas cloud formation and explosion
- installation of blast and fire barriers

2.4.2 Preventing and Limiting Size of Explosive Gas/Vapor Clouds

2.4.2.1 Preventing Gas Leaks from Process Equipment

The following is based on Statoil (2004):

2.4.2.1.1 Piping and Vessels

Piping, pipelines, tanks, heaters, mechanical equipment, vessels design integrity and related inspection, maintenance, and operation activities shall ensure that leaks of hydrocarbon fluids (and chemicals and toxic gases) do not occur. This is accomplished by the following measures:

- the inspection system and execution shall verify that the chosen monitoring parameters are within specified limits
- a tailored condition monitoring strategy shall be established for any given plant
- a working process for establishing inspection programs shall be available
- inspection acceptance criteria shall be defined
- systems for reporting and follow-up of findings shall be established
- blow-down systems to be designed with a particular focus on pressure transients, surge and vibration

2.4.2.1.2 Flanges and Connections

The design, activities of operation, inspection and maintenance of flanges and other mechanical connections shall ensure that hydrocarbon leaks do not occur. The amount of flanged connections in systems containing hydrocarbon fluids shall be kept at a minimum by observing the following:

- prior to undertaking constructive modifications of the plant the possibility of using welded connections as opposed to flanges shall be investigated, based on a coherent safety philosophy. For example, any mechanical connections (e.g. flanges) on the pipeline side of risers on offshore installations should be avoided. If such connections are chosen, appropriate protection must be applied
- mechanical connections shall be designed against vibrations
- procedures for assembling flanges and other connections, including bolt tensioning, must be developed

- compact-type flanges are preferred to the ANSI type due to a lower predicted leak frequency
- de-pressurization (blow-down) pipes should normally be designed without flanges

2.4.2.1.3 Instrumentation, Valves, and Rotating Machinery

The design, activities of operation, inspection and maintenance of in-stream instrumentation, valves, and rotating machinery (e.g. pumps and compressors) shall ensure that hydrocarbon leaks do not occur. A maintenance program shall cover all equipment that could be a potential gas/vapor leak source.

2.4.2.2 Minimizing Size of Gas Cloud in Case of Accidental Leaks by Early Leak Detection and Effective Shut-Down

The following is based on Statoil (2004):

2.4.2.2.1 Overall Objective and Performance Criteria of Gas Detection Systems

Detectors for flammable gas shall be installed in areas (risk volumes) where explosion loads from ignited gas clouds could cause unacceptable damage, i.e. loss of safety functions. In these areas, the smallest gas cloud that has the potential of causing such damage shall be specified as the detection target. The gas detection function shall provide reliable and fast detection of leaks before any accidentally released gas cloud reaches a concentration and size, which could cause significant risk to personnel and plant. The detector number and distribution in the process area of concern shall be based on realistic worst-case gas leakage scenarios (source locations and leakage rates) within the process area of concern by considering the dispersion mechanism transforming the leak jet into a gas cloud. Due to the complex influence of the geometry and arrangement of process equipment environmental conditions such as ventilation, numerical gas dispersion simulations, using e.g. the FLACS code, may have to be performed for optimization of location of detectors within specific areas.

2.4.2.2.2 Detector Types

For hydrocarbon gases, infrared detectors shall be the preferred type whenever suited for the gas to be detected. Optical beam detectors with long range shall be used to increase detection probability of small leaks. Catalytic detectors should only be used if proper detection performance by other types can not be achieved. Maximum response time of the gas detection function shall be defined in order to ensure that total reaction time for each safety function can be fulfilled. Typical response times are as follows:

- infrared detectors: < 5 s for general area applications
- catalytic detectors: < 10 s from low alarm (20% LEL) to high alarm (30% LEL)
- acoustic detectors: < 30 s including delays to improve false alarm immunity

2.4.2.2.3 Coverage and Location of Gas Detectors

The following overall principles shall be applied with respect to location of detectors:

Detectors shall be spread out evenly, but not symmetrically, to ensure adequate coverage of long "corridors" in the main ventilation flow direction. Natural flow "corridors," e.g. walkways along expected flow direction, should be covered.

Detectors shall not be located in zones with no or little ventilation (dead zones) as gas concentration rise is slower in these areas as compared to areas with higher ventilation rates. Examples of dead zones are corners, between large roof beams and congested areas without leak sources.

In process areas of considerable height, e.g. modules on offshore platforms, detectors should be positioned in at least two levels.

Crucial gas detector locations include:

- zone 1 and zone 2 areas (see zone definitions in Chapter 7)
- inlets of heat, ventilation, and air conditioning systems (HVAC)
- ventilation outlets from classified areas

- combustion air intakes and turbine hoods
- rooms that contain potential ignition sources and is equipped with doors, trapdoors, hatches etc. towards classified areas (see Chapter 7)
- enclosed areas where gas clouds can enter or be formed
- rooms containing potential ignition sources
- battery rooms if hydrogen can be formed in dangerous concentrations
- gas turbine hoods

Gas detectors in ventilation intakes shall ensure that a possible gas cloud is detected and dampers are closed before a dangerous concentration can reach the ventilated areas. In the case of hydrocarbons, use of infrared type detectors located as close as possible to the ventilation intake allows actions to be initiated within short period of time, typically a few seconds.

2.4.2.2.4 Gas Detection Alarms

The gas detection system shall automatically activate pre-determined actions in accordance with the actual plant-safety-strategy-and-philosophy documentation. Each individual gas detector shall generate alarm upon gas exposure. The explosive gas concentrations at which alarms are activated are typically in the range 20–30% of LEL. Alarms shall be given as quickly as possible to warn and guide personnel upon detection of gas. Detection of gas, failure to execute actions upon demand, and system defects shall be presented as alarms in central control room.

Other issues of concern related to gas detection systems include:

- systems independence
- systems operation integrity
- systems survivability
- systems inspection, maintenance and modification
- man-machine interface
- documentation

2.4.2.2.5 Spray/Mist Detectors

In the case of release of droplets of flammable liquids that evaporate slowly after being released to the atmosphere, suitable detectors shall be located close to the floor of the process area, e.g. 0.5 m above the deck on offshore platforms.

Oil mist detectors (optical obscuration) should be installed in machinery spaces and turbine exhaust channels, indicating the release of high pressure flammable liquids such as fuel or lubrication oil that may lead to aggressive and damaging fires.

2.4.2.3 Emergency Shut-Down Systems (ESD)

The following is based on Statoil (2004).

2.4.2.3.1 Purpose of ESD Systems

The purpose of emergency shutdown systems at large is to prevent escalation of abnormal conditions into major hazardous events and to limit the extent and duration of any such conditions. In the present context this means to limit any accidental releases of explosive gas, vapor, and mist from process equipment, to levels that do not give rise to hazardous explosions and fires.

2.4.2.3.2 Manual ESD Activation

Manual activation of the ESD system shall be possible from strategic positions along escape routes and at stations where accessibility and manning in a hazard situation is adequate. Manual activation stations shall be provided in the central control room, and in strategic locations in the process areas depending on plant arrangement and applied safety strategy. Manual activation buttons shall be protected against inadvertent activation, e.g. by protective covers or rotary switches. Each post shall be clearly marked, and a consistent color coding shall be applied for the activation buttons.

2.4.2.3.3 ESD Actions and Levels

Once initiated, all required ESD actions shall be automatically activated. The ESD actions shall be arranged in a tree-structured level hierarchy according to extent of actions and plant process facilities. All ESD levels and actions shall be described in detail within the documentation for operation of the plant. A superior ESD level will initiate lower levels. The following automatic actions shall be executed:

- shut-down of oil and gas wells (platform and sub-sea) and import/export pipelines including stop of export from upstream installations
- shut-down and sectioning of hydrocarbon process facilities
- initiation of de-pressurization of pressurized process plant ("blow-down") at higher ESD levels
- ignition source isolation/deactivation (see Section 2.4.4)
- shut-down of main power generation
- start/stop of emergency power generator
- shut-down of drilling and work-over equipment not required for oil/gas well control

2.4.2.3.4 ESD Alarms

Effective alarms shall be given as quickly as possible to warn personnel upon ESD level initiation. Activation of main ESD levels, failure to execute actions upon demand and system defects shall be presented as alarms in the central control room. An ESD shut-down action shall, in addition to general plant alarm, also be announced at other strategic locations, e.g. by visual display of ESD alarms in local control rooms. The ESD system status shall be continuously available in the central control room, and the system shall raise alarms there for operator awareness or action.

2.4.2.3.5 ESD Response Time Rate

The required response characteristics of the various elements of the ESD function loop shall be quantified to ensure that the total reaction time for each safety function is sufficiently short. The actual response times/rates of any specific ESD action to achieve safe state, shall be clearly identified

in relevant documentation. An example is that the time from activation to start of de-energizing of a solenoid valve shall be less than two seconds.

2.4.2.3.6 ESD Independence, Operating Integrity, Reliability, and Survivability

The following requirements shall be satisfied:

- The ESD system shall operate as an independent system.
- The classified ESD safety instrumented functions shall be functionally and physically segregated from other systems/functions.
- ESD functions shall be fully operational whenever the plant is "live" with combustible process liquids or gases, and appropriate testing of the system shall be possible without interrupting operation.
- Reliability and safety unavailability criteria for ESD functions shall be documented and maintained throughout the life cycle of the ESD system.
- The logic solver and essential utilities shall be located in a protected area.
- Equipment that is critical for the effectuation of system actions shall be protected against accidental loads such as explosion, fire and falling loads.

2.4.2.3.7 ESD Fail-To-Safe Principle

In the event of malfunction, the entire ESD-system or affected parts of it shall automatically go to a predefined safe state. If this *fail-to-safe* principle is not applied, an equivalent level of safety shall be achieved by redundancy, diagnostics, and alarms to control room.

2.4.2.4 ESD Man-Machine Interface

The man-machine interface in the central control room shall provide the means for operator awareness and actions and be suitable during emergency situations. The ESD status overview should be clearly displayed in the central control room at all times. All man-machine-interface facilities

shall be based upon an established philosophy, and the presentation of the system shall be clear and concise. Shutdown alarms and critical system failures shall be presented in proper sequence/priority. The hard-wired safety panel shall provide the means for manual activation and status observation of relevant actions, e.g.:

- plant alarm
- de-pressurization
- ignition source isolation/de-activation at relevant levels
- fire water pump start
- fire extinguishing systems status

As regards documentation and modification the following points must be observed:

- Documentation describing the ESD system shall be available in a formal updated revision that reflects current design and operation.
- Relevant documentation for operation and maintenance shall be available and updated.
- Modifications and changes to the ESD system shall be made in accordance with appropriate written procedures for initiation, documentation, review and approval.

2.4.2.5 Fast Dilution with Air of Gas Leaks by Natural Ventilation and Forced Heating, Ventilation and Air Conditioning Systems (HVAC)

The following is based on Statoil (2004).

2.4.2.5.1 Overview

The purpose of natural ventilation and HVAC systems is to prevent transformation of accidental leaks into major hazardous gas clouds by limiting the extent and duration of any explosive clouds that may occur. Natural ventilation shall:

- dilute the escaped gas by means of air and hence reduce the size of explosive gas clouds
- reduce harmful concentrations of smoke or toxic gases
- ensure acceptable working and equipment environment

HVAC shall:

- dilute the escaped gas by means of air and hence reduce the size of explosive gas clouds
- provide pressurisation of rooms to prevent ingress of smoke or gas
- provide smoke extract ventilation in case of internal fire
- ensure acceptable working and equipment environment

2.4.2.5.2 Natural Ventilation in Classified Areas

The natural ventilation system shall for ninety-five percent of the time provide a minimum specified average ventilation rate throughout the area to avoid stagnant zones. The following points should be considered:

- location and sizes for ventilation louvers (e.g. in walls of offshore platform modules) shall be optimized to give required minimum ventilation air change rates for dilution of flammable gases but shall in addition comply with safety risk analysis (gas dispersion, explosion loads etc.) and maintain acceptable working environment/ weather protection for personnel.
- natural ventilation shall be documented by calculations and/or model testing. Potential stagnant zones shall be evaluated and precautions taken where considered necessary. During operation, compliance with design documentation shall be verified, i.e. has additional equipment been installed, has louvers been changed or blocked etc.

2.4.2.5.3 Forced/Mechanical Ventilation in Classified Areas

For definition of classified areas see Chapter 7. The following points must be observed:

- A ventilation rate to satisfy specified minimum requirements shall be provided throughout the area to avoid stagnant zones.
- Alarm shall be given upon too low airflow.
- In the event of an internal gas leak in special systems such as turbine enclosures, battery rooms, gas analyser houses, etc. all available ventilation capacity should be used to maintain the "unclassified" status.
- There shall be a reliable overpressure of minimum $5 \cdot 10^{-4}$ bar in unclassified rooms relative to surrounding classified areas. Alarm shall be given upon too low overpressure.
- It must be possible to perform a manual disconnection of ignition sources in rooms with self-closing doors towards classified areas.
- Rooms without such doors, but which are next to classified areas shall be treated as rooms with doors. Access from non-classified to classified areas should be avoided.
- Fire/smoke/gas "dampers" shall provide quick, reliable and effective prevention of spread of gas, fires and smoke into other areas via the HVAC system. Gas tight fire dampers shall be installed in HVAC inlets. Dampers shall close automatically within six seconds upon gas and/or smoke detection in the air intake. Dampers and fans shall be interlocked to avoid abnormal pressure configurations. Damper position shall be monitored from the central control room.
- Ventilation functions and equipment shall be fully operational whenever the plant is live with flammable gases/liquids, and appropriate testing of the function shall be possible without interrupting operations.
- Reliability and safety unavailability criteria for the HVAC system shall be documented and maintained through the life-cycle of the system. In case of malfunction the HVAC system or affected parts of it shall go to a predefined safe state. The HVAC system and its components shall be designed and protected to ensure that they will remain operative during incidents where the system has a role as, or forms, a safety barrier.

2.4.2.5.4 Documentation and Modification

The following points must be observed:

- Documentation describing the ventilation system shall be available in a formal updated revision that reflects current design and operation.
- Relevant documentation for operation and maintenance shall be available and updated.
- Safety requirements and philosophies shall be adhered to, and safety assessment shall be performed prior to changes affecting system safety functions.

2.4.3 Preventing Ignition Sources

2.4.3.1 Introduction

The various types of potential ignition sources and their properties are discussed in Section 2.2. Chapter 7 discusses requirements to ensure that *electrical* equipment does not become an ignition source if making contact with explosive gas. Chapter 7 also discusses the concept of "hazardous area classification," which implies that the various regions/areas of the industrial plant are grouped in three categories depending on the likelihood that explosive gas may exist there. Regions where this likelihood is regarded as negligible are denoted "un-classified" or "non-classified".

Electrical as well as non-electrical equipment installed and used in locations where an explosive gas atmosphere may occur shall comply with the requirements corresponding to the hazardous area class of the area in which the equipment is to be used. According to the new European Union regulations, equipment—both electrical and non-electrical—used in hazardous areas shall provide a minimum level of explosion protection as follows:

- Zone 0: Certified equipment of category 1
- Zone 1: Certified equipment of category 2
- Zone 2: Certified equipment, or equipment suitable for use, of category 3

In general, specifications for electrical equipment to be used in areas where an explosive gas atmosphere will or may exist are to be in accordance with recognized international/European (IEC/EN) or other relevant standards/norms. For non-electrical equipment specifications should be based on recognized European (EN) or other relevant standards/norms.

2.4.3.2 Open Flames and Hot Exhaust Gases

The use of open flames, for operational or any other purpose, should be prohibited in all areas where explosive gas atmospheres may occur. Certain devices containing flames may be permitted if the flames are safely enclosed and if the temperature of the external enclosure surface does not exceed the prescribed maximum value for the gas in question. Furthermore, it should be ensured that the enclosure is adequately resistant to the effect of the flame and that flame propagation into classified areas cannot occur.

Hot exhaust gases may be introduced into areas that may contain explosive gas atmospheres if suitable devices are available to guarantee that the temperature of the exhaust gases at the inlet to the area is sufficiently low for ignition of the explosive atmosphere to be impossible.

2.4.3.3 Hot Surfaces

The standard rule is that the temperature of hot surfaces such as exhaust pipes and channels shall not exceed the Auto-Ignition Temperatures (AIT) of flammable fluids that can be present due to accidental leaks. (As discussed in Section 2.2.4, this may in some cases be a very conservative requirement.) Any thermal insulation materials shall cover the entire hot surface. Necessary means of protection shall be provided to prevent flammable fluids from penetrating into the insulating material.

In addition to easily recognizable hot surfaces such as heating coils, mechanical processes can also lead to hazardous temperatures. These processes include equipment which converts mechanical energy into heat, i.e. all kinds of friction clutches and mechanically operating brakes. Furthermore, all moving parts in bearings, shaft passages, glands, etc. can become sources of ignition if they are not adequately lubricated. In tight housings of moving parts, the ingress of foreign bodies or shifting of the

axis can also lead to friction which in turn can lead to high surface temperatures, in some cases quite rapidly.

The temperature of any surface in an area where an explosive atmosphere is likely to be present during normal operation should never exceed eighty percent of the minimum ignition temperature in °C of a combustible gas or liquid, not even in the case of rare equipment malfunctions. In areas where explosive atmospheres are likely to occur only in abnormal situations, the temperatures of all apparatus surfaces that can come into contact with the explosive gas mixture should be lower than the minimum ignition temperature.

2.4.3.4 Burning Metal Particles, "Thermite" Reactions, and Transient Hot Spots

Grinding and cutting operations must be prohibited in areas where explosive atmospheres may occur. Hot spots and spark showers from such operations may ignite explosive gas atmospheres, including natural gas/air.

Impacts involving rust and light metals (e.g. aluminum and magnesium) and their alloys can generate thermite flashes that can cause ignition of a wide range of mixtures of combustible gas or vapor and air, e.g. natural gas/air. Titanium and zirconium can form incendiary metal particle sparks by impact or friction against any sufficiently hard material, e.g. concrete.

Hot metal beads that are produced during welding or cutting operations may be regarded as metal sparks with large surface areas. They are effective ignition sources, e.g. for natural gas.

In areas where an explosive atmosphere is likely to occur under normal operation, apparatus and equipment which, even in the case of rare operational malfunctions, can give rise to incendiary friction, impact or abrasion sparks should be excluded. The possibility of impacts between aluminum (excluding alloys with less than 10% aluminum and paints and coatings with less than 25% aluminum by weight) and steel (except stainless steel, when the presence of rust particles can be excluded) must be prevented. Friction and impact between titanium or zirconium and any hard material must be avoided. It is possible in some cases to protect light metals from mechanical contact with rust by a coating. If coated with non-conductive materials such as plastics, precautions against static electricity may be necessary (risk of propagating brisk discharge).

The use of various metal tools for maintenance and repair work may generate metal particle sparks, and the possibility of these as potential ignition sources must be considered. In the case of natural gas, sparks from conventional steel hand tools are unlikely to cause ignition (see Table 2–9).

2.4.3.5 Ignition by Electric Sparks and Arcs and Electrostatic Discharges

All *electrical apparatuses* to be used in areas where explosive gas atmospheres may occur, must be manufactured, installed and operated in such a way that the likelihood of such apparatuses igniting an explosive gas atmosphere is acceptably low. Design of electrical apparatuses for areas containing explosive atmospheres is discussed separately in Chapter 7.

The design of electrical apparatus for use in potentially explosive gas atmospheres so as to prevent ignition of the explosive atmospheres by electric sparks from such apparatus is subject to standardized requirements. This also is discussed in Chapter 7.

Lüttgens and Glor (1989) and Lüttgens and Wilson (1997) give extensive information on fundamental aspects and practical implications for process safety of *electrostatic discharges*.

Objects of conducting materials that may collect an electric charge shall be connected to earth or bonded according to prevailing standards and guidelines. Objects shall be protected against tribo-electric charging (due to physical contact followed by separation of electrically different materials). Such charging can also take place during rapid flow of electrically non-conducting liquids, and gases containing liquid droplets of solid particles in pipes. But charge can also be produced in high-velocity water jets. Particular precautions will be required if poorly conducting construction materials are applied, whre a connection to earth may not provide adequate protection. Electrostatic charging and generation of incendiary electrostatic discharges may be a problem in situations including:

- fuelling operations
- filling of containers, tanks and pressure vessels
- high fluid velocities (high water sprays, gas jets)
- steam cleaning
- loading operations

Stray currents flowing in electrically conductive systems or parts of systems can give rise to spark or arcs. Such currents can appear:

- as return currents in power-generating systems, e.g. in the vicinity of large welding systems, for example when conductive electrical system components such as pipes and cable sheathing laid underground lower the resistance of the return current path
- as a result of any short-circuit, or of a short-circuit to earth, owing to faults in the electrical installations
- as a result of lightning
- as a result of magnetic induction, e.g. near electrical installations operating at high currents or high frequencies

If parts of a system able to carry stray currents are disconnected or connected/bridged, even in the case of slight potential differences, electric sparks or arcs can be formed, which may be able to ignite an explosive gas cloud. Moreover, ignition can also result from the heating of the current paths.

When impressed-current *cathodic corrosion protection* is used, this ignition risk has to be expected and special protective measures should be provided. However, if *sacrificial anodes* are used, ignition risks due to electric sparks do not normally have to be accounted for.

The possibility of *compensating for the electric potential* should be considered for all conductive parts of the installation, including those that are not adjacent to electrical equipment. It is permissible to deviate from this requirement within areas enclosed by conductive walls that are included in a potential compensation system. Before introducing new conductive parts, e.g. ventilation and suction pipes in tanks, such parts should be included in a potential compensation system.

For conductive parts of the system that are not adjacent to electric installations, it is permissible to dispense with special measures to compensate the potential difference when a compensation system already exists, formed by interconnected parts of the electrically conductive systems, e.g. pipe networks or extensive earthing systems.

Before the connections of conductive parts of the system are opened or closed, e.g. during dismantling of fittings and parts of pipes, bridges comprising connection lines with an adequate cross-section should be pro-

vided if it is possible that the adequacy of the interconnection might be impaired.

Ignition will always occur if *lightning* strikes an explosive gas cloud. Moreover, there is also a possibility of ignition due to the high temperature reached by lightning conductors. Large currents flow from where the lightning strikes and these currents can produce sparks in the vicinity of the point of impact. Even in the absence of lightning, thunderstorms can cause high induced voltages in plants and equipment, which in turn can give rise to sparks and arcs. Installations should be furnished with appropriate lightning protection measures, as specified in installation guidelines for electrical systems and for cathodic corrosion protection systems.

As discussed by Harrold (1992) *high-frequency electromagnetic waves* are emitted by all systems that generate high-frequency (HF) electrical energy (10^4 to 10^{11} Hz), e.g. radio transmitters or industrial high-frequency generators for heating, drying, hardening, welding, cutting, etc. All conductive parts located in the radiation field function as receiving aerials. Ignition sources may be generated if the field is powerful enough and if the receiving aerial is sufficiently large. The primary process is generation of electric currents via electromagnetic induction. The currents induced may make thin wires glow, or may generate sparks during the contact or interruption of conductive parts. If induced currents and inductances of conducting parts can be estimated, Figure 2–31 can give guidance as to whether such sparks can ignite the explosive atmosphere of concern. In the case of particularly strong HF fields, for example in the immediate vicinity of high-power HF generators, non-conductive materials can also become heated due to resonance absorption, and can thus become a source of ignition. The energy picked up by the receiving aerial, and the received HF power at any particular wavelength, depend mainly on the distance between the transmitter and the aerial and the dimensions of the aerial.

2.4.3.6 Ignition by a Jet of Hot Combustion Products

The basic knowledge about ignition of explosive gaseous atmospheres by jets of hot combustion products is applied in practice in the design of so-called "flameproof" enclosures. These are enclosures for electrical components and apparatuses designed to satisfy two essential requirements. First, the enclosure must be sufficiently strong to withstand the full

explosion pressure if a gas explosion occurs inside the enclosure. Secondly, any slits or holes in the enclosure wall must be sufficiently narrow to eliminate the hazard of ignition of a possible explosive atmosphere on the outside by the jet of hot combustion products emerging from the slit. International standards specify requirements for design of flameproof enclosures for industrial use. This kind of electrical apparatus is discussed in Chapter 7.

2.4.3.7 Ignition by Light Radiation

Focused radiation in the optical spectral range (10^{11}–10^{15} Hz) can become a source of ignition through absorption. For example, sunlight can cause ignition if focused, e.g. by bottles acting as convex lenses or by concave reflectors. Focused light beams may occur as short pulses or as continuous radiation. A small solid object in the explosive gas mixture, hit by a hot focal point created by such radiation, can resemble either an electric spark or arc, and depending on the temperatures generated, on the solid object, it may become an effective ignition source.

With laser radiation, e.g. in communication systems, distance-measuring devices, surveying work and visual-range meters, the power density of even an unfocused beam, even at great distances, can be so great that ignition is possible. Here, the heating process occurs mainly when the laser beam is absorbed by a solid body surface or dirty transparent parts.

2.4.4 Controlling Ignition Sources

This section is based on Statoil (2004).

2.4.4.1 Introduction

The concept of ignition source control, as opposed to ignition source prevention, implies that potential ignition sources may exist in an area as long as there is no significant concentration of combustible gas or vapor in the atmosphere. However, as soon as significant concentrations of combustible gas/vapor are detected these potential ignition sources shall be removed from the area, de-energized, or in some other relevant way be brought in a state/situation where they are unable to initiate an explosion.

2.4.4.2 Equipment in Naturally Ventilated, Normally Non- Hazardous Areas

Electrical and non electrical equipment installed in non-hazardous, naturally ventilated areas shall at least comply with the requirements for equipment to be used in zone 2 (see definition in Chapter 7). This is to ensure that the equipment does not act as an ignition source in a situation where explosive gas enters the normally non-hazardous area. Use of equipment that is not specifically protected with regard to becoming an ignition source shall be based on a specific safety assessment considering realistic worst-case scenarios of explosive gas cloud formation. In the offshore industries dispensations for use of non compliant equipment shall be based upon a specific safety assessment as well as measures for avoidance of effective ignition sources, e.g. isolation (e.g. de-energizing) upon gas detection.

Safety related equipment that should remain operative after completion of an extensive evacuation of e.g. an offshore oil/gas installation, shall as a minimum comply with hazardous area zone 2 requirements (see definition in Chapter 7). Such equipment includes:

- emergency lighting
- communication equipment
- navigation aid flooding lights for lifeboat area and helicopter deck
- fire pumps

2.4.4.3 Isolation of Electrical Ignition Sources

Potential electrical ignition sources, equipment and activities shall be identified and a philosophy/strategy shall be implemented for disconnection/isolation in case of an accidental gas leak. Explosion-unprotected equipment is to be categorized in one of three groups, depending on the need for the equipment to be in operation in the case of a gas leak.

Group 1 contains electrical equipment that does not affect production availability or safety integrity. Such equipment should be disconnected/isolated on one single "low" gas alarm anywhere on the installation, or by the signal "confirmed gas in hazardous area." Group-1 equipment includes:

- welding sockets and sockets for hand tools
- air driven tools

- temporary electrical equipment
- electrical heating elements
- hot work activities

Explosion-unprotected electrical equipment including electrical heating elements located in non hazardous mechanically ventilated areas shall be isolated upon a single low gas detection alarm in HVAC intakes. Electrical equipment, except heating elements, located inside rooms may be kept alive until detection of confirmed gas in HVAC intake, provided a safety analysis has documented sufficiently low ignition probability.

Group 2 contains electrical equipment that must be kept powered to maintain production, drilling operations etc. Such equipment should be disconnected/isolated on the signal "confirmed gas in non hazardous areas." Group-2 equipment includes:

- main power generators
- main electrical distribution panels
- diesel engines
- heaters
- boilers
- ventilation systems
- emergency generators, unless defined as a safety critical item

Explosion-unprotected equipment located inside mechanically ventilated rooms shall be subject to instantaneous automatic isolation upon confirmed gas detection in HVAC inlets. Explosion-unprotected equipment located inside rooms with doors/air locks facing classified areas or located in the close vicinity of classified areas shall be subject to instantaneous automatic isolation; either, upon confirmed gas detection in the room due to ingress via doors/air locks or, by coincident loss of ventilation flow and single low gas alarm in adjacent areas.

Group 3 contains safety critical electrical equipment that must be in operation to ensure safe escape and evacuation, and/or to prevent accident escalation. In the case of a major emergency situation such equipment shall be disconnected/isolated with a time delay. Typical Group-3 equipment constitutes parts of larger systems like:

- un-interruptable power supplies (from batteries) (UPS)
- fire and gas detection systems (F&G)
- emergency shut down systems(ESD)
- process shut down systems (PSD)
- public address systems (PA)
- emergency communication systems (EC)

On offshore platforms, equipment required to secure oil or gas wells in an emergency situation shall be defined in an emergency shutdown procedure. A back-up procedure for manual shutdown shall also be established.

On offshore installations safety control units, un-interruptable power supplies and emergency switchgear shall be located in protected rooms inside the living quarters, or in the vicinity of the central control room, based upon safety assessments. If such equipment is not properly located, the minimum requirement is that it shall be suitable for use zone 2 areas (see definition in Chapter 7). This will normally apply to all active field equipment such as detectors and loudspeakers including those located inside non hazardous rooms and living quarters.

2.4.4.4 Isolation of Non-Electrical Ignition Sources

Diesel or gas fuelled equipment such as prime movers and heaters shall be stopped immediately upon confirmed gas detection in ventilation air intake to the enclosure or combustion chamber, and the equipment (non explosion protected) shall be shut down in accordance with the assigned ignition source isolation groups and equipment location.

Group-1 equipment comprises non essential, non electrical equipment. For such equipment the following points should be observed:

- This type of equipment shall normally be located in enclosed areas/ compartments.
- Explosion-unprotected equipment being accepted for location in naturally ventilated areas shall be shut down instantaneously by single low gas alarm anywhere on the installation. Explosion-unprotected equipment located in a non-classified mechanically ventilated area shall be isolated/de-energized upon specified gas alarms or confirmed gas.
- Exhaust pipes and hoses shall be fitted with spark arrestors.

Group 2 is composed of essential non electrical equipment, which shall be located in enclosed areas/compartments that are either safe by pressurization or the probability of gas ingress is found to be sufficiently low to ensure an acceptably low risk of explosion. For such equipment the following points should be observed:

- Explosion-unprotected equipment located inside mechanically ventilated rooms shall be subject to instantaneous automatic isolation by confirmed gas detection in the HVAC inlet of the room or in combustion inlets, or by over-speed of rotating machinery.
- Explosion un-protected equipment located inside rooms with doors/air locks facing classified areas, or located in the close vicinity of classified areas, shall also be isolated/de-energized automatically upon either confirmed gas detection in the room due to ingress via doors/air locks, or upon coincident loss of ventilation flow, or gas alarm in adjacent areas.

Group 3 contains safety-critical, non-electrical, and explosion-unprotected equipment, e.g. emergency generators. Such equipment shall in general be subject to the requirements for fire pumps, but it shall also be de-energized upon confirmed gas detection in air inlets, and/or, after a specified time delay, in the case of a major emergency (APS).

2.4.4.5 Cranes

In order to reduce the ignition probability all crane operations shall be stopped and cranes shut down, once a leakage of flammable gas has been detected.

If the crane is in operation in a situation of a single "low gas" alarm in any location on the installation, the crane operator shall immediately take the crane into a safe position, secure the load and perform a manual shutdown. Upon confirmed gas detection in crane ventilation/combustion air intake, an automatic shutdown of the crane, without time delay, shall be initiated. Adequate means for crane operator warning and consistent operating procedures shall be available.

When the crane is not in operation, potential ignition sources in the crane shall be isolated/de-energized automatically upon a single "low gas" alarm anywhere on the installation. In the case of cranes located in non

hazardous areas external equipment shall generally be certified for use in hazardous area Zone 2 (see definition in Chapter 7), as a minimum. The same applies to internal/local equipment, if operative after main crane shut-down and isolation/de-energizing.

2.4.4.6 Human Machine Interface (HMI) in Central Control Room

Human Machine Interface (HMI) in the central control room shall present extensive system information and means for operator interactions. Therefore the HMI shall include the operator stations (VDU) and a simplified hardwired safety matrix panel allowing manual activation of major safety barrier functions from the central control room. This also includes operator monitoring and control of ignition sources in an acceptable manner.

2.4.4.7 Operation, Inspection, and Maintenance

All equipment that are potential ignition sources shall be operated, inspected and maintained to ensure that any such equipment is in a proper condition at all times. A maintenance program including descriptions and instructions for testing, inspection and repair shall be available. Safety-critical functions and components shall be tested regularly according to accepted standards to verify that the specified safety integrity level and failure rates are fulfilled.

2.4.5 Mitigating Gas/Vapor Cloud Explosions that Occur Despite Preventive Measures

2.4.5.1 Control and Mitigation—Two Different Concepts?

Tam (2002) suggested that the terms *control* and *mitigation* of accidental explosions should have different specific meanings. He defined *mitigation* as an action causing a reduction of the consequences of an explosion event, without there being any reduction in the explosion severity at the explosion source. *Control*, on the other hand, would be when a device or technique has a direct impact on the severity of an explosion at its source. Automatic explosion suppression and partial

inerting would then be defined as controlling techniques, whereas explosion venting would be a mitigatory technique.

Tam's distinction between control and mitigation is intriguing, but the distinction may not always be entirely clear in practice. In the present text the term mitigation will be used to cover both aspects.

2.4.5.2 Explosion Isolation: Minimizing Explosion Propagation *inside* Process Equipment Using Physical Barriers

2.4.5.2.1 Reasons for Using Explosion Isolation

The term "isolation" is not entirely unambiguous. In some cases it implies closure of pipelines and other connections between sections of a process plant to stop the flow of liquids and gases. In the present context the concept "explosion isolation" comprises means for preventing gas explosions from spreading from one process volume to another via pipes and ducts. There are three main reasons for adopting this concept. Firstly, there is always a desire to limit the extent of the explosion as far as possible.

Secondly, a gas flame propagating in a duct between two process units tends to accelerate due to flow-induced turbulence in the gas ahead of the flame. For a sufficiently long duct this may result in a vigorous flame jet entering the process unit at the down-stream end of the duct. The resulting extreme combustion rates can generate very high explosion pressures even if the process unit is generously vented. This effect was demonstrated in a dramatic way for flame-jet-initiated explosions of propane/air in a generously vented 50 m^3 vessel, by Eckhoff et al. (1984).

The third main reason for preventing flame propagation between process volumes is *pressure piling*. This implies that the pressure in the unburned dust cloud in the downstream process unit(s) increases above atmospheric pressure due to compression caused by the expansion of the hot combustion gases in the unit where the explosion starts, and in the connecting duct(s). Experiments have shown that the final explosion pressure in a closed vessel is proportional to the initial pressure of the explosive mixture. Therefore, in a coupled system, higher explosion pressures than would be expected from atmospheric initial pressure can occur transiently due to pressure piling.

This was demonstrated in a laboratory-scale gas explosion experiment by Heinrich (1989) as shown in Figure 2–60. In spite of the marked cooling by the walls in this comparatively small experiment, the transient peak pressure in V_2 significantly exceeded the adiabatic constant volume pressure of about 7.5 bar(g) for atmospheric initial pressure. Very serious situations can arise if flame jet ignition and pressure piling occurs simultaneously. It is important, therefore, to take effective measures to prevent propagation of gas and vapor explosions from one process item to another through connecting pipes and ducts.

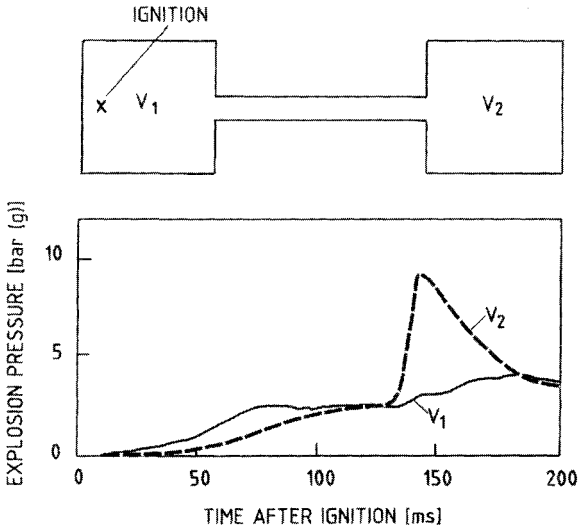


Figure 2–60 Pressure development in two closed vessels of 12 liters each, filled with 10 vol.% methane in air at atmospheric initial pressure and normal temperature, and connected with a 0.5 m long duct, following ignition at the point indicated in the figure. From Heinrich (1989).

2.4.5.2.2 Mechanical Valves

One means to stop explosion propagation is to install various types of mechanical valves in the connecting pipes/ducts. Such valves can either be passive or active.

One passive device for interrupting gas and vapor explosions in ducts is the Ventex valve described by Rickenbach (1983) and illustrated in Figure 2–61. In normal operation the gas flowing in the duct passes

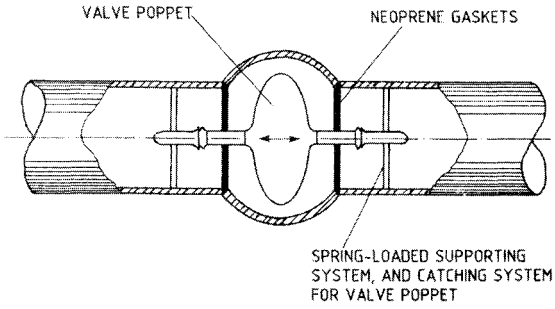


Figure 2–61 Illustration of the Ventex valve described by Rickenbach (1983). From Eckhoff (2003).

around the valve poppet without causing any significant off-set as long as the flow velocity is less than about 20 m/s. However, in case of an explosion in the duct, the preceding blast pushes the valve poppet in the axial direction until it hits the neoprene gasket, where it is held in position by a mechanical catch lock, which can be released from the outside.

Active Ventex valves are also being used. In this case a remote pressure or flame sensor activates a separately powered system that closes the valve in the desired direction prior to arrival of the flame.

Other active mechanical valves include various types of *fast-closing slide valves*. The closure is triggered by the signal from a flame or pressure-rise probe positioned upstream of the valve. The required closing times of such valves depend on the distance between the remote pressure or flame sensor, and the valve, and on the type of explosive gas. Often closing times as short as 50 ms, or even shorter, are required. This most often is obtained by using an electrically triggered explosive charge for releasing the compressed air or nitrogen operating the valve. The valves must be sufficiently strong to resist the high pressures that can occur on the explosion side after valve closure. In the case of pressure piling effects and/or detonation, the transient pressure peaks can be very high.

2.4.5.2.3 Flame Arresters

Flame arresters are a well-proven means of effectively stopping flame/explosion propagation in pipelines, as well as out of or into process enclosures. The following account is based on Halstrick (1995).

Depending on the individual operating conditions, different combustion processes such as deflagrations, detonations and endurance burning are possible. This calls for flame arresting devices that are suitable for the actual situation, i.e. deflagration-proof, detonation-proof or endurance-burning-proof devices. Except for very small devices for special applications, all-purpose flame arresting devices, i.e. devices that stop flame propagation in all three combustion modes, do not presently meet all standard requirements for being installed. The reason for this is that the design/construction of flame arresters for protection against deflagrations, detonations and endurance burning differ considerably.

One of the most important components of a flame arresting device is the *flame trap*, the purpose of which is to stop the flame propagation, whether a deflagration, a detonations or endurance burning. The different design concepts in use can be categorized as either *static* or *dynamic* flame traps, of which the former if further subdivided into *dry* and *liquid* types.

Dry static flame traps are available as close-meshed sieves, plate arresters, sintered metals and crimped ribbon-type traps. It is crucial that the width of gaps in the trap can be manufactured to exact, reproducible tolerances. This is not possible for flame traps of metal-wool, metal-foam or knitted metal mesh. Simple wire gauze ("Davy screens") is also inadequate because it is easily destroyed even when exposed to quite harmless low-over-pressure explosions. The most common and reliable design of dry static flame traps is the crimped-ribbon-type trap illustrated in Figure 2–62. Such traps can be manufactured with high reproducibility to the desired gap width based on the MESG (see Section 2.2.7 and Table 2–2) of the combustible gas or vapor in question.

Depending on the application of the flame trap, whether deflagration-proof, detonation-proof or endurance-burning-proof, two, three or more disc modules are mounted in series in a specially designed enclosure that satisfies the mechanical strength requirements for the actual application and provides the appropriate gap length for the required heat dissipation.

Liquid static flame traps are used in process systems where the combustible fluid is partly or fully in the liquid phase. This liquid is then used as

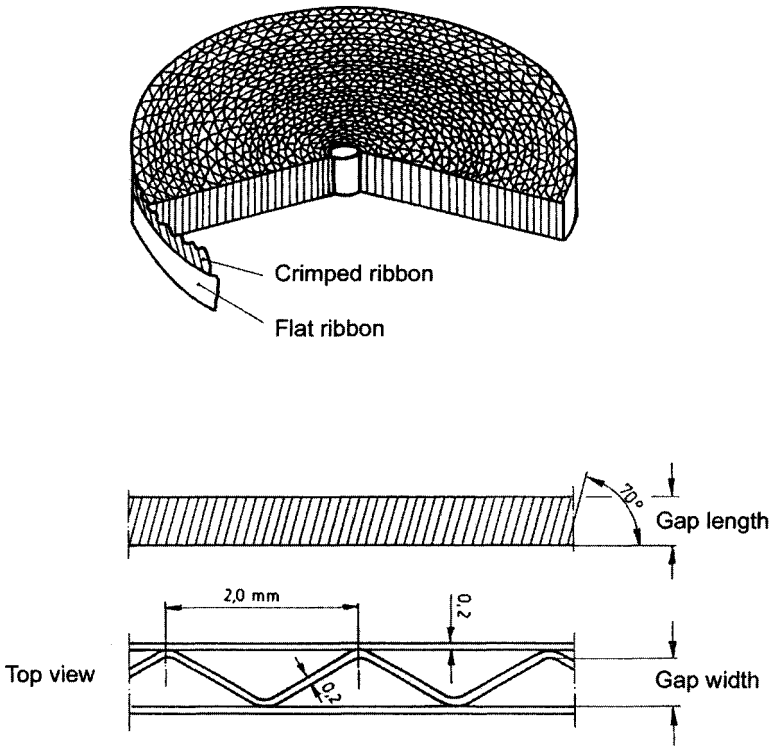


Figure 2–62 Single flame arrester disc module for interrupting propagating gas or vapor flames, based on flame quenching/cooling of hot combustion gases in narrow channels made from corrugated metal strip. Typical standard gap widths are 0.15, 0.2, 0.3, 0.5, 0.7 and 0.9 mm. From Halstrick (1995).

the sealing medium in the flame trap. This trap design is always adopted for flame arresters to be used in liquid-filled pipelines that may drain empty. An example is petrol filling and emptying lines, where the petrol is used as the sealing liquid operating as the flame trap. The absence of oxygen after a deflagration/detonation prevents ignition of the flammable sealing liquid. Static liquid flame traps can only be used for stopping deflagrations and detonations, and not in process systems where endurance burning can be expected. Figure 2–63 illustrates the principle of a liquid static flame trap.

The basic principle of a dynamic flame trap is that the explosive mixtures must have a flow speed which is well above the flame propagation speed. One application of dynamic flame traps is in safety relief valves that open

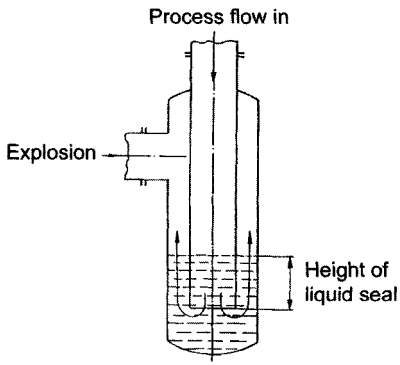


Figure 2–63 Illustration of a liquid static flame trap for stopping deflagrations and detonations in pipelines. From Halstrick (1995)

at quite high pressures, which results in high flow speeds. Examples are incinerator inlets and flare systems in which the flow speeds are maintained well above the flame propagation speed in the respective mixture. *Explosion proof end-of-line flame arresters* must be designed to prevent flame flash-back into the pipeline in the case of an external atmospheric vapor cloud explosion.

Volume explosion flame arresters are designed for specified volumes. This means that these arresters provide safety only for conditions that are almost equivalent to the original test conditions (size and type of explosion volume and size of opening).

Explosion proof in-line flame arresters consist of an explosion-pressure-tight body with the flame arresting element which in turn is composed of the actual flame filter disc (see Figure 2–62) of two or more layers and the enclosing cage. The gap width of the flame filter disc has to be adjusted to the appropriate MESG to prevent flame transmission. It is important that this type of arrester is not exposed to other flame propagation speeds and explosion pressures than the range for which the devices have been tested and approved.

Dry detonation arresters with static flame trap consist of a detonation shock absorber in front of a flame arresting element, as illustrated in Figure 2–64.

The purpose of the shock absorber is to convert a detonation into a deflagration before the flame arrester is reached. It is crucial that the detonation shock absorber points towards the possible point of ignition. If

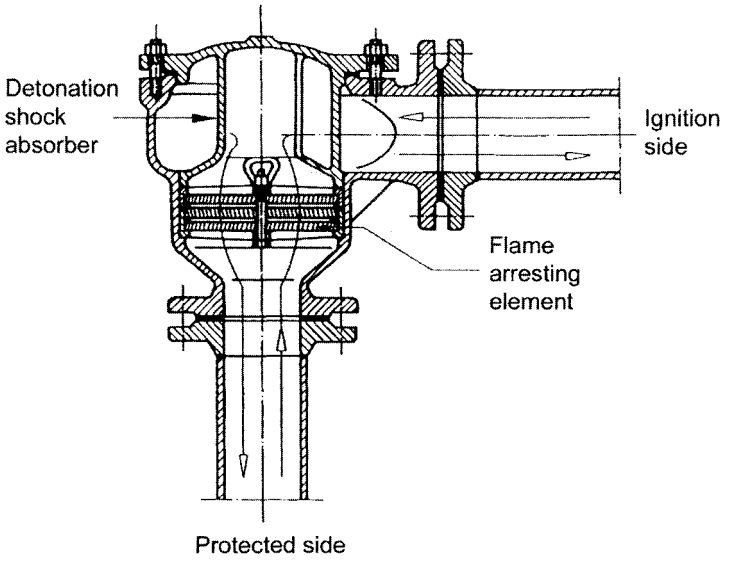


Figure 2–64 Dry detonation arrester with static flame trap. From Halstrick (1995).

pointing in the opposite direction only deflagrations will be arrested. In some situations approach of detonations has to be expected from both sides, and bi-directional detonation arresters are required. It is important to note that in-line flame arresters in general and detonation arresters in particular do not prevent flash back in endurance burning situations, and in order to handle such situations additional measures have to be taken.

Filling and emptying lines of storage tanks for flammable liquids have to be protected by detonation arresters unless, during operation, the lines are permanently filled with liquid. If this is not the case, there is a possibility of explosive vapor/air mixtures being developed in the system, and *liquid-type detonation arresters* may be installed. In this case the liquid itself forms the barrier absorbing the detonation shock. Such devices constitute a complete protective system and shall therefore ensure that the liquid barrier is maintained under all circumstances. For example, if the line connected to the vessel is to be emptied by a suction pump, a static flame trap is installed as a by-pass to prevent that the suction operation empties the liquid barrier.

Special *flash-back-proof detonation foot valves* are also used for preventing flame transmission between vessels and connecting lines, while emptying the line. These devices are also based on the used of liquid barriers.

To prevent flame flash-back into a line in the case of *endurance burning* outside the line end both dry static and dynamic end-of-line flame traps are used. Where explosive atmospheres may be expected in storage tanks for flammable liquids, endurance-burning-proof flame arresters are used in the direct venting system to the atmosphere to prevent flames in the lines from entering the tanks. Vent caps are used if loss of flammable liquid by evaporation is of minor importance, whereas pressure/vacuum relief valves are applied whenever evaporation losses have to be avoided for environmental or economical reasons, or both.

2.4.5.2.4 Flame Interruption by Automatic Injection of Suppressant

A third active explosion isolation method is interruption by fast automatic injection of extinguishing chemicals ahead of the flame. This is a special application of the automatic explosion suppression technique described in Section 2.4.5.8. Important design parameters are type of combustible gas/vapor, size of the enclosure where the explosion starts, ignition location, duct diameter, distance along the duct from the primary explosion enclosure to the point of suppressant injection, method used for detecting onset of primary explosion, and type, quantity and rate of release of suppressant.

2.4.5.3 Minimizing Gas Cloud Size and Controlling Gas Cloud Location *outside* Process Equipment Using Physical Barriers

The following account is in part based on the paper by Tam (2002):

2.4.5.3.1 The Hazardous Situation

In the chemical and process industries, controlling/mitigating potential gas/vapor explosions outside process equipment, due to leaks from such equipment, has been and still is a major concern. In the offshore oil and gas industries, the main emphasis has been on controlling explosions in large, fully or partially enclosed areas containing process equipment ("modules"). The hazardous explosive gas clouds are then generated by leaks from process equipment and pipe works operated at pressures considerably higher than atmospheric. In the past "Halons" (halogenated hydrocarbons) were used for controlling the development of gas and

vapor explosions in confined enclosures of smaller volumes, e.g. turbine housings. However, due to their negative effects on the environment the "Halons" are no longer used. In smaller confined volumes such as turbine housings it has to a large extent been possible to replace "Halons" by water mist. In more open larger volumes (modules), the explosion controlling methods used include optimization of equipment and pipe work layout to reduce flow-generated turbulence, and hence reduce the explosion rate. Other controlling methods include the use of blast walls and total water deluge.

2.4.5.3.2 Definition of a Physical Barrier

As the name suggests, a *physical barrier* is a physical arrangement that mitigates the severity of a gas/vapor explosion. Such an arrangement can for example be a wall. Total area water deluge would not be classified as a barrier technique, but as a total flooding technique. Tam (2002) distinguished between two main categories of barrier methods, viz. cloud size and location limitation and explosion development control.

2.4.5.3.3 Cloud Size and Location Limitation Barriers

The main purpose of this class of barrier is to minimize the volume of flammable gas that a gas explosion is allowed to consume, and to control the location of this volume. Generally the over-pressure load on structural surfaces is expected to increase with increasing size of an explosive gas cloud. It is therefore important to limit the size of any potential accidental explosive cloud as far as possible. The most commonly used barrier type is the *blast wall*. In some cases a main function of the blast wall is to limit the size of the explosive cloud, in other cases its purpose is mainly to reduce the impact of a potential explosion, by shielding the structures outside the wall from excessive over-pressure loading. As pointed out by Tam (2002) the concept of using a wall to restrict the size of the explosive gas clouds also includes *weaker walls* than blast walls, and even deliberately weakened walls. The concept of using weak walls is based on demonstration calculations using a Computational Fluid Dynamic (CFD) gas explosion code. The process area was divided into two halves using different arrangements of weak walls. It was shown that maximum over-pressure developed on the structural surfaces could be reduced by

fifty to seventy-five percent, depending upon the arrangement of the weak barrier. *Soft barriers* are very weak and light physical barrier walls. This type of barrier has the ability to separate an area into a number of cells, and thereby control the size of the gas cloud within a hazardous area. The soft barrier control method has been used on oil and gas platforms.

Soft triggered barriers are either *non-suppressive* or *suppressive*. The most commonly used *non-suppressive soft barrier* is a water curtain that has no specific suppressive effect. According to Tam (2002) such curtains have been used successfully in the process industries onshore to contain hazardous gas clouds within a limited area, or to divert such clouds to safer areas. A number of extensive experimental research programs comprising various scales up to full size, have demonstrated that water spray barriers can be effective in containing hazardous gas clouds. However, some types of water curtains are also categorized as *suppressive barriers.* Experiments have shown that this type of curtain, in addition to limiting the size of a dispersing explosive gas cloud, also caused the flame propagation to decelerate as the flame passed through the curtain. The effect was a significant reduction of the maximum explosion over-pressure. *Automatic explosion suppression* (see Section 2.4.5.8) can also be used for establishing soft triggered suppressive barriers.

2.4.5.4 Use of Physical Barriers for Controlling Explosion Violence

Tam (2002) emphasized that two consecutive effects contribute to the increased hazard associated with gas explosions in congested areas, viz.

- pre-compression of explosive gas cloud in the congested area by an explosion developing in adjacent areas (pressure piling), and
- flame propagation in the congested, pre-compressed volumes

2.4.5.4.1 Controlling Pre-Compression (Pressure Piling)

In a process area, one may find regions where both the degree of confinement and the local aspect ratio (length-to-width/diameter) are high. If a gas explosion occurs in such regions, following precompression of the unburned gas there, one can expect very high explosion over-pressures.

As shown in Section 2.4.5.2, pre-compression has a multiplicative and not an additive effect on the maximum explosion pressure. Barriers that isolate the confined volumes from the rest of the area could be an effective means of controlling the maximum explosion pressure that can occur, by preventing the explosive gas to enter these confined volumes. Hard barriers that can withstand the explosion force will effectively eliminate the problem, but soft barriers can also have a significant controlling effect.

2.4.5.4.2 Controlling Turbulence Generation

It is well known that severe gas explosions can develop in regions of high congestion, due to the enhanced combustion rate caused by the turbulence generated in the still unburned explosive cloud ahead of the flame when it is pushed past equipment and pipe work. Experiments of the type illustrated in Figure 2–10, have demonstrated the dramatic increase of the explosion violence that can result with certain obstacle configurations. Tam (2002) presented results from small and medium scale experiments demonstrating the effect of soft barriers in such geometries. The experimental arrangement is illustrated schematically in Figure 2–65.

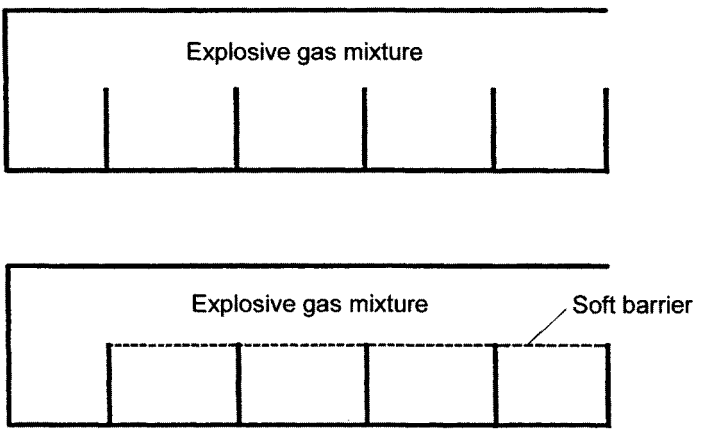


Figure 2–65 Schematic illustration (side elevation) of small and medium scale experiments conducted in Norway to investigate the effect of soft barriers on the violence of methane/air explosions in confined spaces containing turbulence-generating obstacles. From Tam (2002).

The tests comprised two arrangements, one without a soft barrier covering the obstacles (upper part of figure) and one with a soft barrier (lower part of figure). With the barrier, the explosive gas occupied only the upper part of the volume of the test structure prior to ignition. In the series of small-scale experiments the maximum explosion over-pressures were > 250 mbar without the barrier, and only about 5 mbar with the barrier. In the intermediate-scale experiments the effect of the barrier was also significant, viz. 600 mbar without barrier, and 150 mbar with barrier.

Experiments, as pointed out by Tam (2002), have shown that for a given volume of obstructions, the combustion enhancing effect is greater with a large number of small obstacles than with a small number of large obstacles. One potential technique for controlling the explosion violence could then be to enclose a large number of small obstacles (e.g. pipe work) in a larger hard or soft shield into which the explosion would not propagate. Tam discussed a number of potential solutions. In some process areas pre-compression and turbulence generation can mutually enhance each others effects. It is then important to limit the volumes of such areas as far as possible.

2.4.5.4.3 Suitable Materials for Soft Barriers

Tam (2002) pointed out that a soft barrier does not have to sustain the total gas explosion. It only needs to survive long enough to prevent substantial amounts of flammable gas from entering the fuel-free zone during the passage of the flame. For this purpose, low-pressure and low-tension membranes such as meteorological balloons could be used. Such balloons, in effect, divide the fuel-free zone into many fuel free cells. Furthermore, the balloons can be filled with oxygen-free gases or mixtures of air and inert gas. Low-tension and low-pressure balloons can "mold" themselves to the shapes of hard surfaces and overcome the problem of flammable gas being trapped in the volumes between conventional spherical balloons (high pressure, high tension). Tam (2002) refers to large-scale experiments that have demonstrated that low-pressure low-tension balloons could last long enough for the explosion flame to pass the area of the balloon before the balloon collapses. However, so far these methods are not widely used in practice.

2.4.5.4.4 Additional Hazards to be Considered when Using Soft Physical Barriers Other than Water

Focusing only on the reduced maximum explosion pressure, when using soft physical barriers, is not sufficient. Some other important safety issues must also be resolved, viz.;

- possible emission of missiles or fragments resulting from collapsing barriers
- creation of small volumes/pockets of explosive gas that cannot be reached by the ventilation systems, because of the barriers
- unacceptable reduction of operational efficiency due to the presence of the barriers
- more complicated maintenance procedures due to the presence of the barriers

For these and other reasons the use of soft barriers (other than water) has only gained limited terrain so far.

2.4.5.5 Design of Plants and Buildings to Prevent Damage by Gas Explosions

It is important to distinguish between ideal requirements and realistic possibilities. In all circumstances it is strongly recommended that the explosion problem be taken into account as early as possible in the planning process, whether a completely new plant is to be constructed, or an existing plant rebuilt.

Ideally any plant in which gas explosions may occur should be located at a safe distance from other activities, buildings etc. Furthermore, the various parts of the plant should be separate to enable effective isolation of the explosion to within the section of the plant where it starts. Buildings should be one-story whenever otherwise suitable. If multi-story buildings have to be used, the parts of the process representing the greatest explosion hazard should be located as high up as possible, preferably on the roof.

In the past, floors and roofs of factory buildings were often supported by recesses in comparatively weak walls with no reinforcement, as illustrated in Figure 2–66(a). In the case of an explosion, the walls were easily displaced outwards even at very modest over-pressures, and the floors

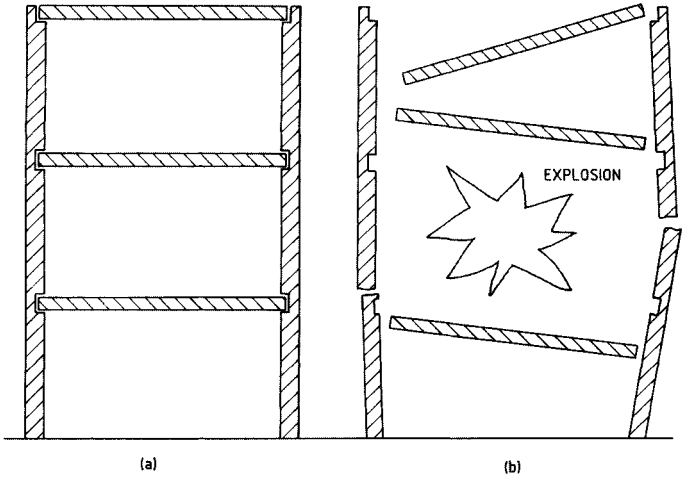


Figure 2–66 Inappropriate construction of multi-storey building with regard to preventing hazardous collapse of the floors in case of an explosion inside the building. From Eckhoff (2003).

and roof fell down into the building as illustrated in Figure 2–66(b). Clearly, under such circumstances the consequences of even minor explosions in the building could be catastrophic. A much better solution is that the roof and intermediate floors be supported by a strong frame structure. The walls are then low-mass panel sections that function as vent covers, should an explosion occur. If required, the panels must be anchored to the frame structure.

In some cases, hazardous plant items can be placed in special, isolated, well-vented niches as illustrated in Figure 2–67. Depending on the location, the floor and roof of the niche may also have to be explosion proof. However, if the building is constructed of reinforced concrete, it can be made sufficiently strong to enable the windows to serve as vents It is then important to ensure that flying pieces of glass do not present a hazard to humans. To avoid this hazard it may be necessary to replace glass panes by anchored, transparent plastic panes.

Some final points to be taken into account when planning layout and construction of process buildings with the aim to reduce the explosion and fire hazard include:

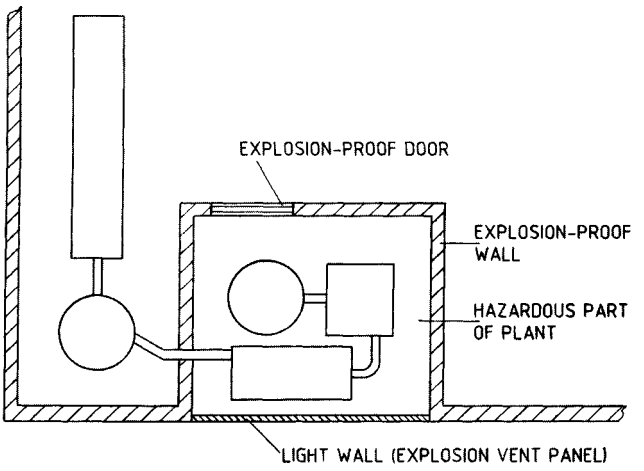


Figure 2–67 Illustration (top view) of arrangement where the hazardous part of the plant is located in an isolated well vented niche inside the building. From Eckhoff (2003).

- safe escape routes in case of explosion and fire
- fire resistant construction materials
- fire resistant doors
- electrical installations according to latest regulations/ recommendations

2.4.5.6 Explosion Venting

2.4.5.6.1 What is Explosion Venting?

The basic principle is illustrated in Figure 2–68.

The maximum explosion pressure in the vented explosion, P_{red}, is a result of two competing processes:

- burning of the explosive cloud, developing heat and increasing the pressure
- flow of unburned, burning and burned cloud through the vent, relieving the pressure

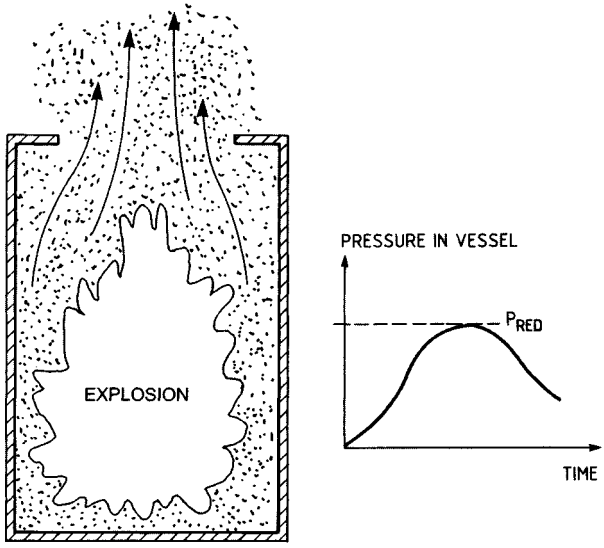


Figure 2–68 Illustration of the principle of explosion venting.

The two processes can be coupled via flow-induced turbulence that can increase the burning rate. The maximum permissible pressure, P_{red}, depends on the construction of the enclosure, and on whether plastic or only elastic deformation is acceptable in the case of an explosion.

Explosion venting of simple process enclosures is less common in the case of gases and vapors than in the case of explosive dust clouds (see Section 5.6.4.4). Gas explosion venting is more often applied to rooms of onshore facilities and processing modules on offshore gas and oil installations. This reflects the fact that accidental gas explosions most often occur in explosive clouds generated outside process equipment by accidental leaks, whereas dust explosions are practically always initiated in dust clouds generated inside process equipment as part of the process.

2.4.5.6.2 Vent Covers

Vent covers for process enclosures are discussed in Section 5.6.4.4. For rooms, e.g. offshore process modules, as pointed out by Bjerketvedt et al. (1997), typical vent covers include:

- louvered walls
- solid wall cladding
- relief walls (also called wind walls or weather cladding)
- glass windows (not recommended)

2.4.5.6.3 Potential Hazards Caused by Venting

See Section 5.6.4.4 (dusts)

2.4.5.6.4 Use of Vent Ducts

See Section 5.6.4.4 (dusts)

2.4.5.6.5 Reaction Forces Caused by Venting

See Section 5.6.4.4 (dusts)

2.4.5.7 Use of Mathematical Simulation Models in Design for Mitigating Accidental Gas Explosions in Complex Geometries

In some industries, e.g. petrochemical plants and oil refineries, the process areas are highly congested, as illustrated in Figure 2–69. In such cases the inherent positive feedback loop of the explosion development can generate very high burning rates and pressure piling effects, and hence very high, destructive explosion pressures. Today one can use comprehensive numerical computer simulation models to predict the development and strength of gas explosions even in such extreme situations. This makes it possible to take appropriate precautions, e.g. in terms of adequate safety distances to neighbouring activities.

The following summary is based on the text of Czujko (2001):

Figure 2–69 Highly congested process plant in which very high explosion pressures can be generated in case of a gas explosion.

2.4.5.7.1 CFD-Models in General

In the general Computational Fluid Dymanics (CFD) approach the aim is to resolve the physics of flow development numerically by dividing the actual space into small boxes (control volumes) and implement models for the various physical phenomena like fluid flow and turbulence. In each cell, all variables are assumed constant in one time step, and based on the flow balance and fluxes, as well as the physics taking place inside the cell in the next time step, the variables may change. Furthermore, in CFD explosion simulation not only physical flow and turbulence must be modelled. It is also essential to model combustion/flame development. Thus the conservation equations for

- mass
- impulse

- turbulence fuel transport and mixture fraction, and
- enthalpy

are solved for each time step and control volume.

Some CFD models are designed specially for simulating gas explosions, and they have been extensively validated by experiments on various scales. The most well-known models of this type are FLACS, EXSIM and Auto Reagas. During the last years, the development of computers has substantially increased the possibility of applying CFD for simulating explosion development in complex large-scale systems. There are nevertheless still strong limitations on the spatial and temporal resolution that can be obtained in realistic industrial scenarios. For this reason, one incorporates so called sub-grid models to resolve the phenomena that occur on a smaller scale than typical numerical cell sizes (0.5–1 m). Such models take care of the influence on the combustion of e.g.:

- turbulence
- flame folding
- water deluge

The need for sub-grid models is one reason why development of comprehensive CFD codes, has to be based on extensive validation by experiments.

2.4.5.7.2 The FLACS Model and Its Applicability

FLACS has been developed at GexCon, Bergen, Norway (previously Christian Michelsen Research (CMR), and initially Christian Michelsen Institute, (CMI)) over a long period beginning in 1980. The first version of FLACS was created by B.H. Hjertager. The work has been supported by major oil and gas companies.

FLACS uses a 3-D Cartesian implicit Navier Stokes solver based on the Simple method by Patankar with a compressibility extension. Implicit/explicit second-order terms are implemented for increased accuracy. Flow equations are closed by means of a k-ε model for turbulence. Models for wall function and turbulence generation from sub-grid geometry are required and implemented. A distributed porosity concept is used for mapping the complicated geometries onto the simulation mesh. In devel-

oping such a mapping model there are several pitfalls, and the "perfect" solution seldom exists. Further, two equations for handling fuel transport and consumption are solved. To model the combustion process properly a range of sub-grid models are needed. A flame model describes the extent of the reaction zone, whereas burning velocity models decides the velocity of the flame. This velocity, in turn, is a function of fuel composition and concentration, oxygen concentration, temperature, pressure, turbulence level, various flame instabilities, and presence of non flammable gas dilution or water mitigation. Corrections for increased flame area due to flame folding around sub-grid obstacles are included.

Special FLACS models simulate gas leaks, gas flow across boundaries, effect of water spray on explosion development, effect of pressure relief panels, and heat, ventilation, air conditioning systems (HVAC). A multi-block solver for propagation of blast waves in the far field was developed in 1994. In-house versions of FLACS also have models for heat transfer through walls and thermal radiation.

Further models for implementation in FLACS are under development or being planned, including:

- models for predicting deflagration-to-detonation transition (DDT) and for simulating detonations
- models for predicting auto-ignition
- models for simulating various types of flame instabilities (important for low confinement/low congestion scenarios)

In addition the following models have been developed based on FLACS for gases:

- models for oil mist leaks, evaporation, deposition, dispersion, and explosion (Chapter 3)
- models for dust cloud generation and dust explosions (Chapter 5)

2.4.5.7.3 Adequate Representation of Geometry is Essential

Because the geometry of the system in which the explosion takes place can have a decisive influence on the explosion development, and hence on the maximum explosion pressures produced, it is essential that CFD

models for gas explosion simulation represent this geometry properly. Geometries can be either defined manually or imported from CAD systems. In FLACS simulations a porosity-distributed flow resistance concept is used for automatic geometry representation and contains no subjective estimates of blockage, which is often encountered in other CFD simulators.

2.4.5.7.4 Ventilation and Dispersion Simulations

Transient leaks of combustible gas, sonic, as well as low momentum, in combination with natural and forced ventilation, can be simulated by means of the FLACS code. Liquid leaks that are converted into equivalent gas leaks through evaporation, can also be simulated. The numeric handling of the geometry of the explosion environment includes minute details, which are important for ventilation calculations, and can be very important for gas dispersion calculations.

2.4.5.7.5 Gas Explosion Simulations

The range of scales that may be simulated include gas volumes from less than one liter to thousands of m^3. The gas cloud can either be pre-defined, or be calculated from a dispersion simulation preceding the simulation of the explosion. For the present (2005) the gas clouds that can be handled by FLACS comprise any combination of methane, ethane, propane, butane, propylene, ethylene, acetylene, hydrogen, CO and H_2S, either mixed with air or with oxygen-enriched air. Initial pressure, temperature, and turbulence can be varied. Systematic simulation studies can be performed to quantify the effects on the maximum explosion pressure of re-location and re-arrangement of process equipment, altering the degree of confinement of the area studied, use of various types of relief panels, use of inert gas dilution (N_2 or CO_2) and water deluge etc.

2.4.5.7.6 Far Field Blast Calculations

Using a multi-block concept, blast waves from the explosion that propagates into the far field and interacts with objects there, can also be simulated.

2.4.5.8 Automatic Explosion Suppression

2.4.5.8.1 General Concept

The first patent for a fast fire suppression system was allotted to a German company as early as 1912. The Second World War accelerated the development. The British Royal Air Force found that eighty percent of the total losses of aircraft in combat were due to fire. Based on this evidence a military requirement was issued that specified a light-weight, high-efficiency fire extinguishing system for protecting air craft engines and their fuel systems. A similar situation arose in Germany. As a result new, fast acting fire extinguishers were developed based on three main principles:

- extinguishing agent permanently pressurized
- large diameter discharge orifice
- very fast opening of valve for immediate release of extinguishing agent by means of an explosive charge

These principles, combined with a fast-response optical or pressure rise flame detection system, form the basis of today's automatic explosion suppression systems, as illustrated in Figure 2–70.

The suppressor contains a suitable extinguishing agent (suppressant) and a driving gas, normally nitrogen at 60–120 bar. If ignition of the explosive atmosphere should occur, the onset of pressure rise in the enclosure due to the growing flame is detected automatically by the pressure sensor, which produces an electric signal that triggers an explosive charge opening the suppressor valve almost instantaneously. A special nozzle design ensures that the suppressant is distributed as evenly throughout the enclosure volume as possible. In principle the pressure sensor can be made sensitive enough to detect even a very small initial flame. However, if the minimum pressure rise for triggering the opening of the suppressor valve is chosen to be so small that similar pressure variations may occur due to normal plant operation, false activation of the suppression system becomes likely. The use of two pressure detectors oriented perpendicular to each other can make it easier to discriminate between pressure rise due to explosions and other disturbances. Also, natural pressure variations can be filtered out electronically if the rates of rise of such normal variations are significantly lower than typical initial rates of rise of the explosion pressure.

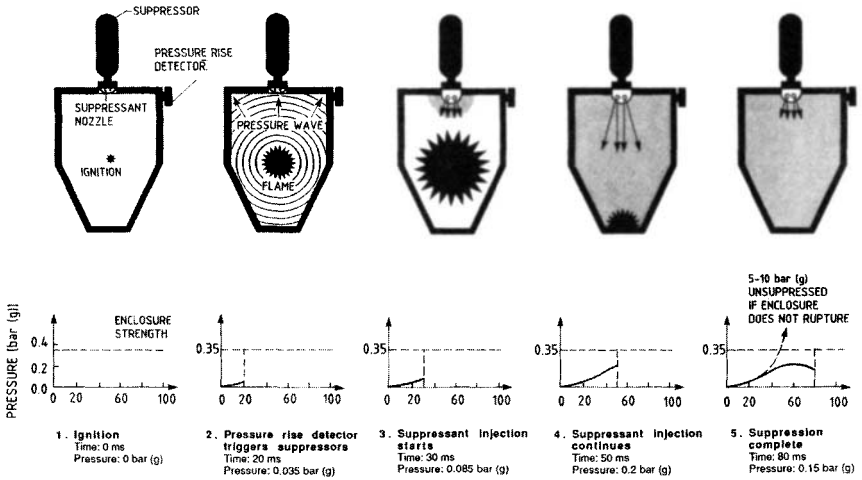


Figure 2–70 Illustration of system for automatic suppression of accidental gas, mist/spray and dust explosions. From Eckhoff (2003).

In the case of gas and vapor explosions ultraviolet or infrared optical flame sensors can be used instead of pressure sensors for detecting the initial flame. A problem with the ultraviolet detectors is that they do not discriminate between hydrocarbon flames and some other light sources, for example welding arcs. Therefore, false alarms may become a problem. As a result, ultraviolet detection systems are often switched off during maintenance and repair work requiring welding, in spite of the fact that the risk of gas leaks may be at a maximum just during such periods of irregular operation. In some cases, this problem has been solved by using a combined system of ultraviolet radiation and gas detectors, requiring that both types of detectors trigger before the suppressant is released. Then radiation from welding will be detected, but without this leading to suppressant injection unless a significant concentration of hydrocarbon gas is detected simultaneously by the independent gas detectors.

As mentioned in Eckhoff (1983) the infrared pyrometer technique offers an alternative solution to this problem. The basic principle is to measure the relative intensities of the radiation emitted by the actual source at two or more discrete wave lengths. From this information, using an integrated computer, the Planck radiation distribution curve that has the same intensity ratios for the same wave lengths can be determined. Any such set of

ratios can be associated with a specific light source temperature, and the detector system is thus able to indicate whether the actual temperature indicated is typical of say a hydrocarbon flame, or whether it is significantly higher, e.g. of the order of welding arc temperatures.

2.4.5.8.2 Basic Approach to Design of Explosion Suppression Systems

Moore (1981) presented a basic approach to design of industrial suppression systems. He introduced the concept of critical mass M_t of suppressant that is just sufficient for suppressing the flame when being evenly distributed throughout the flame volume. He assumed that there exists a critical minimum mass concentration of any given suppressant for suppressing a flame of a given explosive mixture, and that a suppressant cloud of this concentration or higher must occupy at least the flame volume for successful suppression. It then follows that the critical mass M_t increases with time because the flame volume increases with time. A similar line of thought was applied to the mass of suppressant actually delivered at any time after onset of flame development. Successful suppression would result if

$$M_{t,delivered} > M_{t,required}$$

One distinguishes between three different suppression strategies:

- *Advance inerting*: Detect initial explosion, identify its location, activate appropriate suppressors, and establish suppressant barriers to prevent explosion spread to adjacent areas.
- *Local suppression*: Detect initial explosion, identify its location, and activate appropriate suppressors for ensuring no flame propagation beyond explosion kernel.
- *Total Suppression*: Detect explosion and deluge entire system with suppressant to ensure that explosion is totally suppressed.

The design of any particular industrial suppression system depends on the suppression strategy chosen, the type of suppressant, the chemical and explosivity properties of the explosive mixture, the nature of the process/enclosure to be protected, the volume and shape of the enclosure, and on other actions taken to prevent or mitigate accidental explosions in the plant.

In large process volumes it can be difficult to obtain adequate distribution of the injected suppressant if the gas leak to the atmosphere is in the form of a strong gas jet. Depending on the velocity and size of the leak jet, the injected suppressant may not be able to penetrate the jet region.

2.4.5.8.3 Sophisticated Optical Detection Systems may Be Required in Large Process Volumes

If the volume of the enclosure or area to be covered by the suppression system is very large, e.g. of the size of typical offshore oil and gas platform modules, very large flames can develop before a detectable rise in pressure has been generated. In such cases, as described in Eckhoff (1983), pressure rise detection of the onset of explosion is inadequate, and optical flame detectors, either ultraviolet or infrared, have to be used. Ultraviolet detectors are able to detect a fireball of 1 m diameter at a distance of 10–20 m within about 15 ms, during which an initial flame will only have propagated a distance of a few cm. In principle, therefore, such detectors should be able to detect any initial fireball in modules typical of offshore platforms at a sufficiently early stage to ensure effective local suppression. However, an important condition is that the light path between the initial fireball and the ultraviolet detector is not obstructed by walls or process equipment etc. Furthermore, the suppressant containers have to be located close to the origin of the explosion, and be able to throw the suppressant in the appropriate direction. It is clear that for such applications a quite advanced detection system, supported by computers, will be required. First the precise location of the local flame must be identified, and then the adequate quantity of suppressant must be released locally in the area where the flame has been detected.

In order to cover large process volumes adequately, a number of such detectors may have to be located in an optimal matrix covering the entire volume, with the aim to localize the initial flame with the required accuracy. This requirement is quite severe if it is necessary to identify the precise location of flames as small as 1 m^3 in volumes of several thousand m^3 or more.

Chapter 3

Explosions in Clouds of Liquid Droplets in Air (Spray/Mist)

3.1 Introduction

Sprays and mists of combustible liquids, e.g. hydrocarbons, in air at atmospheric pressure and normal temperature, with a droplet size of < 100 μm and a droplet mass concentration in the range 100-500 g/m^3 are explosive. This is so regardless of whether the liquid is of a low or a high boiling point. In the case of a low boiling point liquid, the droplets will evaporate readily and the cloud will very soon become a mixture of combustible vapor and air. If the boiling point is high, i.e. the vapor pressure at normal ambient conditions is low, the droplets will, with regard to the combustion process, behave similarly to solid particles of an organic material, which are also known to be able to cause explosions when dispersed as clouds in air (see Chapter 5).

The phenomenon of spray/mist explosions is complex. Similarly to a gas or a dust explosion it is the result of two main consecutive processes, viz. generation and combustion of an explosive cloud. Ignition may be regarded as the coupling between the two main processes. Because of the partly unstable two-phase nature of a spray/mist, the cloud generation process is in itself very complex.

The terms spray and mist are used somewhat interchangeably in this text, although one can argue that there is a clear difference between the two. *Spray* then means a cloud of liquid droplets extending from one or more nozzles (e.g. accidental leaks from high-pressure process equipment), whereas *mist* denotes a cloud generated by condensation from a super-saturated fuel vapor. This implies that droplets in mists are generally smaller than typical spray droplets. With regard to accidental explosions, both spray and mist may be relevant depending on the actual accidental situation.

Spray/mist combustion research may be motivated by two different needs. The first is to prevent and control accidental explosions, whereas the second is to improve diesel engines, gas turbines, furnaces, rockets etc. in which liquid fuel sprays are burned deliberately. Published research on accidental spray/mist explosions is scarce, and it is therefore useful to consider literature related to deliberate generation, ignition and combustion of sprays. The basic phenomena are the same, whether the research is conducted to prevent explosions or to improve combustion engine performance, and work related to the latter is therefore also to a large extent relevant for resolving the explosion problem. Zehr (1965) pointed out that a spray/mist and a cloud of solid particles (dust) have common features in that both consist of a finely divided dense fuel phase suspended in an oxidizer gas. However, explosive mist clouds are less stable than explosive dust clouds because collisions between droplets give rise to coalescence and transformation to fewer and larger droplets. When the droplet size gets sufficiently large, the droplet sedimentation velocity in the gas becomes significant and the droplets settle out of the cloud.

In his instructive book on combustion of liquid fuel sprays, Williams (1990) has covered a wide range of topics relevant to spray and mist explosions. This includes various properties of liquid fuels, characteristics of sprays in terms of drop size and drop velocity distributions, processes for atomization of liquid fuels and combustion of single droplets and droplet clouds. Both experimental work and theoretical modeling is discussed. Förster (2000) also gives useful information on generation, ignition and combustion of sprays/mists.

3.2 Generation of Clouds Liquid Droplets in Air (Spray/Mist)

3.2.1 General

The production of mists is discussed by Green and Lane (1964) in their classical reference book on particulate clouds. As mentioned above, mists are produced by condensation, following sudden expansion or cooling of high-concentration vapor clouds. A classical theory describing this kind of condensation processes is discussed by Green and Lane. Sometimes small solid particles act as condensation nuclei. Mists can also be generated by slow condensation on solid nuclei.

3.2.2 Mechanisms of Spray Generation

Valuable insight in the mechanisms of accidental spray generation can be obtained by studying the methods used for producing sprays for research purposes in the laboratory. Williams (1990) reviewed various methods for spray formation by "atomization" of liquids. When a liquid is atomized, energy is expended mainly in three ways, viz. in forming new surface, in overcoming viscous forces in changing the shape of the liquid, and in losses due to inefficient application of the energy to the liquid. Atomizing devices commonly used in generating liquid sprays may be classified in three main types:

- *Air-blast or aero-dynamical atomizers,* in which compressed air or other gas at high velocity is used to break up liquid emerging from a nozzle, and to produce a fine degree of atomization. This kind of break-up is adopted in conventional paint spray guns, in Venturi atomizers and in many atomizers that generate aerosols for insecticidal, disinfectant and therapeutic purposes. It is characteristic of atomizers of this type that they give a very wide range of droplet sizes, though in some cases the range is narrowed by trapping the larger droplets within the atomizer.
- *Generators depending on centrifugal action.* In this case the liquid is fed on to the centre of a rotating disc, cone or top and centrifuged off the edge. The spray is characterized by uniformity of the main droplet size, in marked contrast to the heterogeneity of sprays produced by other methods.
- *Hydraulic or hydro-dynamical spray generators*, in which liquid is forced through a nozzle and breaks up into droplets. Here the disintegration depends more upon the physical properties of the liquid and the conditions of ejection from the nozzle than upon interaction between the liquid and the surrounding gas. Probably the most successful hydraulic atomizer, and indeed the only one which has application for fine atomization, is the swirl chamber atomizer used in agricultural spraying equipment, oil-fired furnaces, internal combustion engines and gas turbines. The swirl is produced by leading the liquid tangentially into the chamber and forcing it to spray out through a central orifice of small diameter.
- *Special atomizers*. These include the electrostatic atomizer which breaks up liquid by the action of electrostatic forces and the acoustic atomizer which utilizes high intensity sonic or ultrasonic vibrations.

Lefebvre (1989) summarized the challenges in spray generation as follows:

> The subject of atomization and sprays is one that has attracted the attention of many research workers and has been the focus of numerous theoretical and experimental studies. However, our knowledge of the atomization process is far from complete. The physics is not well understood, the available data and correlations for drop size distributions are sometimes of questionable validity, and there is little agreement between the various investigators as to the exact relationships between liquid properties, nozzle dimensions, and mean drop size. These comments are especially true for pressure-swirl atomizers. Many factors contribute to this unsatisfactory situation. They include the great complexity of the atomization process, differences in the design, size, and operating conditions of the nozzles tested, and the difficulties involved in accurate drop size measurement. These difficulties include the very large number of drops in a spray, the high and varying velocity of the drops, the wide range of drop sizes encountered in most practical sprays, and the changes in drop size with time due to evaporation and coalescence.

3.2.3 Coalescence of Drops in Sprays/Mists

A comprehensive review of the state of the art more than forty years ago was given by Green and Lane (1964). Much of this information is still valid. The same applies to the comprehensive review by Zebel (1966), who considered three main categories of coalescence mechanisms. The first was thermal coalescence, driven by Brownian movement of the droplets, the second coalescence influenced by inter-droplet forces (van der Waals and electrical/magnetic forces). The third category was coalescence under the influence of external force fields (electrical and magnetic, gravitational and centrifugal, sonic, and laminar and turbulent flow fields).

Förster (1990) argued that accidental generation of large, explosive clouds of sprays/mists of organic liquids of high boiling points is not very likely. The reason is that the mean droplet-droplet distance in the explosive range is of the order of only 10 droplet diameters, which in a turbulent cloud makes fast coalescence of the small droplets to larger ones highly probable. The larger droplets will then "rain out" and the fuel concentration in the cloud will fall below the explosive range. Förster confirmed experimentally that with a high boiling point liquid (octanol) it was indeed very difficult to generate an explosive spray in a 1 m^3 explosion chamber unless a highly sophisticated spray nozzle system was adopted.

3.3 Combustion of Clouds of Liquid Droplets in Air (Spray/Mist)

3.3.1 General

Zehr (1965), emphasized that a spray/mist and a dust cloud have common features in the sense that both consist of a finely divided dense phase suspended in a gas (see Eckhoff, 2003). However, spray/mist clouds of high mass densities are less stable than high-density dust clouds because collisions between small drops inevitably give rise to coalescence and transformation of the cloud into one of fewer and larger drops. When the drop size gets sufficiently large, the drop sedimentation velocity in the gas becomes significant and the drops settle out of the cloud.

The ignition and combustion of droplets of organic liquids, and of particles of solid organic materials, is similar. The thermal energy from an ignition source, or from nearby burning droplets/particles, first causes the liquid of the droplet/particle to evaporate partly of fully (via pyrolysis for liquids of high boiling points, and solids). Further heating causes the vapor to ignite and burn, either as individual diffusion flames around each droplet/particle, or as a more or less homogeneous premixed vapor/air flame. Faeth (1987) discussed the theoretical description of mixing, transport, and combustion in sprays and mists. In particular, the description of drop/turbulence interactions was considered.

In his instructive book, Williams (1990) covered a wide range of topics relevant to spray explosions. This includes various properties of liquid fuels, characteristics of sprays in terms of drop size and drop velocity distributions, processes for atomization of liquid fuels, and combustion of single droplets and droplet clouds. Both experimental work and theoretical modeling are discussed.

Understanding the details of the ignition and combustion of individual liquid droplets is a premise for in-depth understanding of the combustion of collectives of such droplets, i.e. spray and mist combustion.

3.3.2 Laminar or Close-to-Laminar Flame Propagation in Sprays/Mists

In their classical investigation, Burgoyne and Cohen (1954) developed a method whereby suspensions of controlled uniform drop size could be prepared from pure liquids. Using tetralin (see caption of Figure 3–1) as the fuel, it was possible, within limits, to study the effect of the drop size on the combustion properties of a suspension of liquid drops in air. Through a study of limits of flammability, nitrogen dilution limits, and burning velocities, it was shown that the mechanism of flame propagation was completely changed over the drop-size range from 7–40 μm. Thus, below about 7 μm the suspension behaved like a premixed gas, whereas above 40 μm the drops burned individually, in their own air envelope, one burning drop igniting adjacent ones, thus spreading combustion. At intermediate sizes, behavior was transitional.

In a later study Burgoyne (1963) discussed the concept of "concentration" of the droplets in a spray or mist. The following conclusions were drawn:

- For suspensions formed by uncontrolled condensation of saturated vapor (mists), the drop size is small and the lower limit of flammability, measured as mass of liquid per unit volume of air, is essentially the same as that of the vapor. For saturated hydrocarbons this is of the order of 45 g/m^3.
- For suspensions formed mechanically by liquid break-up, the drop size is considerably larger, and in specifying lower limit concentrations account must be taken of the distinction between "static" and "kinetic" concentrations, arising from the sedimentation of the drops.

Burgoyne (1963) also pointed out that in very coarse sprays, with drop diameters between 0.6 and 1.4 mm, normal flame propagation becomes impossible although burning drops may carry flame downwards. However, the presence of smaller droplets in such suspensions may, depending on their concentration, make upward flame propagation possible. Furthermore, the entry of a fast-moving explosion flame (e.g. from a pipe) into a chamber containing a droplet suspension that is normally non-inflammable due to large drop size, flame propagation may nevertheless occur, due to the shattering or break-up effect on the drops of the pressure waves preceding the explosion flame.

Published work on exact burning velocities of explosive sprays/mists is scarce. This may in part be due to the rather complex experiments required, where fuel concentration and droplet size have to be controlled independently. Also, with low-boiling-point liquids, droplet evaporation during droplet cloud generation introduces further complications. Figure 3–1 gives some results from two independent studies of close-to-laminar flame propagation of tetralin droplets in air.

As Figure 3–1 shows, the two independent investigations gave somewhat diverging results for droplet diameters between 10 and 20 μm. However, both studies conclude that there are two different combustion regimes of the tetralin droplets, depending on droplet size. For sizes <10 μm the droplets evaporate completely before combustion gets underway, and the flame propagation process resembles that of premixed gaseous fuel/air. For sizes > 50 μm, however, each droplet burns individually, and flame propagation is more discrete, from droplet to droplet. The distinct peak in vertical burning velocity in the data of Chan and Jou (1988) occurs in the region of drop sizes between these to regimes.

Flame propagation in sprays/mists is considerably more complex than in premixed gases. However, apart from the high values observed by Chan and Jou (1988) in the transition range, Figure 3–1 suggests that close-to-laminar burning velocities of sprays and mists of organic liquids in air are of the same order as laminar burning velocities of saturated gaseous hydrocarbon/air mixtures (alkanes).

3.3.3 Turbulent Flame Propagation in Sprays/Mists

Figure 3–2 shows some results from flame propagation experiments in turbulent clouds of kerosene droplets in air. The Reynolds numbers were of the order of $Re \approx 10^5$. By increasing the flow velocity in the channel (Re is proportional to the flow velocity) by a factor of 3, the turbulent burning velocity (net velocity in relation to unburned cloud ahead of flame front) with 90 μm droplets increased by about the same factor. Also when the droplet size was reduced to 50 μm and further to 30 μm, the turbulent burning velocity increased systematically with flow velocity, but the relative increase got smaller with decreasing droplet size.

Figure 3–3 shows the variation of the turbulent burning velocity with the kerosene/air ratio, for three different droplet sizes. For the smallest droplets the maximum burning velocity occur red at about stoichiometric

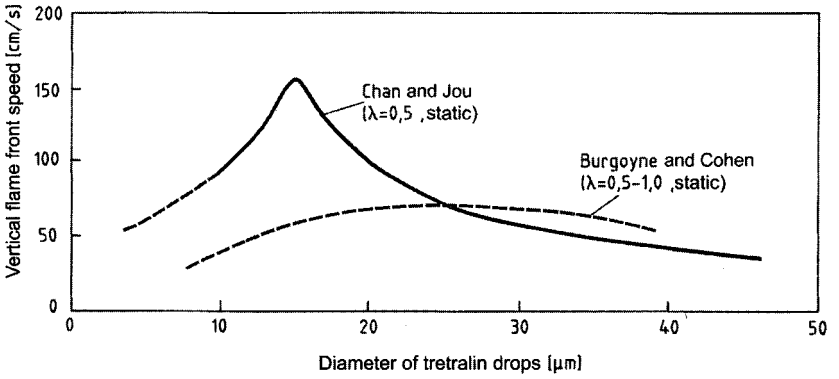


Figure 3–1 Experimental flame speeds (vertical upwards flame propagation) in suspensions of mono-sized tetraline droplets in air, as a function of droplet size. Tetralin is a cyclic organic compound of molecular weight 132, boiling point 207°C, and flash point 71°C. From Burgoyne and Cohen (1954), and Chan and Jou (1988).

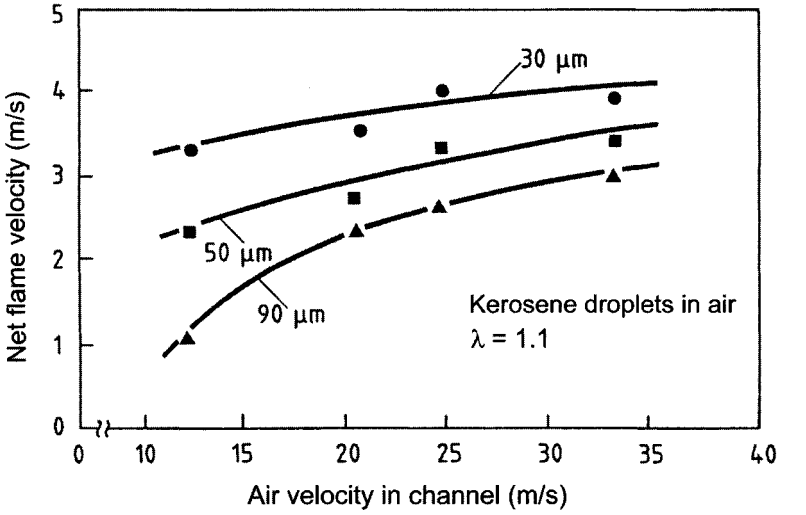


Figure 3–2 Influence of flow generated turbulence on the burning velocity of suspensions of kerosene droplets in air, for various mean droplet sizes. Kerosene is a petroleum distillate with boiling point range 180–300°C, and flash point range 38–66°C. Slightly over-stoichiometric fuel concentration (equivalence ratio $\lambda = 1.1$). Horizontal flow of droplet suspension in a channel of rectangular cross-section (width 30 cm, height 10 cm). From Richards and Lefebvre (1989).

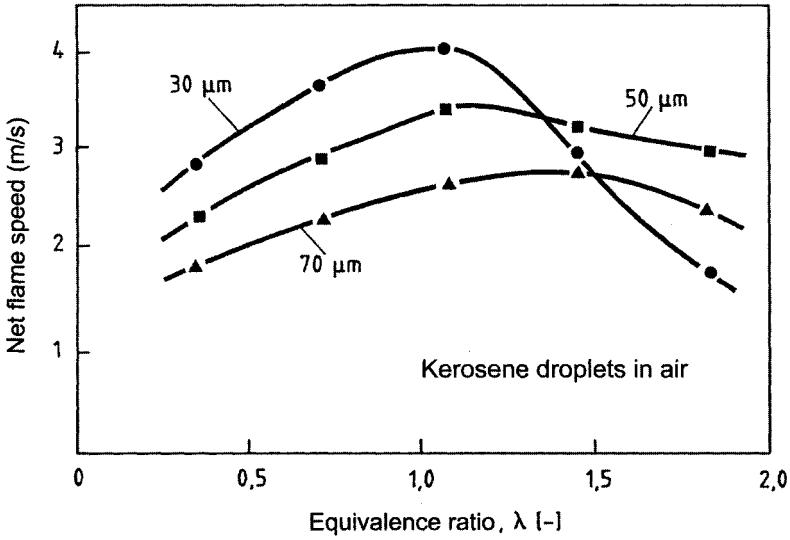


Figure 3–3 Influence of fuel/air ratio and droplet size on burning velocity of suspensions of kerosene droplets in air, flowing at 25 m/s in a horizontal channel of rectangular cross-section (width 30 cm, height 10 cm). Kerosene is a petroleum distillate with boiling point range 180–300°C, and flash point range 38–66°C. From Richards and Lefebvre (1989).

composition ($\lambda = 1.0$), whereas the peak was shifted systematically towards higher kerosene/air ratios ($\lambda > 1.0$) as the droplet size increased.

Figure 3–4 shows that with the prevailing flow conditions and range of λ values studied, the turbulent burning velocities for 30 μm droplets were nearly identical for the three different fuels kerosene, toluene and dekalin. However, other data obtained by Richards and Lefebvre (1989) indicate that for larger droplet sizes the burning velocity tends to decrease with increasing boiling point of the liquid. The complete relationships between burning velocity on the one hand, and degree of turbulence, droplet size, fuel/air ratio etc. on the other, are very complex.

Hansen and Wilkins (2004) conducted a series of interesting spray explosion experiments in a straight horizontal vented laboratory-scale channel of length 2.5 m and 0.3 m x 0.3 m square cross section. They found that:

- Sprays of hexane droplets (high volatility) exploded at least as violently as premixed propane/air.

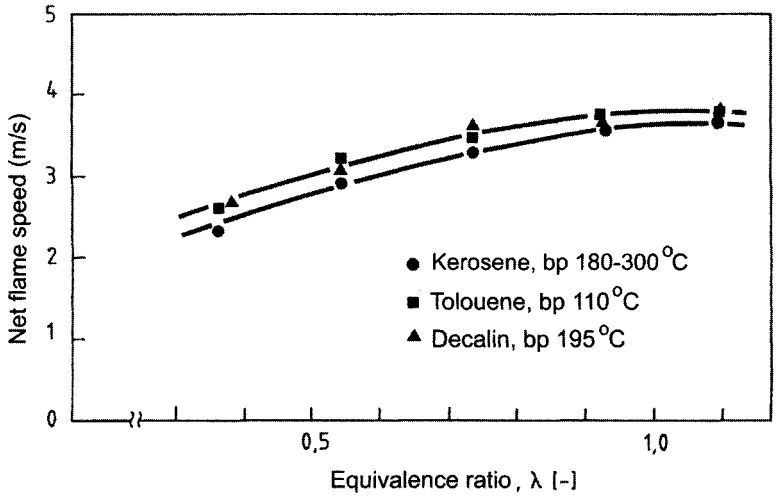


Figure 3–4 Influence of fuel/air ratio on burning velocity of suspensions of 30 µm diameter droplets in air of three different liquid fuels, flowing at 20 m/s in a horizontal channel of rectangular cross-section (width 30 cm, height 10 cm). From Richards and Lefebvre (1989).

- The violence of spray explosions remained high also for over-stoichiometric spray concentrations.
- Mounting of a baffle in the channel raised the maximum explosion pressure by a factor of 8, probably due to droplet break-up by intense flow-induced turbulence.
- Under conditions causing droplet break-up, even sprays of high-flash-point oils exploded violently.
- Worst-case spray concentrations increased with decreasing volatility and increasing droplet size.

3.3.4 Maximum Constant-Volume Adiabatic Explosion Pressures in Sprays/Mists

Experimental data presented by Förster (2000) indicate that the maximum pressures generated in constant-volume adiabatic combustion of clouds of droplets of some organic liquids in air is of the order of 5 bar.

3.3.5 Detonation in Sprays/Mists

Sprays and mists of combustible liquids in air can detonate, depending on fuel chemistry, droplet size and liquid mass concentration. This is not surprising in view of the fact that both homogeneous mixtures of combustible gases and air and clouds of some organic dusts in air can propagate detonation waves. Bull et al. (1979) performed experiments to determine the marginal conditions for spherical detonation in unconfined sprays in air. The sprays were generated in a 5 m^3 plastic foil tent by sonic air-blast atomizers producing measured droplet sizes in the range 5–30 μm. Detonation of the spray cloud was initiated by detonating plastic high explosive charges positioned inside the cloud. Particular attention was paid to studying the influence of fuel volatility on the detonability of the spray. N-hexane and n-dodecane were used to represent extremes of very high and very low vapor pressure fuels. Fuels having intermediate volatilities were synthesized by mixing the two fuels in appropriate proportions.

It was found that n-hexane sprays detonated within fuel/air mass concentration limits similar to those for gaseous detonations, although the measured detonation front velocities were somewhat lower than the calculated Chapman-Jouquet (CJ) values, even with droplet diameters smaller than 50 μm. However, sprays of n-dodecane and n-decane did not detonate, even when using 0.5 kg initiator charges. It was concluded that this was because with these fuels the minimum effective equivalence ratio necessary for detonation propagation was not being attained behind the shock wave within the limiting induction period available. It was further concluded that propagation of self-sustained detonations in n-alkane sprays in the spherical mode, with droplet sizes larger than 10 μm, may require the presence of a certain quantity of fuel vapor prior to ignition.

Smeets (1985) performed experimental spray detonation studies in a 50 cm diameter vertical shock tube. The sprays were stochiometric clouds of droplets in air of methanol, ethanol, propanol, hexanol, decanol, or decane. The droplets had a defined narrow size distribution with a mean diameter of 350 μm. It was found that even in sprays of low vapor pressure fuels, like n-decane and hexanol, self-sustained detonations can exist. However, a minimum concentration of fuel vapor is required, in addition to the droplets, as also indicated by Bull et al. (1979) and discussed above. By measuring several physical quantities within the reaction zone, using laser Doppler techniques, the processes of liquid fuel fragmentation, evaporation, turbulent mixing, and combustion could be observed in

detail. From the data recorded it was concluded that for sprays of 350 μm diameter droplets, the controlling mechanism for the reaction is turbulent mixing of the vapor originating from the droplets, with air.

Tang et al. (1986) wanted to explain the difference between the theoretically predicted and the experimentally observed detonation velocities in clouds of droplets of low-vapor-pressure liquids in air or oxygen. Monosized n-decane droplets of diameter 400 μm were allowed to settle vertically in air or oxygen at ambient pressure and temperature in a vertical shock tube. A shock wave was emitted vertically downwards into the droplet suspension, initiating a self-sustained detonation wave in the cloud. The propagation of this wave was monitored using pressure switches and pressure transducers. Measured detonation velocities were compared with theoretical predictions. Quite satisfactory agreement between experiment and theory was obtained for lean mixtures, but there were quite substantial differences for very rich mixtures. In the latter case, the experimental velocities reached a peak somewhat on the rich side and then decreased very little for richer mixtures.

Förster (2000) gives some references to published work on detonation of sprays and mists, confirming that clouds of combustible liquid droplets in air can detonate in the same way as premixed clouds of gaseous fuels and air. Typical measured detonation velocities are in the range 1,300–1,600 m/s, i.e. somewhat lower than typical values for saturated hydrocarbon gas/air mixtures.

3.4 Ignition of Clouds of Liquid Droplets in Air (Spray/Mist)

3.4.1 Ignition by Hot Surfaces

The problem is of the same nature as discussed in Chapter 2 with hot-surface ignition of gases and vapors. However, it has not been possible to trace any standardized test methods for measuring minimum ignition temperatures of sprays/mists. Figure 3–5 gives some results from hot-surface ignition of suspensions of JET-A fuel droplets flowing through a vertical furnace. Note that the vertical axis in Figure 3–5 covers only 120°C. This figure also gives corresponding data for premixed propane/air, and it is

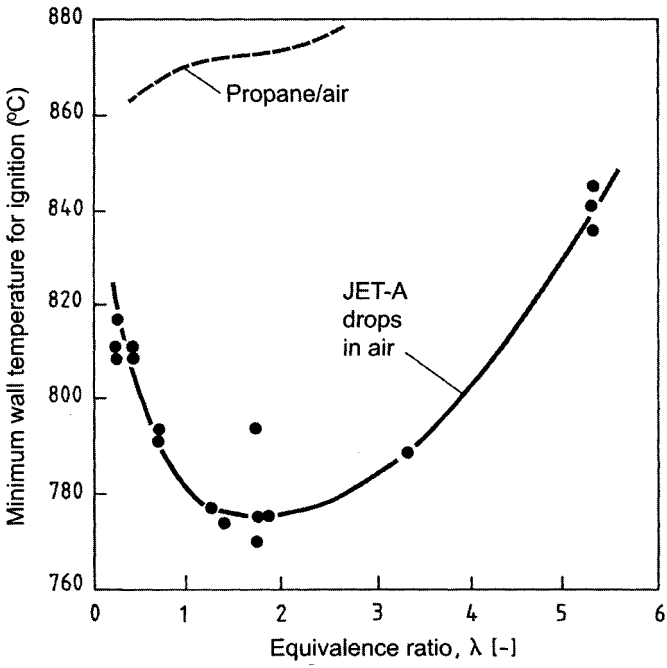


Figure 3–5 Influence of fuel/air ratio on the minimum temperature of a hot wall in a vertical furnace required for igniting a suspension in air of JET-A fuel droplets flowing at 1 m/s. JET-A is a petroleum distillate similar to kerosene (see caption of Figure 3–2). Droplet diameters < 50 μm. Comparison with ignition of propane/air in same apparatus and at same mean flow velocity. From Graves at al. (1986).

seen that the JET-A droplet cloud ignites at significantly lower furnace wall temperatures than the propane/air under the prevailing conditions.

A comparison of the T_{min} values for propane/air in Figure 3–5 with those in Figure 2–17 gives further support to the view that T_{min} for gases and vapors determined by standard flask tests, which are sometimes also applied to sprays/mists, are highly conservative for many practical situations.

Figure 3–6 shows that T_{min} for a cloud of JET-A fuel droplets increases systematically with the velocity of the cloud across the hot surface. As expected, the trend was the same for propane/air, although less pronounced.

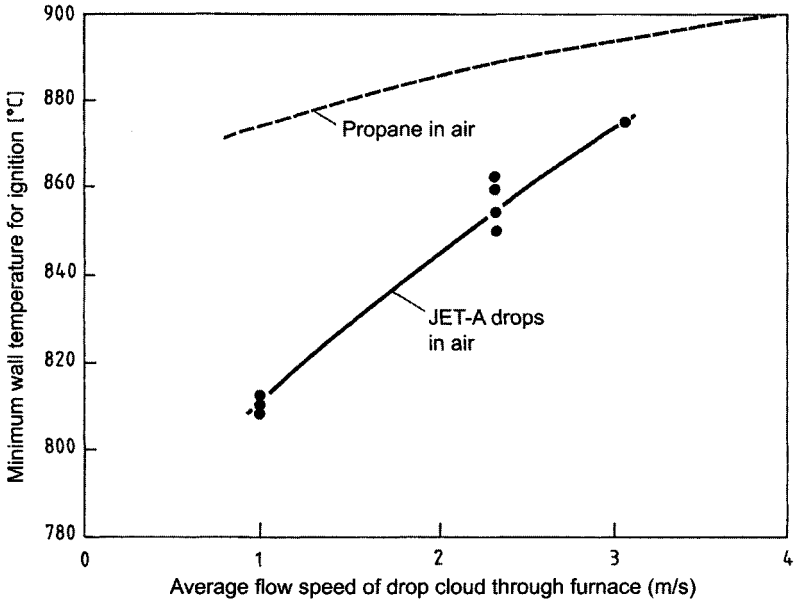


Figure 3–6 Influence of mean flow velocity on the minimum temperature of a hot wall in a vertical furnace required for igniting a suspension in air of JET-A fuel droplets. JET-A is a petroleum distillate similar to kerosene (see caption of Figure 3–2). Droplet diameters <50 μm. Stoichiometric fuel/air ratio. Comparison with ignition of propane/air in same apparatus. From Graves et al. (1986).

3.4.2 Ignition by Electric/Electrostatic Sparks/Discharges

Explosive sprays/mist can be ignited by electric spark discharges in the same way as premixed gaseous fuel/air mixtures can. In principle, therefore, the concepts of Minimum Ignition Energy (MIE) and Quenching Distance (QD) are equally valid for sprays/mists as for premixed gases. However, in the case of spray/mists the experimental determination is considerably more difficult.

Figure 3–7 gives a set of data showing the influence of spark gap length on minimum ignition energy of a suspension of kerosene droplets in air (see caption of Figure 3–2). The data indicate that the minimum ignition energy in this particular case, with the droplet suspension flowing at a fairly high velocity past the spark gap, was about 20 mJ and the quenching distance 3–4 mm.

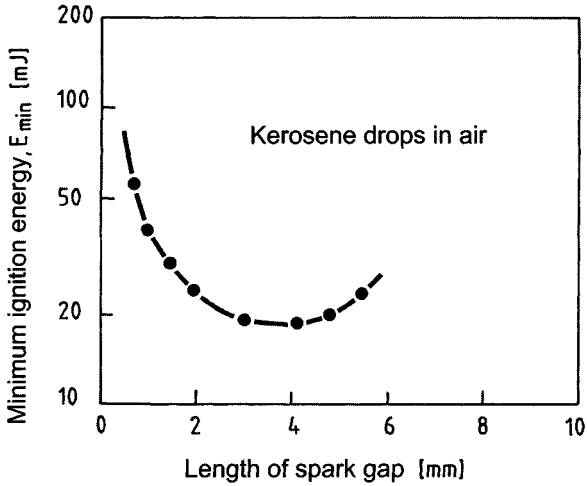


Figure 3–7 Influence of spark gap length on minimum ignition energy of a suspension of kerosene droplets in air at atmospheric pressure and normal temperature. Flow velocity of suspension past spark gap 3 m/s, and equivalence ratio $\lambda = 0.65$ ($\lambda = 1$ for stoichiometric mixtures). From Rao and Lefebvre (1976).

Ballal (1976) performed detailed measurements of the ignition characteristics of kerosene sprays injected into a flowing air stream. Measurements of spark discharge parameters (breakdown voltage, optimum spark duration, energy released in the gap) and ignition characteristics (optimum spark gap, minimum ignition energy, lean ignition limit) were performed. Three different spray generators (atomizers) were used at air velocities up to 38 m/s, producing fuel droplet diameters in the range 20 to 110 μm. The initial pressure was varied between 0.2 and 0.9 bar (absolute), the initial temperatures between 228 and 350 K, and the equivalence ratios between 0.3 and 0.9. The results showed that optimum spark duration for ignition increased with decreasing initial pressure and spray cloud velocity, and with increasing droplet size. Minimum ignition energy decreased markedly with decreasing fuel droplet size.

Figure 3–8, from the subsequent work by Ballal and Lefebvre (1978), shows the variation of MIE for clouds of heavy-oil droplets in air, with the equivalence ratio λ in the range of $\lambda < 1$, and the mean droplet diameter in the range 40–150 μm. As Figure 3–8 shows, MIE for 40 μm droplets of heavy oil was only 2 mJ.

Figure 3–9 gives experimentally determined MIEs for clouds in air of 100 μm droplets of a range of different liquid fuels, as a function of the

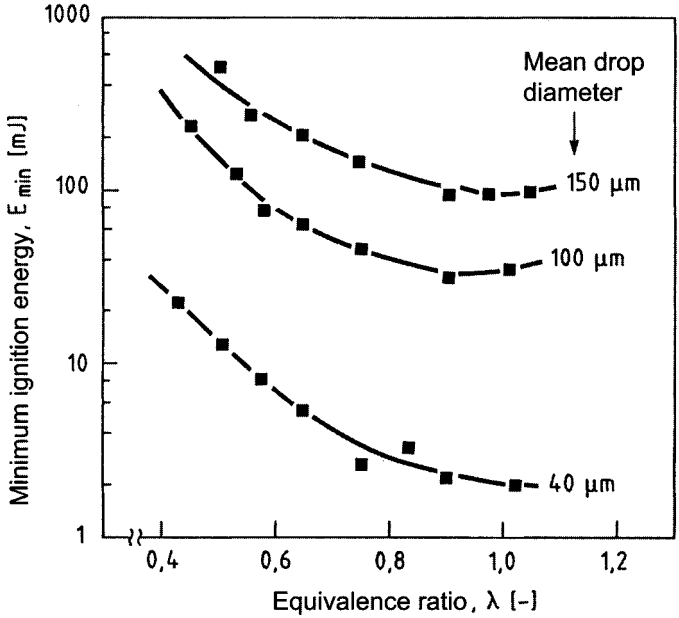


Figure 3–8 Influence of the equivalence ratio λ (fuel/air ratio) and mean droplet diameter on the MIE of close-to-quiescent suspensions of heavy oil droplets in air at atmospheric pressure and normal temperature (λ = 1 for stoichiometric mixtures). From Ballal and Lefebvre (1978).

equivalence ratio λ. Spalding's mass transfer number B, as defined in the figure caption, is used as parameter. B is an amplified function of vapor pressure P_A of the liquid, and Figure 3–9 shows quite clearly that MIE decreases systematically with increasing B, or P_A.

Later Ballal and Lefebvre (1979) extended their studies to measurement of MIE over wide ranges of initial pressures, velocities, turbulence intensities, equivalence ratios, mean droplet sizes, and fuel volatilities. Model predictions showed satisfactory agreement with the experimental data.

3.4.3 Other Ignition Sources

Explosive clouds of liquid droplets in air can also be ignited by fast (adiabatic) compression (diesel engines). Baev et al. (1984) determined induction times for liquid fuel sprays at high temperatures and pressures. The apparatus was a closed cylindrical chamber of 150 mm diameter and

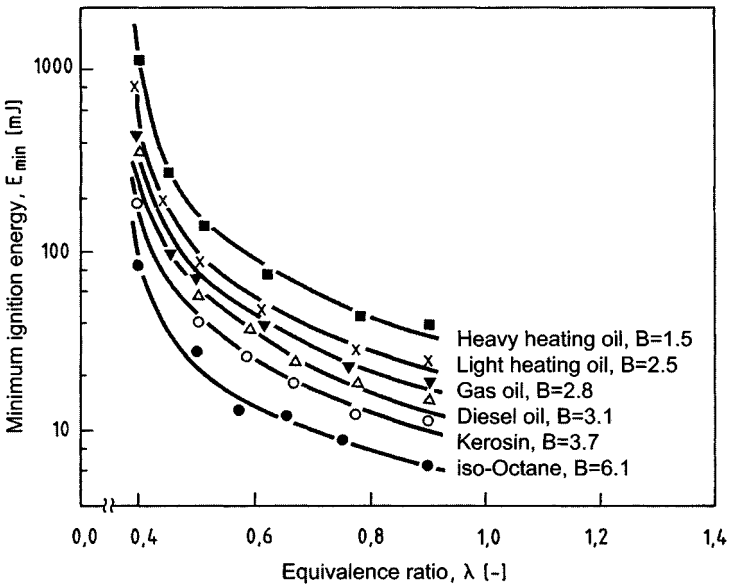


Figure 3–9 Influence of the equivalence ratio λ (λ = 1 for stoichiometric mixtures) on MIE of close-to-quiescent suspensions of 100 µm droplets of various combustible liquids in air at atmospheric pressure and normal temperature. B is Spalding's dimensionless 'mass transfer number' defined as $B = P_A/(P_{tot} - P_A)$, where P_A is the vapor pressure of the combustible liquid, and P_{tot} the total pressure. From Ballal and Lefebvre (1978).

length, filled with air at the desired temperature and pressure. The end plates were fitted with thick quartz windows. The fuel was injected into the chamber during 3–4 ms via a special nozzle system. They also measured the induction time defined as the time interval from onset of fuel injection to onset of the sharp rise in pressure due to combustion in the chamber.

It may also be expected that explosive clouds of liquid fuel droplets in air can be ignited by certain types of metal sparks and thermite reactions from single impacts. However, it has not been possible to trace any experimental data illustrating this.

3.4.4 Standard Test Methods for Ignitability and Explosivity of Sprays/Mists

It has not been possible to trace any standardized methods for assessing ignitability and explosivity of clouds of combustible-liquid droplets in air.

3.5 Case Histories of Spray/Mist Explosions

The main source of this chapter is the review by Eckhoff (1991). Open reports of case histories of accidental spray/mist explosions in the process industries are scarce. Therefore, some spray/mist explosions that have occurred in other contexts have been included in the present section.

3.5.1 Spray/Mist May Have Been Involved in Some Reported "Vapor Cloud" Explosions

Eichhorn (1973) refers, without giving any details or references, to a report, probably from the 1950s, of a mist explosion in the chemical industry, which had prompted an experimental and literature study of mist combustion. The experimental work undertaken suggested that mists of flammable liquids in air can propagate flames, and that the vapor pressure of the liquid is a poor index for judging mist flammability. Spray/mists of chlorinated liquid compounds, however, can be self-quenching.

Quite often the clouds that gave rise to reported vapor cloud explosions were generated by rapid release of combustible liquids from pressurized process equipment. However, such a process may, in addition to producing genuine vapor, also generate sprays of fine liquid droplets that may have played a significant part in the explosion development. The Flixborough disaster in UK in 1974 may be one example (see Section 2.3.4). An investigation conducted by ICI and presented by Kletz (Anonym, 1975) suggested that of all the cyclo-hexane that exploded only one-eighth appeared as true vapor, whereas most of the rest formed a liquid spray. It was pointed out that the spray should be regarded as being just as dangerous (explosive) as the genuine premixed vapor.

3.5.2 Oil Mist Explosion in a Compressor Test Loop

This explosion, discussed by Schmitt (1973), occurred in 1959 in a compressor test facility at Ingersoll-Rand Co.'s plant in Phillipsburg, N.J., USA. The test of a centrifugal compressor was being conducted in accordance with standard procedure, which involved the use of a closed loop for containment and re-circulation of the gas used during the test. The test was nearly completed when, after approximately six hours of test opera-

tion, an explosion occurred. The inlet side of the loop pipe ruptured with great violence, claiming the lives of six men, severely injuring six others, and imposing minor injuries on twenty-four additional men.

The explosion was described as a heavy flash fire accompanied by shock waves. Small localized fires and smoke followed. The floor plates of the platform immediately over the test loop were ripped loose. Heavy brick walls in the immediate vicinity were blown out. The roof of the building immediately over the blast was damaged and moved approximately 30 cm. The lightweight side walls of the building were torn off. Many windows were broken, some of these being located as far as 150 m away from the explosion center. The test loop, the seal oil system, and most of the equipment on the lower level were destroyed or damaged beyond repair. The coupling spacer between the gear and the compressor was shattered, and the coupling end of the gear pinion was twisted off and thrown away to outside the building.

The centrifugal compressor involved was the high-pressure unit of a system employed for compressing natural gas in oil field re-pressuring service. Floating-ring-type oil seals were used to contain the gas within the test loop where the shaft passed through the casing. This design combined the seal with the radial bearings. The sealing medium was oil supplied at a pressure of 3–3.5 bar above the gas pressure in the loop. A small quantity of oil was permitted to pass through the seal for lubrication.

The investigating committee concluded that regardless of the details of the design of this type seal, extremely small leakages of oil into the gas stream are unavoidable. As a result of the high air velocities, which tend to atomize and mix the oil droplets with the gas flow in such a manner that they cannot be effectively drained, the concentration of oil spray in the air tends to increase with time. The definite evidence of a combustion-type explosion confirmed that there must have been enough hydrocarbon oil present within the loop system to produce an explosive oil/air mixture throughout the loop system. Oil had been found to be present when blowing out low points in the piping between test runs. The destructive pressures were produced by heat liberated by the combustion of fine oil spray being circulated by the compressor. The pressure produced by an explosion is approximately proportional to the initial absolute pressure of the explosive mixture. For mixtures of common hydrocarbon compounds in air, the ratio of explosion pressure to initial pressure can be as high as eight.

3.5.3 Spray/Mist Explosions in Crank Cases in Large Diesel Engines

Crank case explosions, their origin and nature, and means of prevention and mitigation were discussed in detail by Minkhorst (1957) in a comprehensive literature survey. Engines fitted with cranks are normally equipped with a lubrication system by which oil is supplied to the bearings by pressurization. In order to protect the environment from oil spray and to prevent loss of oil, the crank system is usually fully enclosed by a case. Explosive spray clouds may be generated within the case by oil being flung into the air from rotating parts. In addition, mists of finer droplets may be generated by evaporation of oil in the hotter parts of the crank case and subsequent circulation to and condensation in the colder parts. If ignition occurs, the ignition source would normally be a hot surface, generated e.g. by a faulty bearing or a cylinder liner where the clearance between piston and liner is too narrow. In the early stages, before the mist concentration has reached explosive levels, the hot surface may act as a mist generator by enhancing evaporation of oil that is subsequently condensed to mist in the colder parts of the system. If the temperature of the hot surface is very high, the oil may be subject to chemical cracking, and highly explosive vapors may also be formed.

A severe crank case explosion, discussed by Minkhorst (1957), occurred in September 1947 in all the four engines of the ship *Reina del Pacifico*. The first warning of a potential hazard was given by some maintenance people who had noticed that a cylinder liner of one of the engines was overheated. Somewhat later a strong explosion occurred in the same engine, which subsequently propagated into the other three engines of the ship. Four distinct explosions were heard at very short intervals. The engines blew up, and twenty-eight men were killed and twenty-three were injured. Minkhorst also discussed a number of experimental investigations related to initiation and propagation of crank case explosions, and to various possible means of preventing and mitigating such explosions.

According to Nordahl (1991), crank case explosions in ship engines have also occurred more recently. In fact, the frequency of such explosions seems to have increased rather than decreased since the time when Minkhorst conducted his investigation. The chain of events leading to the explosions seems to be the same as in the past, the ignition source being

unintentionally heated surfaces. Large 12-cylinder ship engines of up to 60,000 Hp have crank cases of 18–20 m length and 150 m^3 volume. It seems that crank case explosions occur more frequently with 4-stroke engines than with 2-stroke ones.

3.5.4 Major Oil Spray/Mist Explosion in a Transformer Room of a Hydroelectric Power Station

This accident, in which three men were killed and several others injured, was described by Johnsen and Holte (1973). Further details were given by Schjelderup (1990). The accident occurred in 1973 in the hydroelectric power station at Tonstad, Sirdal in Norway. The entire station is located underground. Figure 3–10 gives a top view of the entire underground installation.

Figure 3–11 gives a cross-section of the oil-filled cable junction box, and Figure 3–12 a cross-section of the main underground hall with the control rooms and a transformer with junction box on top in a separate room.

The origin of the event was a flash-over inside a 2 m^3 oil-filled cable junction box of one of the transformers (see Figure 3–11). The probable reason for the flash-over was the presence of significant dissolved quantities of a sulphur-containing cement in the oil, which, in combination with moisture, reduced the flash-over resistance of the oil and gave rise to electrical breakdown. The substantial thermal power of the flash-over arc caused a sudden and substantial rise of the internal pressure in the junction box. This caused the box to rupture and the oil to be expelled as a fine spray throughout the un-vented concrete cell of volume 800 m^3, in which the high-voltage transformers were located. The quantity of oil in the 2 m^3 junction box was more than sufficient to supply the entire 800 m^3 volume of the transformer cell with an oil mist cloud of the most explosive concentration. The mist cloud exploded immediately (see Figure 3–13 and Figure 3–14), ignited by the arc in the junction box, and the wall of the transformer cell was blown out. A strong blast wave swept through the large engine hall and blew in the windows of the central control room. Dense smoke penetrated the entire system including the road tunnels to the open, rendering the rescue operation very difficult.

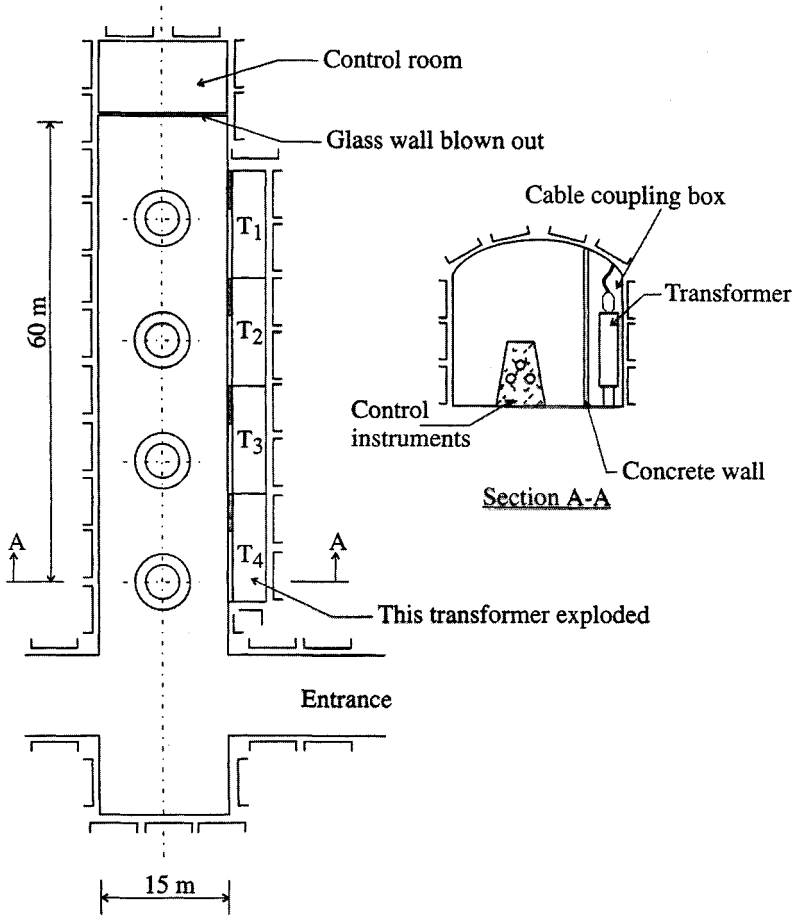


Figure 3–10 Top view, with side view of one section, of the underground hydroelectric power plant at Tonstad, Sirdal, Norway, with vertical section through the main hall, the control instruments above the dynamos, and the transformer in a separate room on the side of the main hall. From Schjelderup (1990).

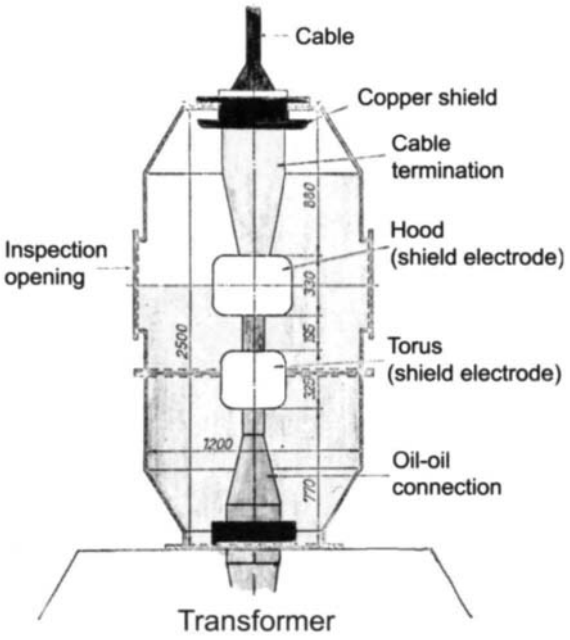


Figure 3–11 Cross-section of the oil-filled cable junction box that exploded in the underground hydroelectric power plant at Tonstad, Sirdal, Norway.

3.6 Means of Preventing and Mitigating Spray/Mist Explosions in the Process Industries

3.6.1 Preventing and Limiting Size of Explosive Clouds

Although the physics, not least the fluid dynamics, of mists and sprays differs appreciably from the physics of homogeneous gases, it is customary to apply the principles developed for gases even to mists/sprays. These are described in Section 2.4.2. This may be a reasonable approach in the case of liquids of low boiling points, in the case of which the droplets will evaporate quickly once released into the open air. But, in the case of liquids of high boiling points, the approach of adopting the gas principles may not always seem reasonable. However, published specific guidance for sprays/mists does not seem to exist.

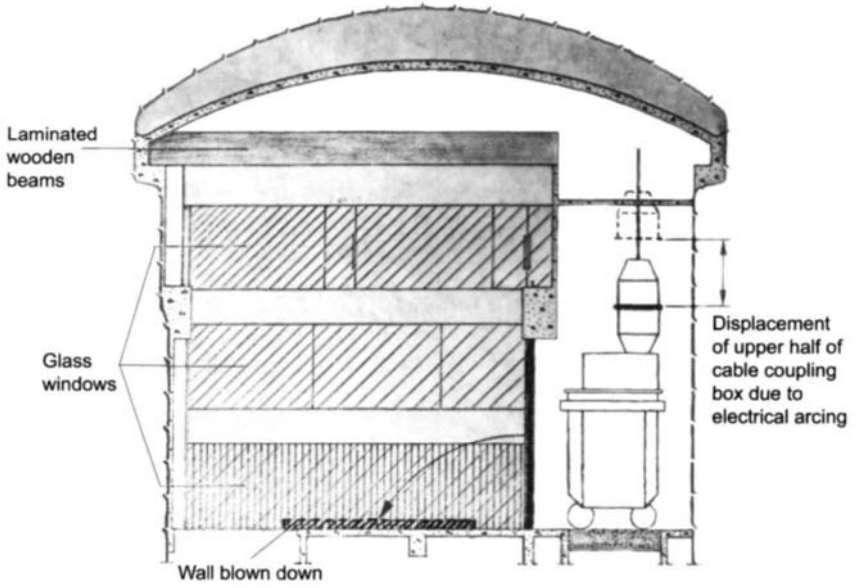


Figure 3–12 Cross-section of the main hall of the underground hydroelectric power plant at Tonstad, Sirdal, Norway, with the separate transformer concrete cell to the right.

Figure 3–13 Photo of debris after the oil spray/mist explosion in the main hall of the underground hydroelectric power station at Tonstad, Sirdal, Norway, illustrated in Figure 3–10. The conical cabinets of the control instruments are seen clearly.

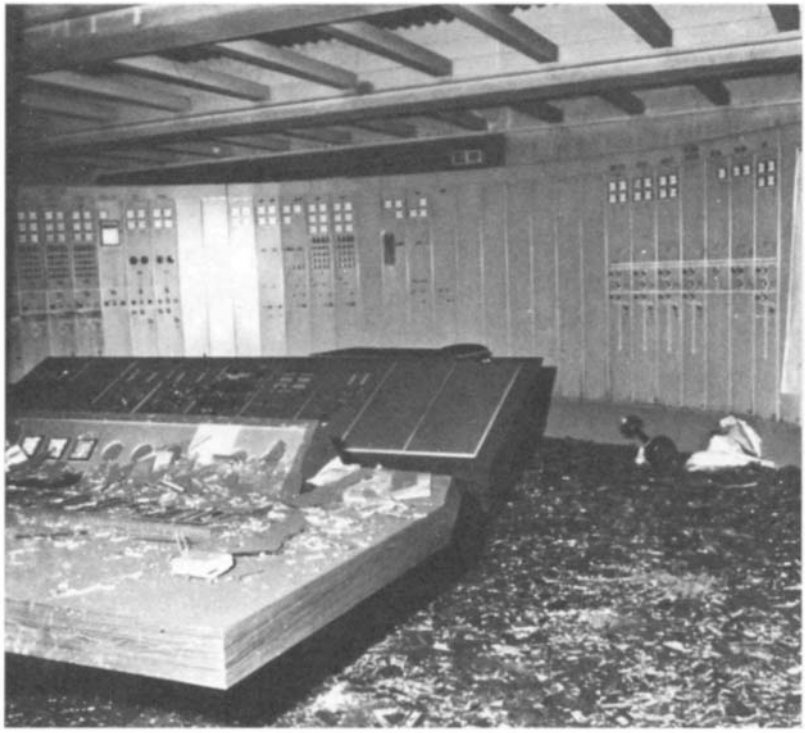


Figure 3–14 Photo of debris and damage in the control room after the oil spray/mist explosion in the underground hydroelectric power station at Tonstad, Sirdal, Norway.

3.6.2 Preventing and Controlling Ignition Sources

The same overall comment as in Section 3.6.1 applies. Guidance for explosive gases is given in Section 2.4.3 and Section 2.4.4.

3.6.3 Mitigatory Measures

The same overall comment as in Section 3.6.1 applies. Guidance for explosive gases is given in Section 2.4.5.

Chapter 4

Gas and Dust Explosions Caused by Smoldering Combustion in Powder Layers and Deposits

4.1 Combustion Velocities/Rates in Powder Layers/ Deposits

4.1.1 Dust Layers

Propagation of combustion through layers and deposits of most dusts will occur very slowly compared with the propagation of a flame through an explosive gas/vapor cloud. This is because the ratio of combustible material/oxygen in a dust layer/deposit is about two orders of magnitude larger than the stoichiometric ratio. Very limited supply of oxygen to support the combustion process, and the heat sink effect of all the excess material that does not take part in the combustion, will limit both the rate of propagation of the combustion process, and the temperature in the reaction zone. Therefore, for most dusts the combustion zone will normally be a smolder or glow rather than an open flame, and typical velocities of the combustion wave will be several orders of magnitude lower than laminar burning velocities in pre-mixed gases and dust clouds.

Figure 4–1 shows some data reported by Palmer (1973) from experiments with glowing/smoldering combustion in horizontal layers of beech dust.

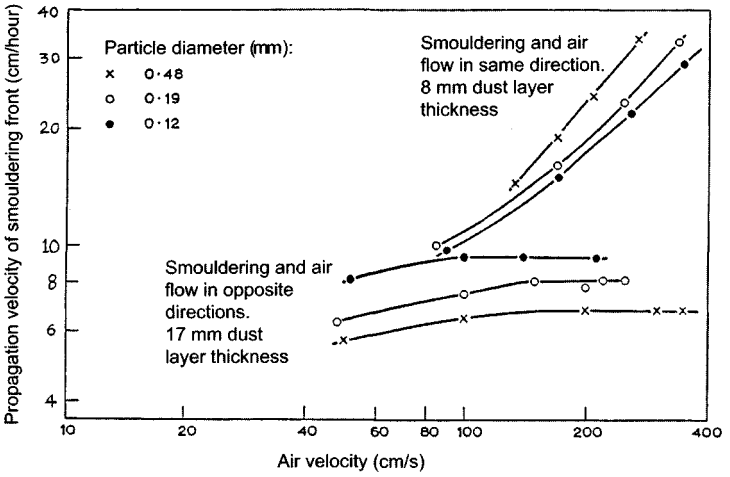


Figure 4–1 *Influence of the air velocity along a layer of beech saw dust on the rate of smoldering front propagation along the layer. From Palmer (1973).*

The layers, of various thicknesses, were ignited at one end and the velocity of the propagating smolder/glow front along the layer was measured as a function of layer thickness, particle size, and the velocity of a horizontal air flow along the surface of the layer. Figure 4–1 exhibits the following trends:

- The observed smolder/glow velocities were in the range of 0.05–0.5 m/hour, as opposed to the order of 0.5 m/s for laminar burning velocities of clouds of organic dust in air (see Chapter 5), i.e. a difference by a factor of the order of 10^4.
- The smolder/glow velocity along the layer increased markedly if an air flow of velocity of some m/s was passed across the layer in the same direction as the propagation of the smolder/glow.
- If the direction of the air flow was opposite to that of the propagation of the smolder/glow, there was no significant influence of the air flow in the velocity range 0.4–4 m/s investigated.

- The influence of particle size on the effect of the air flow was somewhat complex. With the air flow in the same direction as the smolder glow propagation, the latter increased with increasing particle size, whereas with the air flow in the opposite direction the highest smolder/glow velocities was found for the smallest particles. However, in the range investigated the influence of particle size was only modest.
- With high air velocities across the layers, in the direction of the propagation of the glow/smolder, the temperatures in the reaction zone inside the layers were quite high, approaching 1,000°C.
- However, in the absence of an air flow across the layer, there was no visible sign of smolder or glow at the surface of the layer, and the temperature of the smoldering/glowing reaction zone inside the layer was in the range 550–620°C.
- There was little influence of the material of the surface on which the dust layer was resting (asbestos, metal, wood), on the velocity of propagation of the smolder/glow in the layer.

Figure 4–2 illustrates a standardized method for characterizing the ability of layers of powders/dusts to propagate smolder, glow, and in some cases even an open flame.

The foundation of this method was laid by Lütolf (1971). A complete description was given by Verein deutscher Ingenieure (1988). For tests at ambient temperature, a ridge of the dust of triangular cross-section is placed on a ceramic plate, as shown in Figure 4–2(a). The test sample is a 2 cm wide and 20 cm long. For tests at elevated temperatures the sample holder shown in Figure 4–2(b) is used. The sample is then placed in a glass tube heated to the desired temperature. A small air flow of about 0.2 m/s through the glass tube must be ensured.

The dusts are classified according to their ability to propagate combustion at one end of the ridge. Ignition along the ridge tested is accomplished using either a gas flame or a glowing platinum wire at 1,000°C. The definitions of the various flammability classes are:

Class 1: No self-sustained combustion

Class 2: Local combustion of short duration

Class 3: Local sustained combustion, but no propagation

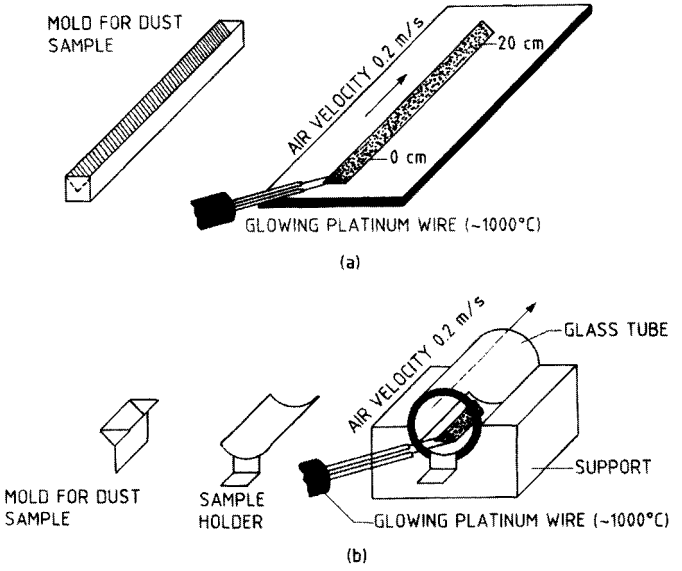


Figure 4–2 Semi-quantitative laboratory method for testing the ability of a horizontal ridge of a given powder/dust to propagate combustion. From Eckhoff (2003).

Class 4: Propagating smoldering combustion

Class 5: Propagating open flame

Class 6: Explosive combustion

4.1.2 Deep Dust Deposits

Figure 4–3 gives the results from investigation of smoldering/glow propagation in deep deposits of beech dusts, also reported by Palmer (1973).

In this case the thickness of the dust deposits was much larger than of the layers discussed above, ranging from 4 cm to nearly 100 cm. The main measurement taken was the time from initiation of the smolder/glow at the bottom of the deposit, till the smolder/glow front reached the deposit top. The following was found:

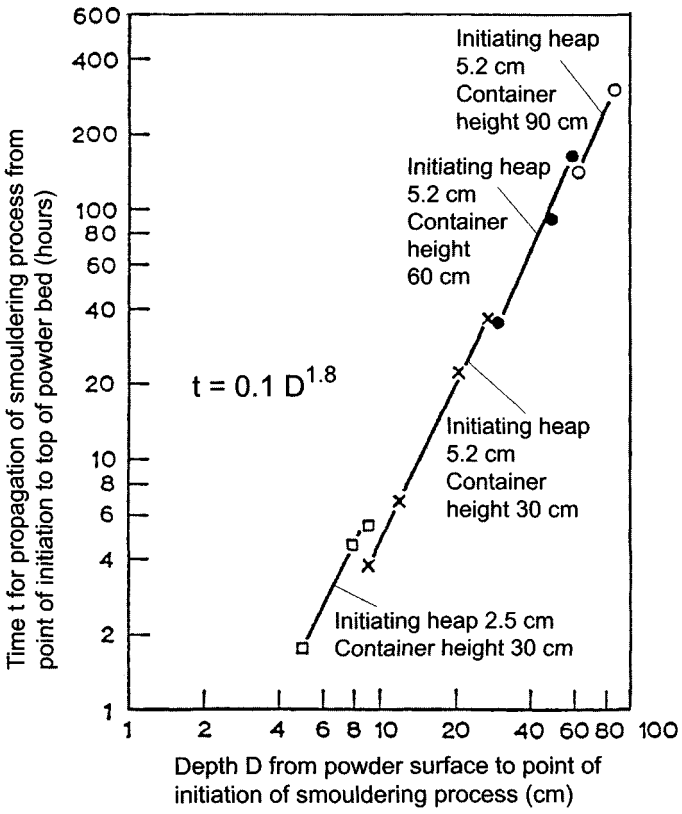


Figure 4–3 Influence of the depth from the top of a deposit of wood saw dust to the point of initiation of smoldering/glow inside the deposit, on the time from initiation to arrival of the smoldering/glow front at the deposit top. From Palmer (1973).

- The average velocity of propagation of the smolder/glow front, from initiation at the bed bottom, to arrival at the bed top, decreased systematically with increasing bed height, according to the approximate relation: $t = 0.1H^{1.8}$, where t is in hours and H is in cm.
- With a deposit height of 85 cm, the propagation of the smolder/glow front from the bottom to the top of the deposit took about 350 hours, i.e. 14–15 days. The average rate of propagation was about 2.5 mm/hour, which is about twenty times less than the lowest layer propagation velocities of about 5 cm/hour given in Figure 4–1.

- As soon as the smolder/glow has reached the bed top, it will start to propagate sideways at the higher velocities typical of horizontal propagation in thin layers.

Smolder/glow processes can propagate inside dust deposits of considerably larger dimensions than about 1 m. The propagation velocities will then be even lower than those typical of the experiment leading to the data in Figure 4–3. In extreme cases, e.g. in huge outdoor storages of coal the times needed for the smolder/glow front to reach the borders of the powder/dust deposit can be up to several years.

The main reason why smolder/glow processes inside large powder/dust deposits are so slow is lack of supply of oxygen to the exothermal chemical reaction. The oxygen has to be transported to the reaction zone by diffusion through the narrow channels between the particles in the deposit. The exothermal process is nevertheless maintained because of the inherent thermal insulation capability of the powder/dust deposit. The very limited amount of heat generated per unit time in the chemical reaction is used effectively for heating the next layer of dust, adjacent to the combustion zone, without being drained into the farther surroundings. The basic principle is the same as that of the general theory of ignition discussed in Section 2.2.2 and illustrated in Figure 2–13.

Besides promoting the development and sustainment of smolder/glow in powder/dust deposits, the high thermal insulation power of the powder/dust creates substantial problems in the efforts to extinguish smoldering/glow processes in large powder/dust deposits, e.g. in silos. Often the main problem is not to extinguish, i.e. to stop the exothermal reaction, but to cool down the hot material to normal temperatures.

Figure 4–10 illustrates a more complex scenario, whereby the smolder/glow process in a silo generates CO, which diffuses upwards through the powder/dust bed and mixes with the air in the empty space above the bed to form an explosive mixture. This mixture is then, in turn, ignited by the glow front as soon as it reaches the top of the powder/dust bed. Further escalation to a secondary dust explosion can occur if dust layers have accumulated in area above silo top.

4.2 Initiation of Combustion in Powder/Dust Layers and Deposits

4.2.1 Minimum Ignition Temperature of Dust Layers

The apparatus shown in Figure 4–4 is often used for determining minimum ignition temperatures of dust layers. It consists of an electric hot plate, a temperature control unit, three thermocouples and a 2-channel recorder.

The hot plate is kept at a given temperature, which is read by one of the thermocouples and displayed on one of the recorder channels. The second thermocouple is used for regulating the plate temperature. When conducting a test, a metal ring of internal diameter 100 mm and height either 5 mm or 15 mm is first placed on the surface of the plate. The powder/dust sample to be tested is then placed on the part of the hot plate that is inside the metal ring and carefully leveled off to the height of the ring. The third thermocouple is placed in the sample through holes in the metal ring. The sample temperature is displayed on the second recorder channel.

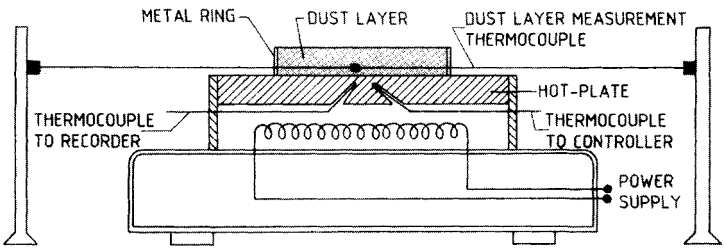


Figure 4–4 Internationally standardized method for determining the minimum hot-plate temperature required for initiating smoldering or flaming combustion of layers of combustible powders/dusts. From Eckhoff (2003).

Typical outcomes of a test are illustrated in Figure 4–5.

Originally it was proposed that the maximum temperature in the dust sample had to exceed the hot plate temperature by more than 20°C for the test to be recorded as ignition. However, it was later decided that a considerably higher temperature rise was required. In order to determine the minimum ignition temperature for the layer thickness tested, repeated

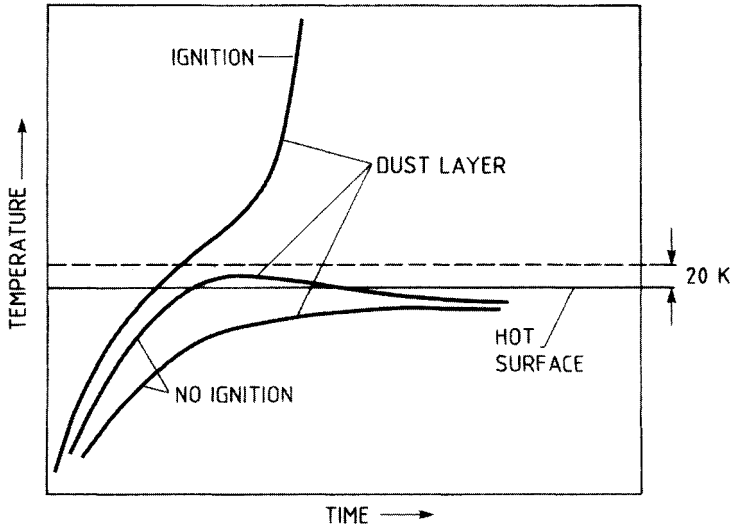


Figure 4–5 Illustration of the three possible outcomes of the test illustrated in Figure 4–4. From Eckhoff (2003).

tests are conducted at different hot plate temperatures until the minimum temperature for ignition has been identified. It is important to note that the minimum hot plate ignition temperature decreases systematically with increasing dust layer thickness. If the values for two different layer thicknesses have been determined, simplified theory enables estimation of the values for other thicknesses.

4.2.2 Minimum Ignition Temperature of Dust Deposits

In this case one considers a powder/dust deposit completely surrounded by air, and the question asked is "what minimum temperature of the air surrounding the deposit is required for self-heating/self-ignition to develop inside the deposit?" The type of apparatus used for determining the minimum ignition temperature in this case is illustrated in Figure 4–6.

The power/dust sample is suspended in a metal gauze basket inside a heating chamber through which a given flow of pre-heated air is circulated. The air temperature and the temperature inside the powder sample

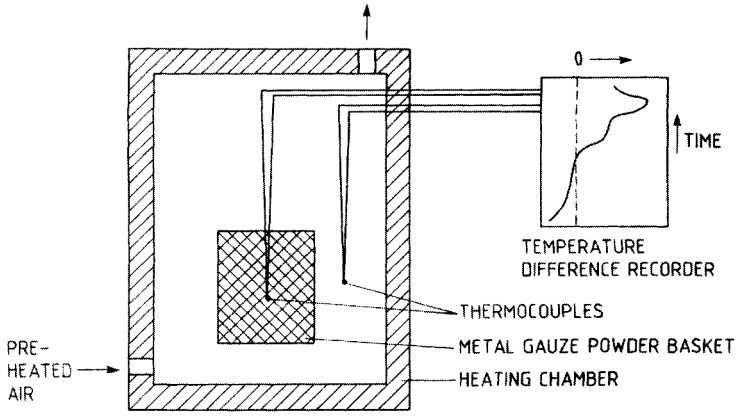


Figure 4–6 Illustration of method for determining the minimum temperature of the surrounding air for initiation of smoldering combustion in a dust sample contained in a wire net basket. From Eckhoff (2003).

are measured and the difference recorded. If the temperature in the powder sample raises beyond that of the air, but not higher than 400 °C, the phenomenon is called "self heating." Temperature rises beyond 400 °C are named "self ignition." Figure 4–7 illustrates the typical linear relationship between the minimum ignition temperature and the sample size resulting from plotting the experimental data in a double-logarithmic form. This enables extrapolation to harger volumes then those tested. More sophisticated instruments for detailed studies of the rate of reaction as a function of temperature, under adiabatic conditions, are also available, including *Accelerating Rate Calorimeters* (ARC), which are essentially highly computerized adiabatic calorimeters. During an ARC experiment, the sample is maintained in a near to perfect adiabatic condition, while time, temperature and pressure data are automatically collected and stored. The data can then be processed by computers. In addition to ARC, *Differential Scanning Calorimeters* (DSC) are also in use. In a DSC the rate at which heat must be transferred to or from the test sample in order to maintain it at the same temperature as an inert reference sample, is measured.

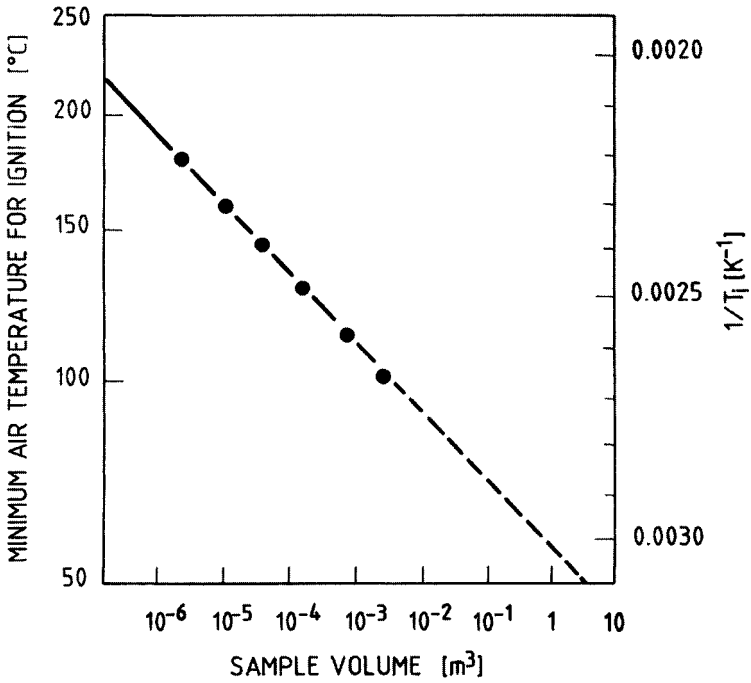


Figure 4–7 Example of results from the experiment illustrated in Figure 4–6 for a given powder/dust, using a range of sample volumes. From Eckhoff (2003).

4.3 Explosion Initiation Processes by Smoldering/Glowing Material

Smolder/glow processes in powder/dust deposits are in themselves a significant problem in some process industries. However, in some cases these processes are only the first step in an escalating chain of events that can lead to both gas and dust explosions. Figure 4–8 and Figure 4–9 illustrate how smoldering powder/dust layers/deposits can give rise to dust explosions.

Figure 4–10 illustrates a more complex scenario, whereby the smolder/glow process in a silo generates CO, which diffuses upwards through the powder/dust bed and mixes with the air in the empty space above the bed to form an explosive mixture. This mixture is then, in turn, ignited by the

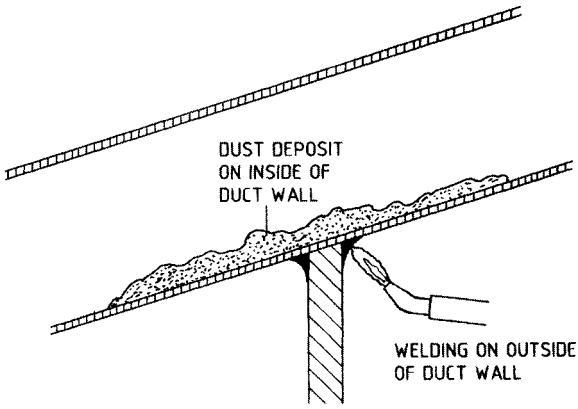


Figure 4–8 Initiation of combustion in a powder/dust layer inside a steel duct, by a welding operation on the outside of the duct. With an explosive dust cloud flowing through the duct, the glowing/flaming dust layer can initiate a dust explosion there. From Eckhoff (2003).

glow front as soon as it reaches the top of the powder/dust bed. Further escalation to a secondary dust explosion can occur if dust layers have accumulated in area above silo top. Case histories confirming this chain of events are given in Section 4.5.

4.4 Case Histories of Accidental Smoldering Combustion in Silos, Resulting in Gas Explosions, and Adopted Extinction Procedures

4.4.1 Fire and Subsequent Explosion in a Silo Plant in Stavanger, Norway (1985)

This accident, described by Braaten (1985), was primarily an explosion of combustible gases released from a solid organic material during self-heating in a silo cell.

The cause of events was in accordance with Figure 4–10. The explosion occurred in a fairly modern reinforced concrete silo complex used for storage of various feed stuffs. Pellets of Canadian rape seed flour had been stored in one of the silos for some time when it was discovered that

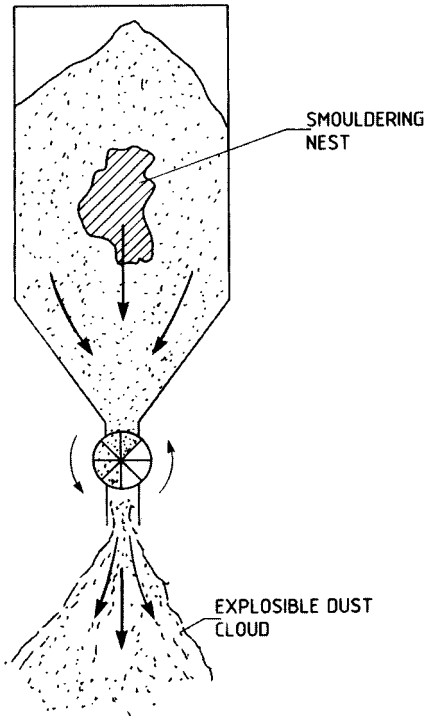


Figure 4–9 Initiation of dust explosion in cloud generated by discharge of powder/dust from silo/hopper, by smoldering nest in powder/dust. From Eckhoff (2003).

the material in the bottom part of the silo had become packed to a solid mass and could not be discharged through the silo exit. Some time later, one week before the explosion, flames were observed in the silo. The fire brigade was called and covered the pellets in the silo with foam from above. Various unsuccessful attempts were then made at discharging the pellets mass at the silo bottom. During this phase there was considerable development of smoke, which mixed with the air not only in the silo cell in question, but also in the silo loft above the cells. It is probable that the smoke contained combustible gases, e.g. CO, and that the strong explosion that occurred just after the top of the pellets had been covered with foam once more, was mainly a gas explosion. However, any dust deposits in the loft may also have become involved. The entire roof of the building was blown up, and debris was thrown into the surrounding area. Because the explosion occurred in the middle of the night (0300), and just after the fire brigade had left, nobody was killed or hurt.

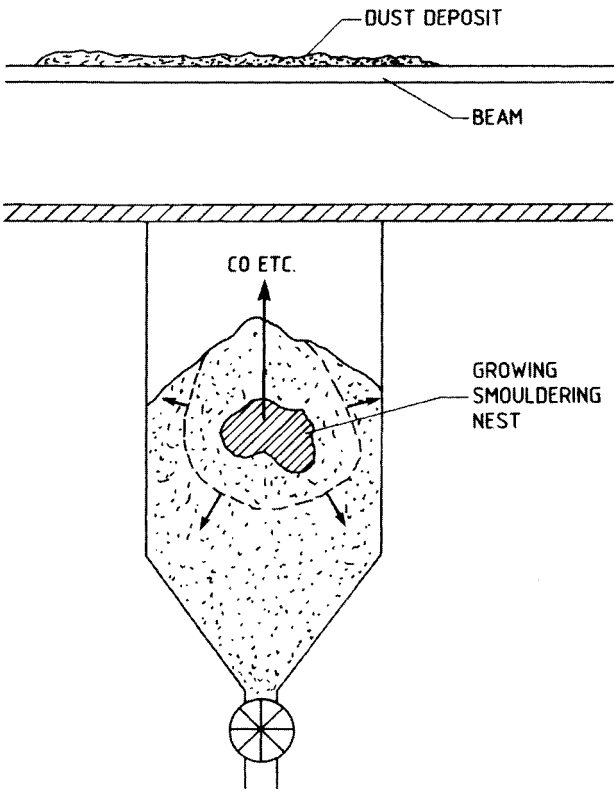


Figure 4–10 Initiation of explosion in a gas/air mixture developed by a smoldering nest inside powder/dust in a silo, and accumulated above the powder/dust in the silo, when the top surface of the smoldering nest reaches the top of the powder/dust. From Eckhoff (2003).

4.4.2 Multiple Silo Fires and Resulting Gas Explosions in a Large Storage Facility in Tomylovo in the Knibyshev Region in Russia (1987-89)

This extensive series of explosions were of the same nature as the explosion discussed in Section 4.4.1. An oral report of the event was provided by Borisov and Gelfand (1989). The large storage facility for grain and feed stuffs, shown in Figure 4–11, consisted of four sections of sixty silo cells each, i.e. 240 silo cells altogether. Each cell was of 3 m × 3 m square cross-section and 30 m height.

Figure 4–11 The concrete silo complex in Tomylovo, Knibyshev region, Russia that was damaged by a series of smoldering gas explosions in the silo cells. (Courtesy of A. Borisov and B. Gelfand, Moscow, Russia).

The first explosion occurred in December 1987 in a silo cell containing moist sunflower seed, which was not supposed to be stored in such silos due to the risk of self-heating. However, this had nevertheless been

done, and the resulting self-heating developed into extensive smoldering decomposition during which methane and carbon monoxide were produced and mixed with the air in the empty top part of the silo, above the powder bed surface. It is reasonable to believe that the primary explosion was in this mixture of explosive gas and air, and that the ignition source was the smoldering combustion when it penetrated to the powder bed top surface, as illustrated in Figure 4–10. However, in addition dust deposits on the internal silo walls and roof may well have been entrained by the initial blast and become involved in the explosion.

This was only the first of a large series of 20–30 subsequent explosions that took place in the same facility, in one silo cell after the other, during 1988 and 1989. There are two main reasons for this continued explosion activity in the silo complex. The most important is the heat transfer from a silo cell in which smoldering combustion is taking place, to the neighboring cells. Such heat transfer was facilitated by the large contact surface area between the cells due to the square cross-section of the cells. The second main reason for the repeated explosions was that sunflower seed was not the only material in the facility that was not supposed to be stored there. Some of the silo cells contained buckwheat and wheat grain of higher moisture contents than the maximum permissible limits for storage in such facilities.

During the period of repeated explosions, a series of attempts were made at breaking the unfortunate chain of events. Cells were opened at the top for inspection. However, this admitted fresh air to the smoldering mass and enhanced the combustion process. Attempts were made at quenching and cooling the powder mass by liquid nitrogen, but this was only partly successful.

It was finally decided to demolish the entire facility using explosives.

4.4.3 Fire and Explosion in Pelletized Wheat Bran in a Silo Cell at Nord Mills in Malmö, Sweden, in 1989

A cross-section of the silo cell is shown in Figure 4–12.

The course of events, as recorded by Templin (1990), was as follows:

Saturday 28th January, 0700: The night shift stopped the production for the weekend according to schedule, and all activity in the grain silo plant terminated.

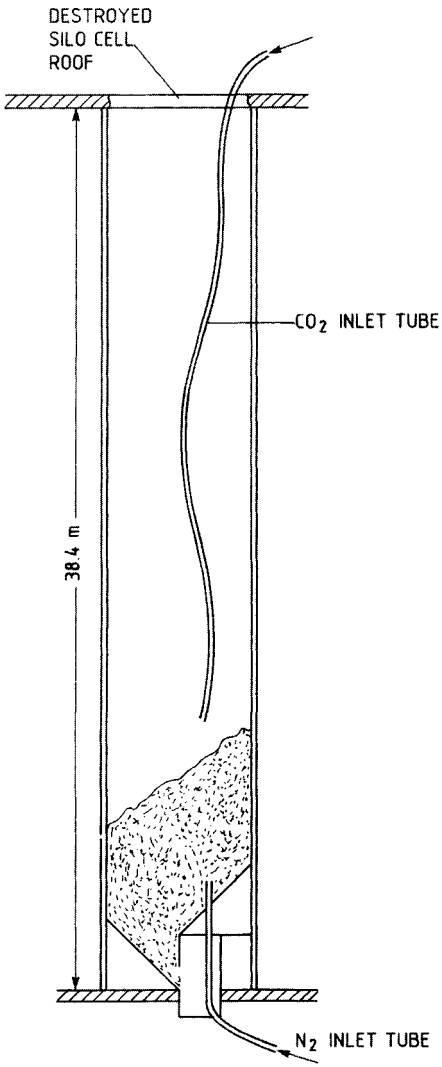


Figure 4–12 Cross-section of silo in Malmö, Sweden, in which a destructive explosion occurred in gas (CO) developed from smoldering combustion of wheat bran pellets in 1989. The inlets for subsequent supply/injection of carbon dioxide and nitrogen for extinction and cooling are also indicated. Courtesy of G. Templin, Nord Mills, Sweden.

Saturday 28th January, 1000: According to Nord Mills' safety procedures, the safety guard team made its inspection round through the entire plant. Nothing special was observed. No persons were encountered.

Saturday 28th January, 2300: A bang, muffled by the noise of strong winds, was heard in the neighborhood, but no action was taken.

Sunday 29th January, 0930: During its scheduled inspection round, the safety guard team discovered fragments of shattered window panes spread over the entire yard. Inspection of the loft of the silo building revealed that the roof of an intermediate star cell had blown up, and that dense smoke was emerging from the open cell top. The height of the cell involved was about 36 m and its cross-sectional area about 20 m^2. Most of the silo was empty, the pelletized wheat bran occupying only the seven first meters above the cell bottom.

Sunday 29th January, rest of day: Fire brigade and other personnel were called, and the entire plant area was cordoned off. About 2,000 kg of gaseous carbon dioxide was pumped into the burning silo from above through a long vertical pipe extending right down to the surface of the smoldering pellets.

Monday 30th January, early morning: The discharge valve at the cell bottom was removed, and discharge of the pellets mass, using a mobile suction unit, was started. This gave rise to increased smoke production, and at 0330 more carbon dioxide was loaded into the silo cell from above.

Tuesday 31st January: The discharge operation was interrupted. Carbon dioxide was emerging through the bottom silo exit, and more was loaded into the silo at the top.

Wednesday 1st February: More carbon dioxide was loaded into the silo at the top. From 0300 to 1250 the smoke development was enhanced by vibrations due to operation of another silo cell. The smoke temperature just above the pellets was 96°C, and just above the silo top 45°C.

Thursday 2nd February–Wednesday 8th February: Smoke development and temperature rise was suppressed temporarily by loading several tons of carbon dioxide into the silo from the top, but there was only slow permanent progress. Temperature rise was observed in the material stored in the four larger adjacent silo cells.

Thursday 9th February–Saturday 11th February: Holes were drilled through the silo bottom and at intervals a total of several tons of nitrogen

were pumped into the pellets from below, while carbon dioxide was charged from above.

Monday 13th February–Wednesday 15th February: Some 6,000 kg of carbon dioxide and 3,000 kg of N_2 was injected into the burning pellets. Temperatures in the burning and adjoining cells and contents of oxygen, CO and CO_2 in the gas above the pellets, were monitored regularly.

Monday 20th February: The smoldering combustion in the wheat bran pellets had finally been brought to an end.

This case history illustrates that fighting smoldering combustion in large silo complexes is not only a matter of quenching, or terminating the oxidation reaction. It is also indeed a matter of cooling massive bulks of poor heat conductors to a temperature level at which the combustion process will not start again once air is re-admitted to the system.

4.4.4 Extinction of Smoldering Fire in a Fish Meal Silo in Norway in 1992 Using Water

During the struggle with trying to fight the smoldering fires in the silo complex in Tomylovo, Russia (see Section 4.4.2) it was agreed that the use of water was not feasible. The arguments were that limited quantities of water would probably enhance the self-heating process rather than quench it, and use of extensive quantities would increase the load on the silo walls (hydrostatic pressure) and cause collapse of the entire structure.

In view of this it is interesting to consider the use of water for extinguishing a fire in a fish meal storage silo in Norway a few years later. In this case, described by Kroken (1997), the silo was cylindrical, of height 20 m and diameter 4.5 m. The bottom hopper was wedge-shaped and fitted with a horizontal screw conveyor for discharging of the meal. The sides of the hopper were at sixty degrees against the horizontal plane.

The entire sequence of events was as follows:

- The fresh fish meal from the production plant was filled into a mixing silo for being recycled and conditioned, before being transferred to a storage silo.

- In order to prevent self-heating in the meal during the conditioning period, recycling by discharging the meal from the silo bottom via the screw conveyor, and returning it to the silo at the top, was started.
- After some days it was discovered that the meal discharge operated by funnel flow rather than by mass flow, i.e. most of the material was not taking part in the recycling process.
- It was realized that digging out the material in the "dead" zones in the silo would be necessary, but due to lack of time, this work was not started until four weeks later. This made it possible for self-heating to develop in the stored fish meal.
- When trying to discharge the meal from the silo bottom after the four week break, bridging of the meal across the entire silo cross-section was discovered.
- Digging out was started from the silo top. When material had been removed down to the level of 2 m above the transition between the cylindrical silo wall and the hopper, the digging had to be stopped. The three main reasons were smoke development, that the meal was too hot to stand on and handle, and that the meal had become very hard very hard (a steel spear could only be forced 10 cm into the meal by hand).
- A nitrogen atmosphere was established in the silo above the meal surface.
- The fire brigade was called. However, they did not know how to deal with the matter, and left.
- The nitrogen atmosphere in the silo was kept for three weeks, but the smoldering fire inside the meal did not extinguish.
- It was then decided to try to fill water into the silo. But first information was collected about the strength of the bottom/hopper part of the silo. It was concluded that the bottom part could just about take the hydrostatic pressure that might result from soaking the meal with water.
- Water was filled into the silo. "Cracking" sounds were heard from the hopper section.
- After ten days the silo bottom was opened, and the water was drained out.

- Workers were admitted into the silo again to resume the digging-out work. The meal was then just like concrete and had to be removed by using pick-axes and hammers and chisels.
- After some digging-out regions with smoldering meal were once more encountered the digging had to be interrupted.
- The silo was once more sealed, and the remaining fish meal soaked with water.
- After ten more days the water was drained out, and the digging-out resumed. This time it was possible to empty the silo completely.
- Outwards buckling of the hopper walls was observed.

The total time, from the onset of the efforts to extinguish the fire to final completion, was about two months. This story illustrates that adequate procedures for extinction of fires in silos still remain to be developed.

4.5 Measures for Preventing Excessive Self-Heating in Silos

4.5.1 Temperature Measurement to Detect Self-Heating/ Self-Ignition

Temperature sensors positioned inside the bulk of material stored in silos by means of wires hanging from the silo roof have been used for detecting the onset of self-heating for many decades. However, in larger silos it is impossible in practice to cover the entire silo volume by temperature sensors, and self-heating and even smoldering fires may develop without being detected by the array of sensors available.

4.5.2 Early Detection of Tracer Gases from Self-Heating

More recently increasing focus has been on the use of very sensitive sensors for detecting the onset of generation of decomposition gases from self-ignition processes. Zockoll (1996) described the development of a new system for early detection of self-heating/self-ignition in deposits of organic powders, with special reference to milk powder in spray dryers, based on detection (infrared light) of low concentrations of CO in the 1–10 ppm range. Figure 4–13 gives some results from Zockoll's work,

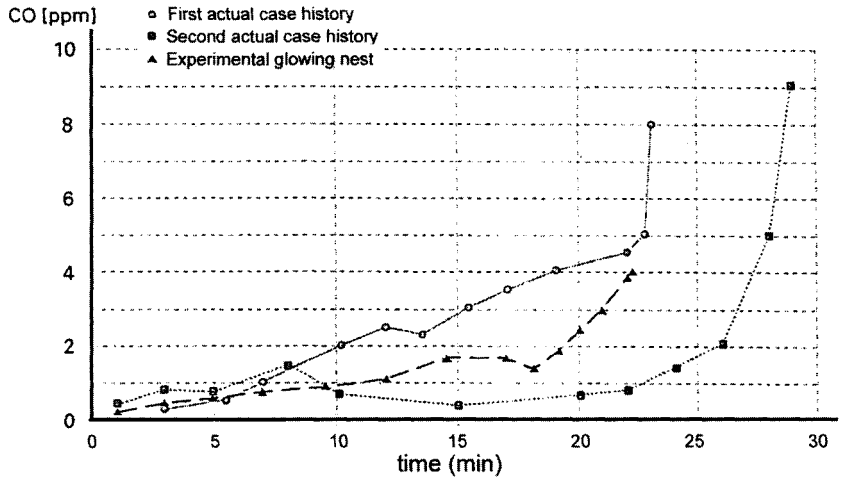


Figure 4–13 CO concentration development with time in two actual industrial cases of accidental self-ignition, and in one controlled self-ignition experiment. From Zockoll (1996).

showing how the measured CO concentrations increase with time. As soon as unacceptably high concentrations are detected, alarms are automatically activated, and measures to handle the problem are taken. In the case of early self-heating, re-circulation of the material in the silo can be an appropriate measure (see Section 4.5.3).

Kohl and Kelleter (1996) described a multi-sensor system for detection of combustible gases developed by slow smoldering fires in lignite. H_2 and CO were monitored continually, using electrochemical sensors. It was found that the ratio of the measured concentrations of the two gases could provide information about the development of the smoldering fire. When implementing systems for early detection of combustion gases, it is important that the supplier of the detection system and the user work closely together to ensure optimal system design for any given application.

4.5.3 Re-Circulation of Material in Silo by Discharging at Bottom and Re-Filling at Top

This method requires adequate silo design (mass flow as opposed to core flow) to ensure that the entire material becomes involved, not only a central

core. As illustrated by the case history described in Section 4.4.4, re-circulation can fail in smoothing out temperature differences in silos if only part of the stored material is participating in the re-circulation process.

4.5.4 Mathematical Modeling/Numerical Simulation of Self-Heating/Self-Ignition Processes

It is anticipated that in the future, the likelihood of onset of self-heating in any given industrial situation can be assessed by numerical simulation. Carson (1996) suggested such a method for assessing the tendency of combustible dusts to self-heat and start to burn spontaneously in a given practical situation. The method comprised various laboratory-scale tests combined with mathematical modeling. The test methods included Differential Thermal Analysis (DTA), Thermo Gravimetric Analysis (TGA), isothermal oven tests and adiabatic calorimetry. Krause and Hensel (1996) presented a numerical method by which non-steady temperature fields in dust deposits can be computed. This enables numerical analysis of a number of practical cases of self heating/self ignition that cannot be analyzed using the classical thermal explosion theory of Frank-Kamenetzki. Krause and Schmidt (2001) studied experimentally critical thermal conditions that may lead to initiation of smoldering processes, or to further development of such processes, once initiated. This type of evidence is important for further development of mathematical simulation models.

Chapter 5

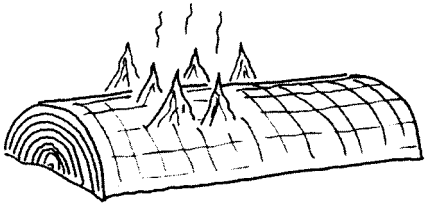
Dust Explosions

5.1 What is a Dust Explosion?

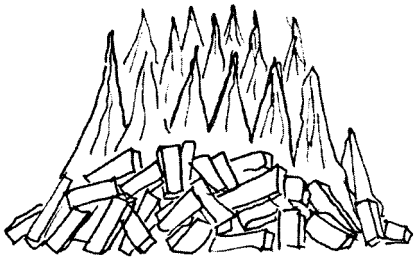
5.1.1 Introduction

The phenomenon of *dust explosions* is in fact quite simple and easy to envision in terms of daily life experience. Any solid material that can burn in air will do so with a violence and speed that increases with increasing degree of sub-division of the material. Figure 5–1(a) illustrates how a big piece of wood, once ignited, burns slowly, releasing its heat over a long period of time. When the wood is cut in small pieces, as illustrated in Figure 5–1(b), the combustion rate increases because the total contact surface area between wood and air has been increased. Also, the ignition of the wood has become easier. If the sub-division is continued right down to the level of small particles of sizes of the order of 0.1 mm or less, and the particles are suspended in a sufficiently large volume of air to give each particle enough space for its unrestricted burning, the combustion rate will be very fast, and the energy required for ignition very small. This is illustrated in Figure 5–1(c). Such a burning dust cloud is a dust explosion.

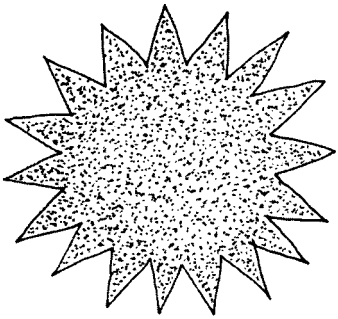
In general, the dust cloud will be easier to ignite and burn more violently the smaller the dust particles are, down to some limiting particle size that depends on the type of dust material. If such an explosive combustion of a dust cloud takes place inside process equipment or work rooms, the pressure in the fully or partly enclosed explosion space may rise rapidly and the process equipment or building may burst, and life, limb and property can be lost.



(a) SLOW COMBUSTION



(b) FAST COMBUSTION



(c) EXPLOSION

Figure 5–1 Illustration of the increase of the combustion rate of a given mass of combustible solid increases with increasing sub-division. From Eckhoff (2003).

5.1.2 Materials that Can Give Dust Explosions

In the same way as gas/vapor and spray/mist explosions do, dust explosions also generally arise from rapid release of heat due to the chemical reaction:

$$\text{fuel} + \text{oxygen} \rightarrow \text{oxides} + \text{heat}$$

In some special cases metal dusts can also react exothermally with nitrogen or carbon dioxide, but most often oxidation by oxygen is the heat-generating process in a dust explosion. This means that only materials that are not already stable oxides can give rise to dust explosions. This excludes substances such as silicates, sulfates, nitrates, carbonates, and phosphates, and therefore dust clouds of Portland cement, sand, limestone, etc. cannot give dust explosions.

The materials that can give dust explosions include:

- natural organic materials (grain, wood, linen, sugar, etc.)
- synthetic organic materials (plastics, organic pigments, pesticides, pharmaceuticals, etc.)
- coal and peat
- metals (aluminum, magnesium, titanium, zinc, iron, etc.)

The heat of combustion of the material is an important parameter because it determines the amount of heat that can be liberated in the explosion. However, when comparing the various materials in terms of their potential hazard, it is useful to relate the heat of combustion to the same amount of oxygen consumed. This is because the gas in a given volume of dust cloud contains a given amount of oxygen, which determines how much heat can be released in an explosion per unit volume of dust cloud. In Table 5–1, the heats of combustion of various substances, per mole of oxygen consumed, are given. Ca and Mg top the list, with Al closely behind. Si is also fairly high up on the list, with a heat of combustion per mole of oxygen about twice the value of typical natural and synthetic organic substances and coals.

Table 5–1 is in accordance with the experience that the temperatures of flames of dusts of metals like Al and Si are very high compared with those of flames of organic dusts and coals.

Table 5–1 Heats of Combustion (Oxidation) of Various Substances per Mole O_2 Consumed

Substance	Oxidation product(s)	KJ/mole O_2
Ca	CaO	1270
Mg	MgO	1240
Al	Al_2O_3	1100
Si	SiO_2	830
Cr	Cr_2O_3	750
Zn	ZnO	700
Fe	Fe_2O_3	530
Cu	CuO	300
Sucrose	CO_2 and H_2O	470
Starch	CO_2 and H_2O	470
Polyethylene	CO_2 and H_2O	390
Carbon	CO_2	400
Coal	CO_2 and H_2O	400
Sulphur	SO_2	300

5.1.3 Factors Influencing the Ignitability and Explosibility of Dust Clouds

A fairly comprehensive list of factors influencing the ignitability and explosibility of dust clouds may look as follows:

- chemical composition of the dust, including its moisture content
- chemical composition, and initial pressure and temperature of the gas phase
- distributions of particle sizes and shapes in the dust, determining the specific surface area of the dust in the fully dispersed state
- degree of dispersion, or agglomeration, of dust particles, determining the effective specific surface area available to the combustion process in the dust cloud in the actual industrial situation

- distribution of dust concentration in the actual cloud
- distribution of initial turbulence in the actual cloud
- possibility of generation of explosion-induced turbulence in the still unburned part of the cloud (location of ignition source important parameter)
- possibility of flame front distortion by other mechanisms than turbulence
- possibility of significant radiative heat transfer (highly dependent on flame temperature, which in turn depends on particle chemistry)

Factors one, two, three, and nine can be regarded as basic parameters of the explosive dust cloud. Factors four to eight are, however, influenced by the actual industrial dust cloud generation process and explosion development. These in turn depend on the nature of the industrial process (flow rates, etc.) and the geometry of the system in which the dust cloud burns. The location of the ignition point is another parameter that can play an important role in deciding the course of the explosion.

In view of the wide spectrum of dust cloud concentrations, degrees of dust dispersion and turbulence, and of locations of potential ignition sources in industry, a correspondingly wide spectrum of possible dust cloud ignition sensitivities and combustion rates must be expected.

This complex reality of the process industry is also shared by laboratory experimentation and represents a constant challenge in the design of adequate experiments and interpretation of experimental results.

5.2 Combustion of Dust Clouds in Air

5.2.1 Explosive Concentration Ranges of Dust Clouds in Air

The explosive combustion of dust clouds, as illustrated in Figure 5–1(c) cannot take place unless the dust concentration, i.e. the mass of dust per unit volume of dust clouds, is within certain limits. This is analogous with combustion of homogeneous mixtures of gaseous fuels and air, for which the upper and lower flammability limits are well established. Figure 5–2 shows the explosive range for a typical natural organic material, such as maize starch, in air at normal temperature and atmospheric pressure. This

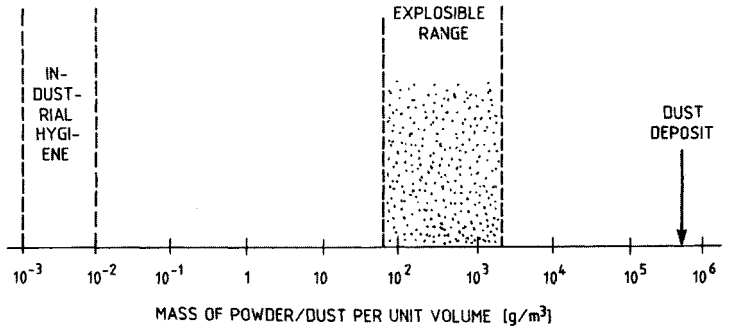


Figure 5–2 Range of explosive dust concentrations in air at normal temperature and atmospheric pressure for a typical natural organic dust compared with typical range of maximum permissible dust concentrations in the context of industrial hygiene, and a typical density of deposits of natural organic dusts.

range is quite narrow, extending over less than two orders of magnitude, from the order of 100 g/m^3 on the lean side to 2–3 kg/m^3 on the rich one. The explosive limits differ somewhat for the various dust materials. For example, zinc powder has a minimum explosive concentration in air of about 500 g/m^3.

Explosive dust clouds have a high optical density, even at the lower explosive limit. This is illustrated by the fact that the range of maximum permissible dust concentrations that are specified in the context of industrial hygiene in working atmospheres, are three to four orders of magnitude lower than typical minimum explosive dust concentrations. This means that the unpleasant dust concentration levels that can sometimes occur in the general working atmosphere of a factory, and which calls upon the attention of industrial hygiene authorities, are far below the concentration levels that can propagate dust flames. Therefore, the minimum explosive concentration corresponds to dust clouds of high optical densities, which are unlikely to occur regularly in work rooms of factories. Figure 5–3 illustrates the high optical densities of explosive dust clouds, based on a rule of thumb by Intelmann, and quoted by Zehr (1965): If a glowing 25 W light bulb is observed through 2 m of coal dust cloud, the bulb light cannot be seen through dust concentrations exceeding 40 g/m^3.

It then follows that the dust clouds in which dust explosions are primarily initiated are practically always found inside process equipment, such as

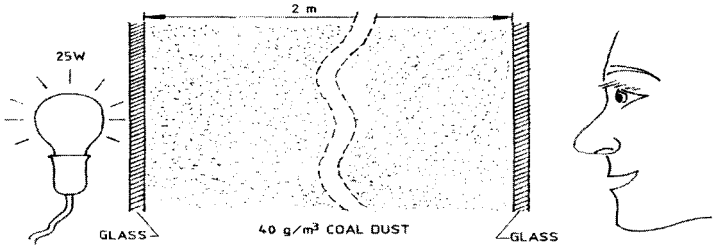


Figure 5–3 A cloud of 40 g/m^3 of coal dust in air is so dense that a glowing 25 W light bulb can hardly be seen through a dust cloud of 2 m thickness. From Eckhoff (2003).

mills, mixers, screens, dryers, cyclones, filters, bucket elevators, hoppers, silos, aspiration ducts, and pipes for pneumatic transport of powders. Such explosions, initiated by some ignition source are called *primary* explosions.

This reveals an important difference between primary dust and gas explosions. In the case of dusts, the primary explosive clouds practically always occur inside process equipment, i.e. in areas where electrical apparatus is scarce or absent. In the case of gases, however, the main hazard is primary gas explosions outside process equipment where gas from accidental leaks gets mixed with air and explosive atmospheres are generated, and where a wide range of potential ignition sources may be found, including various electrical apparatuses.

5.2.2 Flame Propagation Processes in Dust Clouds

5.2.2.1 Basic Differences between Premixed Gas Flames and Flames in Dust Clouds

Leuschke (1965) pointed out some characteristic differences between a laminar premixed gas flame and a laminar dust flame. One important difference is that the reaction zone in the dust cloud is considerably thicker than in the gas, irrespective of the type of dust, and of the order of at least 10–100 mm. When discussing this feature of dust flames, it is useful to distinguish between two flame types. The first, the Nusselt type, is controlled by diffusion of oxygen to the surface of individual particles, where a heterogeneous chemical reaction takes place. In the second type, the volatile flame type, the rate of gasification, pyrolysis, or devolatilization

is the controlling process, and the chemical reaction takes place mainly in a homogeneous gas phase. In Nusselt type flames, the greater thickness of the combustion zone, as compared with that of premixed gas flames, results from the slower rate of molecular diffusion, compared to diffusion in homogeneous premixed gases. With the volatile flame type, the greater flame thickness is due to a pre-heating zone, where combustible gases and vapors (volatiles) are driven out of the particles ahead of the flame. Premixed volatiles and air burn just as readily as any explosive mixture of gaseous hydrocarbon and air. The combustion of the remaining solid char particles occurs subsequently at a slower rate in the tail of the flame, and therefore the volatile flame in clouds of coals and organic dusts is also, in fact coupled to a Nusselt type flame.

In the case of metal dusts, low-melting-point materials may oxidize in the vapor phase, but due to the oxide film round each particle this does not result in a homogeneous metal vapor/air flame. Because of the large heat of combustion per mole oxygen of for example aluminum and magnesium dust, compared with organic dusts, the temperature of the burning particles is very high and thermal radiation plays a significant role in the transfer of heat in the combustion wave. Radiative heat transfer is also supposed to play a role in coal dust flames. However, because the thermal radiation is proportional to the fourth power of the temperature, the role of thermal radiation in coal dust flames is probably less important than in aluminum and magnesium dust flames. Leuschke (1965) conducted an illustrative series of experiments demonstrating the pressure of radiative heat transfer in metal dust flames, using the experimental set-up illustrated in Figure 5–4.

Two transient dust clouds were generated simultaneously on the two sides of a double-glass window, one being ignited immediately by a gas flame. It was then observed whether the radiation from this burning cloud was able to ignite the second cloud. Only flames of Zr, Ti, Al, and Mg were able to produce sufficient radiation to ignite the second cloud. The specific role of thermal radiation in the propagation of dust flames and build-up of the explosion pressure varies with the dust type, and more research is needed to quantify the radiative contribution under various circumstances.

Another difference between flame propagation in a premixed gas and dust cloud has been elucidated by Goral, Klemens, and Wolanski (1988). They studied upwards propagation of flames in a lean methane/air mixture to which had been added inert particles (sand). It was found that the upwards

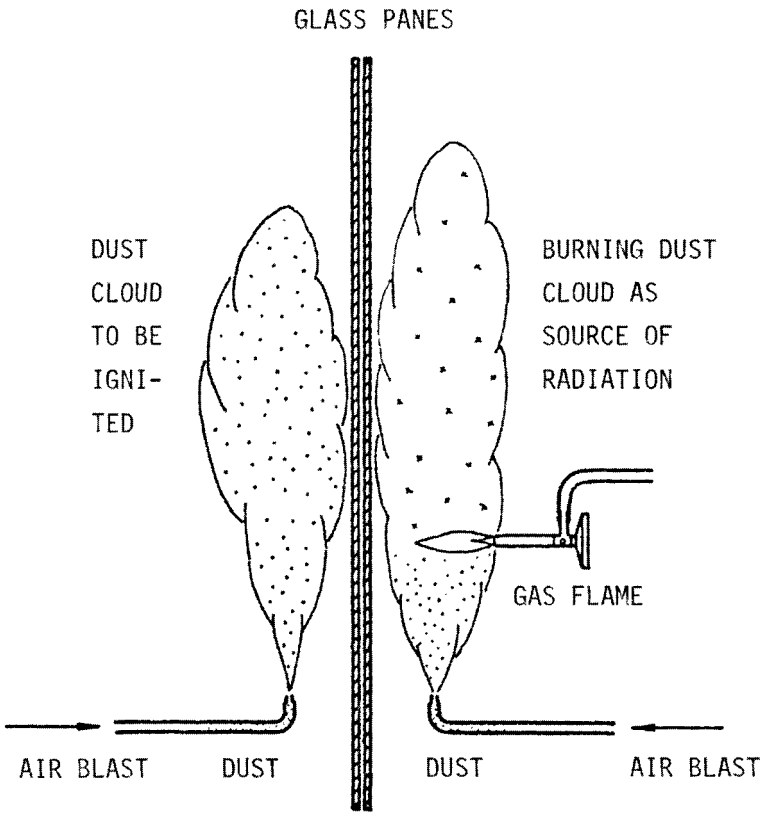


Figure 5–4 Experiment demonstrating the ignition of a cloud of metal dust in air by radiation from a burning cloud of the same dust through a double glass window. From Leuschke (1965).

flame velocity increased with increasing sand grain size. The effect was mainly attributed to the enhanced combustion due to the micro turbulence generated in the wake of the falling particles, although thermal radiation effects were also assumed to play a role.

5.2.2.2 Influence of Dust Chemistry

There are two aspects to consider, namely the thermodynamics of the explosion, and the kinetics. The thermodynamics is concerned with the amount of

heat that is liberated during combustion, the kinetics with the rate at which the heat is liberated.

The maximum rate with which the explosion pressure rises in closed-bomb experiments is a frequently used relative measure of the violence to be expected from explosions of a given dust.

Figure 5–5 shows how the maximum rates of pressure rise of starch (potato and maize starch) are systematically higher than for protein (two fish powders with fat removed) for the same specific surface area. The nitrogen compounds in the protein are probably in some way slowing down the combustion process.

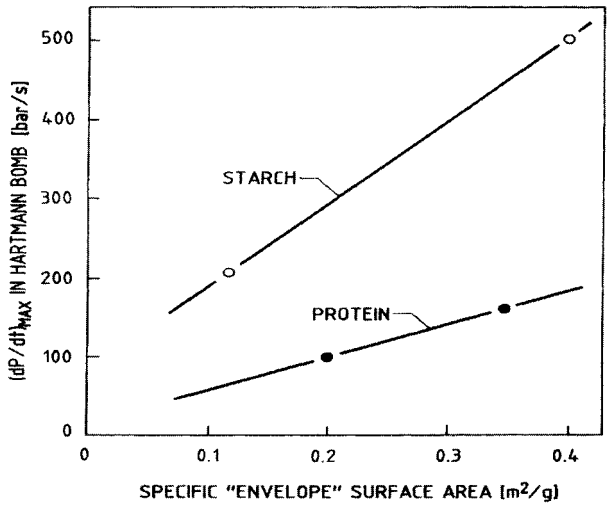


Figure 5–5 Influence of chemistry (starch or protein) and specific surface area of natural organic materials on maximum rate of pressure rise in closed 1.2 liter Hartmann bomb. From Eckhoff (1977/1978).

Another example of the influence of dust chemistry on the explosion kinetics is shown in Figure 5–6. The heats of combustion of polyvinylchloride (PVC) and polyethylene are not very different. Closed-bomb experiments also give about the same maximum pressure for very small particle sizes. However, the chlorine in the PVC causes a quite dramatic drop in the rate of heat release as the median particle size increases beyond about 20 μm. Due to the very slow combustion, P_{max} for PVC also drops much faster as the particle size increases than for polyethylene. The retarding influence of chlorine on the combustion process most probably is of the

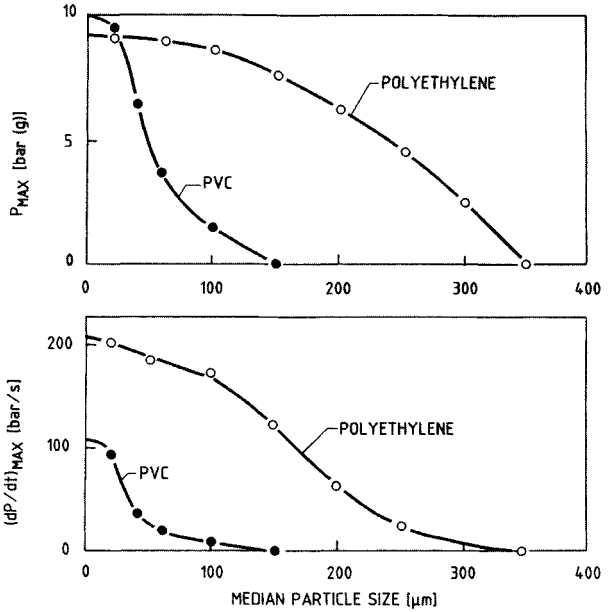


Figure 5–6 Influence of chlorine in molecule of dust material on maximum explosion pressure and maximum rate of pressure rise in 1 m^3 standard ISO vessel, for various particle sizes. From Bartknecht (1978).

same nature as that of the halogens in the halons, which were extensively used for explosion and fire suppression before the negative influence of such materials on the global environment was fully realized.

5.2.2.3 Influence of Moisture

Moisture in the dust reduces both ignition sensitivity and explosion violence of dust clouds markedly. If safety measures against electric spark ignition are based on Minimum Ignition Energy (MIE) data for a given dust moisture content, it is essential that this moisture content is not subsided in practice. The influence of dust moisture on the minimum ignition temperature of dust clouds is less marked than on MIE. The specific role of moisture in reducing both ignition sensitivity and explosion violence of clouds of organic dusts is complex. First, evaporation and heating of water represents an inert heat sink. Secondly, the water vapor mixes with the pyrolysis gases in the preheating zone of the combustion wave, and

makes the gas mixture less reactive. A third factor is that moisture increases the inter-particle cohesion of the dust and prevents dispersion into primary particles.

5.2.2.4 Influence of Particle Size

Figure 5–5, in addition to illustrating the influence of dust chemistry on the dust cloud combustion rate, also shows a clear influence of particle size/specific surface area for both materials. This is a general trend for most dusts. However, this trend does not continue indefinitely as the particles get smaller. In the case of coal and organic materials pyrolysis/devolatilization always precedes combustion, which primarily occurs in the homogeneous gas phase. The limiting particle size, below which the combustion rate of the dust cloud does not increase any more, depends on the ratios between the time constants of the three consecutive processes devolatilization, gas phase mixing and gas phase combustion. Particle size primarily influences the devolatilization rate. Therefore, if the gas phase combustion is the slowest of the three steps, increasing the devolatilization rate by decreasing the particle size does not increase the overall combustion rate. For coals it was found that the limiting particle diameter is of the order of 50 μm. However, for materials yielding gaseous pyrolysis products that are more reactive than volatiles from coal, e.g. due to unsaturated gaseous compounds, one would expect the limiting particle size to be smaller than for coal. For natural organic compounds such as starch and protein, the limiting particle diameter is probably not much smaller than about 10 μm, whereas for reactive dusts such as some organic dyes, it may well be even smaller.

For metals, in particular those at the top of Table 5–1, the limiting particle size, below which there is no longer any increase in ignition sensitivity and explosion violence, is considerably smaller than for most organic materials. This is because these metals do not devolatilize or pyrolyze, but melt, evaporate, and burn as discrete entities. In such cases the aggravating influence of particle size may continue even below 1 μm.

The pronounced decrease of MIE with decreasing particle size is discussed in Section 5.3.6.

5.2.3 Close-to-Laminar Flame Propagation in Dust Clouds

Close-to-laminar 20 mm diameter burner flames of clouds in air of lycopodium and polyvinyl alcohol were studied by Kaesche-Krischer and Zehr (1958) and Kaesche-Krischer (1959). The burning velocity, defined as the ratio of air flow and flame cone area, was determined photographically from the height of the flame cone. Some results are given in Figure 5–7.

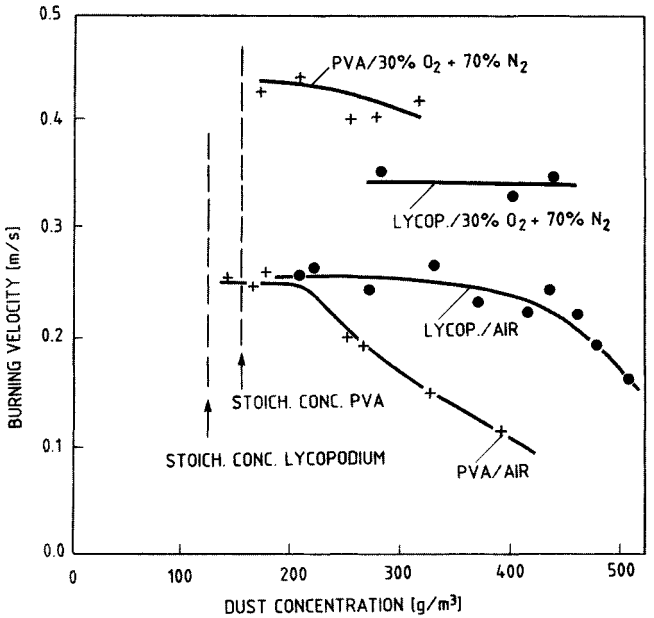


Figure 5–7 Burning velocities of flames of lycopodium and polyvinyl alcohol dust (<60 µm particle diameter) flames as functions of dust concentration. The dotted stoich. conc. lines refer to dust in air only. From Kaesche-Krischer and Zehr (1958) and Kaesche-Krischer (1959).

Lycopodium/air flames of dust concentrations lower than 180 g/m^3 and higher than 500 g/m^3 were difficult to stabilize (stoichiometric concentration $\approx$ 125 g/m^3). The appearance of a stabilized lycopodium/air flame was very similar to that of a rich hydrocarbon/air flame, i.e. a blue flame front followed by a more or less luminous soot tail. Approximate thermocouple measurements of flame temperatures gave about 1800 K for a 180 g/m^3 flame, and 1100 K for a 500 g/m^3 flame. Figure 5–7 shows the

measured burning velocities as a function of the dust concentration. In the range 180–300 g/m^3 the burning velocity of lycopodium flames has a maximum value of about 0.25 m/s. The corresponding concentration range for the PVA dust was 140–220 g/m^3. Figure 5–7 also shows that an increase of the oxygen content in the gas from 21 vol.% in air to 30 vol.%, gave a significant increase of the measured burning velocities for both dusts, in accordance with expectations. The photographs provided by Kaesche-Krischer and Zehr (1958) indicate typical thicknesses of lycopodium flames of a few mm.

Kaesche-Krischer implied that the differences in the concentration ranges giving the highest burning velocities for the two dusts were due to a higher volatile content in the PVA than in lycopodium, assuming that the flame essentially propagates through a homogenous mixture of volatiles and air.

In a comprehensive survey of a number of investigations on the propagation of laminar pulverized coal dust/air flames, Smoot and Horton (1977) discussed factors influencing experimentally determined burning velocities, flame temperatures and flame thicknesses. Most experiments have been performed on stabilized dust flames in burners of various kinds. Due to heat losses by radiation from the hot dust particles, and conduction, typical stabilized burner flames will have temperatures that are lower than the adiabatic flame temperature. In principle heat losses can be avoided by using burners of very large diameters, or equipped with walls having temperature and emissivity profiles matching those of the flame. Smoot and Horton found large differences in burning velocities observed by various investigators which could not be explained in terms of variations in dust properties or dust concentration. They considered incomplete dispersion of fine cohesive dusts, as illustrated in Figure 5–8, as the main source of error.

The experimental results of Smoot and Horton in Figure 5–9 illustrate how improved dispersion of a fine coal dust increased the burning velocity by at least fifty percent.

Non-homogeneity is an inherent feature of any real dust cloud, and this also influences the pattern of flame propagation through such clouds. This not least complicates the experimental determination of the minimum explosive dust concentration. Informative experiments has been conducted in fairly large industrial scale equipment, the work of Palmer and Tonkin (1971) being a good example. Their apparatus is illustrated in Figure 5–10.

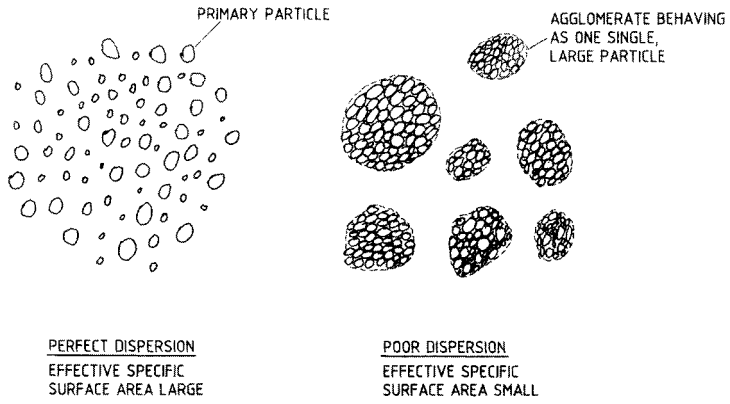


Figure 5–8 Illustration of perfectly dispersed dust cloud and cloud consisting of agglomerates of much larger effective particle sizes than those of the primary particles. From Eckhoff (2003).

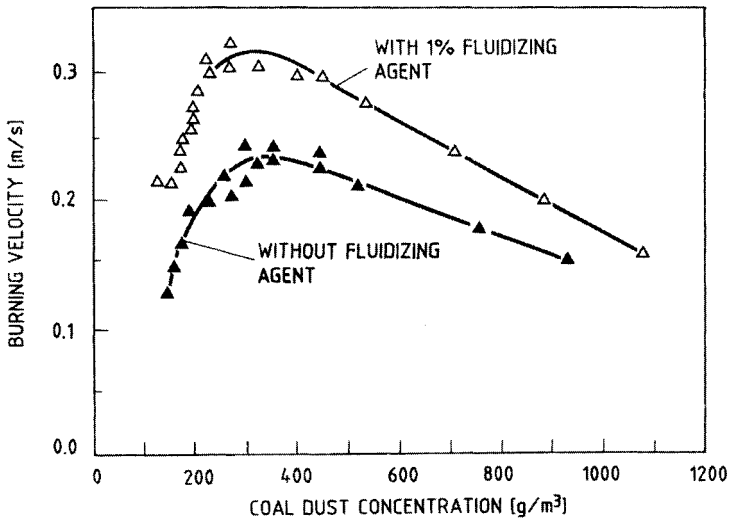


Figure 5–9 Effect of very fine SiO_2 fluidizing agent (Acrosil) on the burning velocity of an air suspension of 10 µm, 28% volatile content Sewell coal dust. From Smoot and Horton (1977).

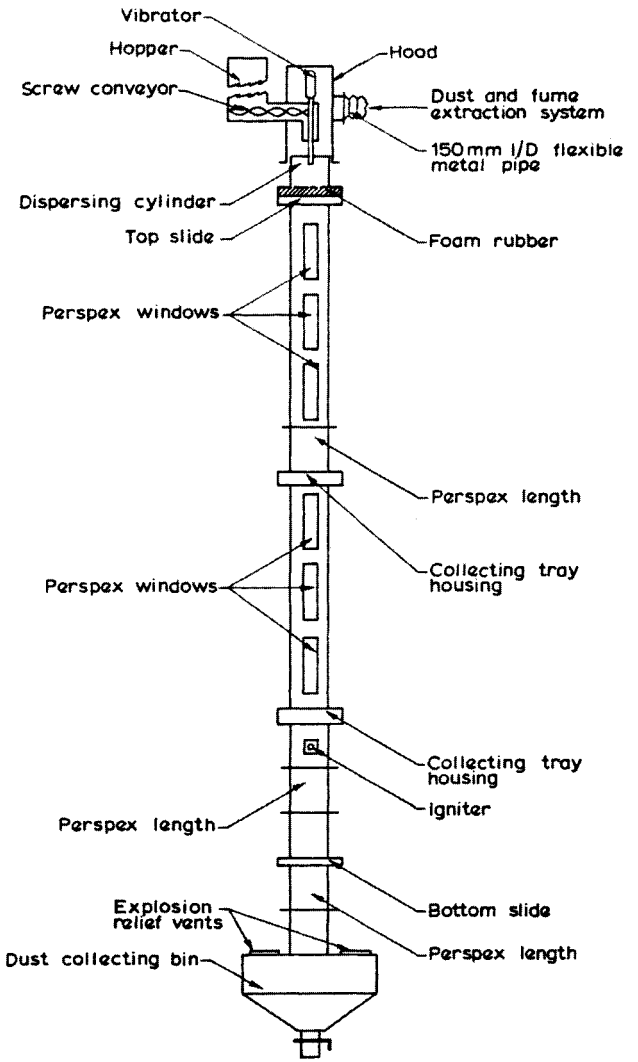


Figure 5–10 Vertical large-scale explosion tube facility for flame propagation studies. Internal tube diameter 0.25 m. Total tube length 5.2 m. Ignition point 1.5 m above bottom. From Palmer and Tonkin (1971).

The dust was introduced at the top of the tube by a screw feeder and dropped into a vibrating 20 cm diameter and 15 cm high dispersing cylinder hanging immediately underneath the screw exit. After having passed through the perforated bottom of the cylinder, the dispersed dust settled

freely under gravity through the entire length of the tube until finally being collected in a bin at the bottom end. Dust concentration and flame propagation could not be measured in the same test, but had to be determined in separate tests at nominally identical dust cloud generation conditions, i.e. rotating speed of the feeding screw conveyor and vibration mode of the dust disperser. The dust concentration was measured gravimetrically. A manually operated sliding tray was inserted into the tube like a gate valve about 3.5 m below the tube top. By simultaneously closing the tube at the top by a conventional sliding gate valve, the volume of dust cloud between the top valve and the tray was trapped. By dividing the amount of dust finally settled out on the tray by the volume between the tray and the top valve, the average dust concentration in this section of the tube could be obtained. Immediately before performing an explosion test the dust feed was stopped and the bottom end of the tube closed by a gate valve located just below the ignition zone. The ignition source was a propane flame generated by injecting a small pocket of a propane/air mixture into the bottom region of the explosion tube and igniting by means of an electric spark located at the tube axis.

By means of this apparatus the average length of upwards vertical flame propagation away from the ignition source could be determined as a function of the average dust concentration in the measurement section.

5.2.4 Maximum Pressures Generated from Constant-Volume Adiabatic Combustion of Dust Clouds in Air

In Table 5–1 calcium, magnesium, and aluminum top the list with 1,100–1,300 kJ/mole oxygen consumed. The lowest values are 300 kJ/mole oxygen for copper and sulphur. It would be expected that this difference will to some extent be reflected in the maximum pressure of explosions, when performed adiabatically at constant volume. Zehr (1957) made some calculations of the maximum pressures to be expected under such conditions. In Figure 5–11, his results have been plotted against data from experiments in either 1 m^3 or 20 liter closed bombs.

For aluminum and magnesium Zehr only indicated that the theoretical values would be larger than 10 and 13.5 bar(g) respectively. This is because at high temperatures the two metals will also react exothermally with the nitrogen in the air. Theoretical maximum pressures based on the assumption that reaction with oxygen is the only contributor to

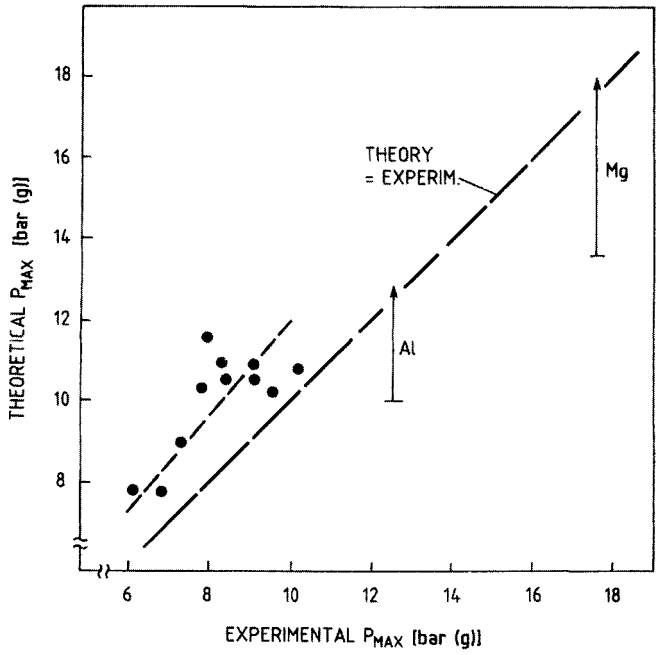


Figure 5–11 Correlation between P_{max} from experiments in 1 m^3 or 20 liter closed vessels, and theoretical adiabatic constant-volume P_{max} values calculated by Zehr (1957).

heat production therefore gives lower values than the experimental ones. In view of the uncertainties involved in the calculations, Figure 5–11 suggests a fair correlation between theoretical and experimental data, with the theoretical results being somewhat higher than the experimental ones for the organic dusts. This would be expected because of the idealized assumptions of stoichiometry and complete oxidation of all fuel, on which the calculations were based.

Figure 5–12 shows experimental data confirming the expectation that the maximum adiabatic constant volume explosion pressure (absolute) for dust clouds is proportional with the initial pressure (absolute) of the unburned cloud, as for explosive mixtures of combustible gases.

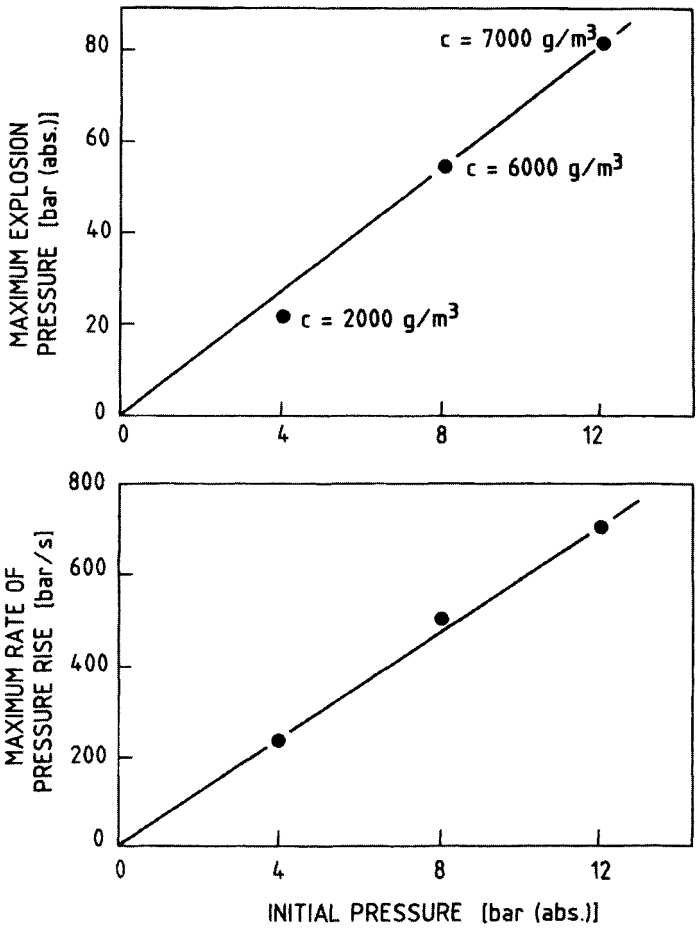


Figure 5–12 Influence of initial pressure on maximum pressure and maximum rate of pressure rise in explosions of clouds of sub-bituminous coal dust in air in a 15 liter closed bomb. Median particle size by mass 100 µm. From Pedersen and Wilkins (1988).

5.2.5 Turbulent Flame Propagation in Dust Clouds

5.2.5.1 Basic Features of Closed-Bomb Studies of Turbulent Dust Flames

Much of the published experimental study of turbulent dust explosions has been conducted in closed vessels of the type illustrated in Figure 5–13.

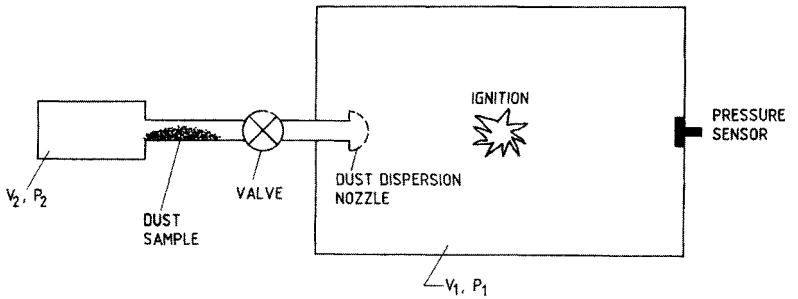


Figure 5–13 Schematic illustration of the type of apparatus commonly used in closed-vessel turbulent dust explosion experiments. From Eckhoff (2003).

The explosion vessel of volume V_1 and initial pressure P_1 is equipped with a dust dispersion system, a pressure sensor and an ignition source. In most equipment the dust dispersion system consists of a compressed-air reservoir of volume $V_2 << V_1$, at an initial pressure $P_2 >> P_1$. In some apparatuses the dust is initially placed on the high-pressure side of the dispersion air valve, as indicated in Figure 5–13, whereas in other apparatus it is placed downstream of the valve. Normally, the mass of dispersion air is not negligible compared with the initial mass of air in the main vessel. This causes a significant rise of the pressure in the main vessel once the dispersion air has been discharged into the main vessel. In some investigations this is compensated for by partial evacuation of the main vessel prior to dispersion so that the final pressure after dispersion completion, just prior to ignition, is atmospheric. This is important if absolute data are required, because the maximum explosion pressure for a given dust at a given concentration is approximately proportional to the initial absolute air pressure. Both the absolute sizes of V_1 and V_2 and the ratio between them vary substantially from apparatus to apparatus. The smallest V_1 used are of the order of 1 liter, whereas the largest that has been traced is 250 m^3. The design of the dust dispersion system varies considerably from apparatus to apparatus. A number of different nozzle types have been developed with the aim to break up agglomerates and ensure homogeneous distribution of the dust in the main vessel. The ignition source has also been a factor of considerable variation. In some of the earlier investigations, continuous sources like electric arcs or trains of electric sparks, and glowing resistance wire coils were used, but more recently it has become common to use short-duration sources initiated at a given time interval after opening of the dust dispersion valve. These sources

vary from electric sparks, via exploding wires to various forms of electrically triggered chemical igniters.

An important inherent feature of all apparatus of the type illustrated in Figure 5–13 is that the dispersion of the dust inevitably induces turbulence in the main vessel. The level of turbulence will be at maximum during the main phase of dust dispersion. After the flow of dispersion air into the main vessel has terminated, the turbulence decays with a time contraint that increases with increasing V_1. Therefore both the strength of the dispersion air blast and the delay between opening of the dust dispersion value and ignition have a strong influence on the state of turbulence in the dust cloud at the moment of ignition, and consequently also on the violence of the explosion. The situation is illustrated in Figure 5–14.

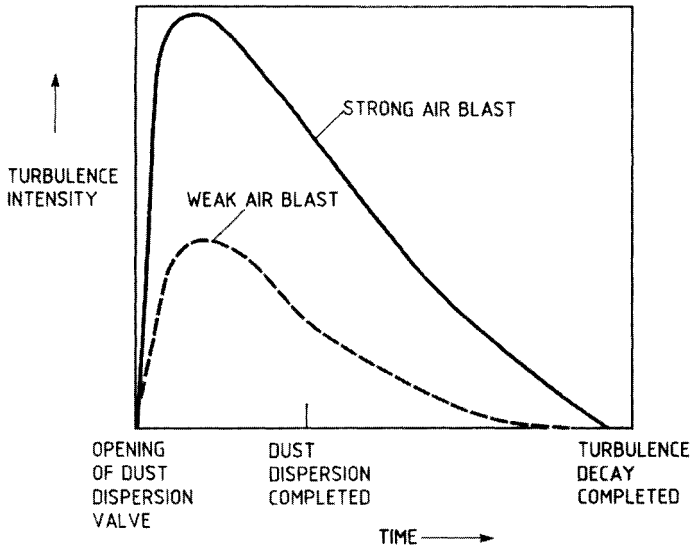


Figure 5–14 Illustration of generation and decay of turbulence during and after dispersion of dust in an apparatus of the type illustrated in Figure 5–13. (Note: A common way of quantifying turbulence intensity is by the Root Mean Square (RMS) of the turbulent velocity fluctuations as measured by e.g. Laser Doppler Anemometry.)

5.2.5.2 Closed-Bomb Experiments

The primary result from a closed bomb dust explosion experiment is a trace of the explosion pressure as a function of time, as illustrated in

Figure 5–15. Two parameters are normally of prime interest, viz. the maximum explosion pressure, i.e. the peak value of the recorded pressure-versus-time trace, and the maximum rate of pressure rise, i.e. the steepest gradient of the rising part of the pressure-versus-time trace.

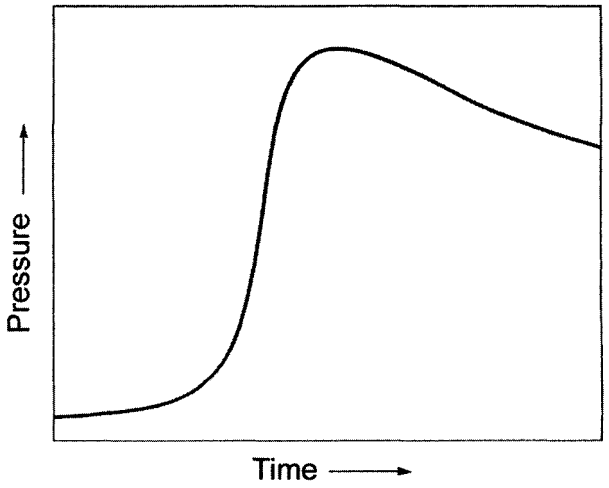


Figure 5–15 Illustration of typical trace of explosion pressure versus time during a dust explosion in a closed vessel.

Figure 5–16 gives an example of a set of results from a complete series of closed-bomb tests of one given dust.

Tests are carried out over a wide range of nominal dust concentrations (mass of dust dispersed divided by vessel volume). Furthermore, due to inevitable scatter, several tests must be performed at each nominal dust concentration. Normally the dust concentrations that give the highest values of the maximum pressure and maximum rate of pressure rise are not the same. The test results are the highest mean values obtained for each of the two parameters.

The data in Figure 5–17 illustrate the influence of the ignition delay on the explosion development in a cloud of lycopodium in air in a 1.2 liter Hartmann bomb.

As can be seen, there is little difference between the maximum explosion pressure obtained with a delay of 40 ms and the one obtained with a delay of 200 ms, whereas the maximum rate of pressure rise is drastically

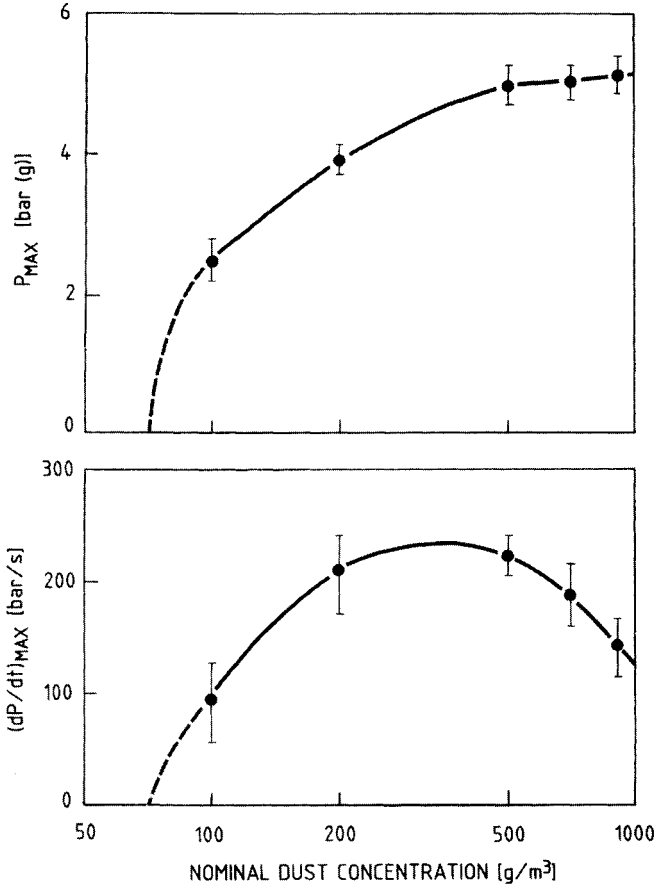


Figure 5–16 Influence of nominal dust concentration on maximum explosion pressure and maximum rate of pressure rise in dust explosions in air in the 1.2 liter Hartmann bomb. Dust: maize starch containing 11% moisture. Bars indicate ± 1 standard deviation. From Eckhoff et al. (1985).

reduced, from 430 bar/s to 50 bar/s, i.e. by a factor of almost ten. There is little doubt that this is due to the reduced initial turbulence in the dust cloud at the large ignition delays (see Figure 5–14). With increasing ignition delay beyond 200 ms, the maximum explosion pressure is also reduced, most probably because the dust has started to settle out of suspension before the ignition source is activated.

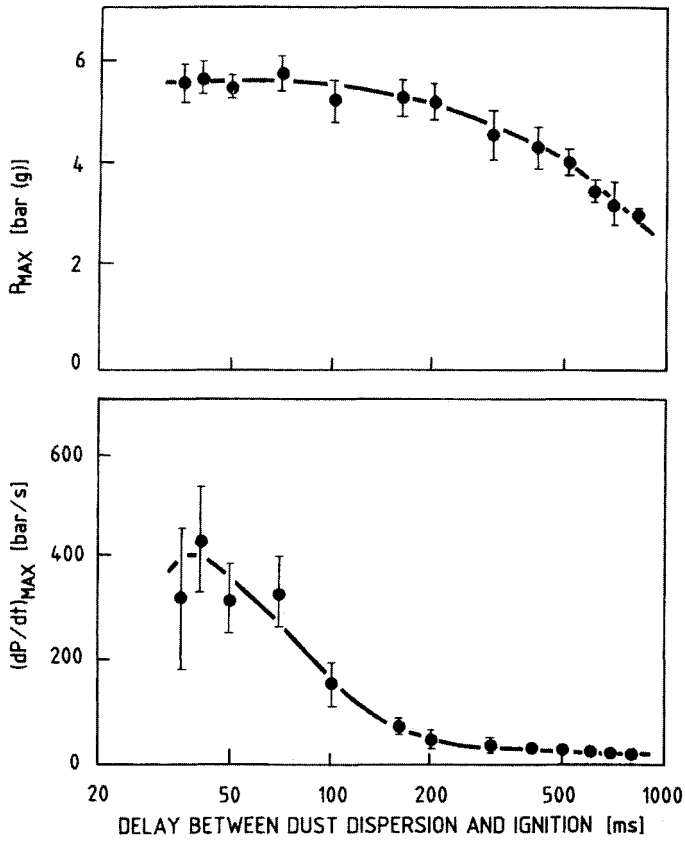


Figure 5–17 Influence of initial turbulence on explosion rate of a dust cloud. Experiments with 420 g/m^3 of lycopodium in air in the 1.2 liter Hartmann bomb. Five experiments per delay. Bars indicate ± 1 standard deviation. From Eckhoff (1977).

As would be expected, the same kind of influence of ignition delay as shown in Figure 5–17 is found in all experiments of the type illustrated in Figure 5–13. One of the first researchers to observe this effect was Bartknecht (1971) when performing dust explosion experiments in a 1 m^3 explosion vessel of the type illustrated in Figure 5–18. As the ignition delay was increased from the lowest value of about 0.3 s to about 1 s, Bartknecht found a marked decrease of $(dP/dt)_{max}$, whereas P_{max} was much less

dependent of the ignition delay. If the ignition delay was increased further, however, there was a marked decrease even for P_{max}.

For several decades the 1.2 liter Hartmann bomb, originally developed by I. Hartmann at the U.S. Bureau of Mines in the 1940s, was used as a standard apparatus in many countries for assessing maximum explosion pressures and maximum rates of pressure rise of various dusts. However, because of the small bomb size, the P_{max} values obtained are normally significantly lower than those attained in larger volumes for the same dusts. The current international (ISO) and European Union standard apparatus for measuring $(dP/dt)_{max}$ and P_{max} for various dusts is essentially the same as the 1 m^3 apparatus developed by Bartknecht (1971), and illustrated in Figure 5–18.

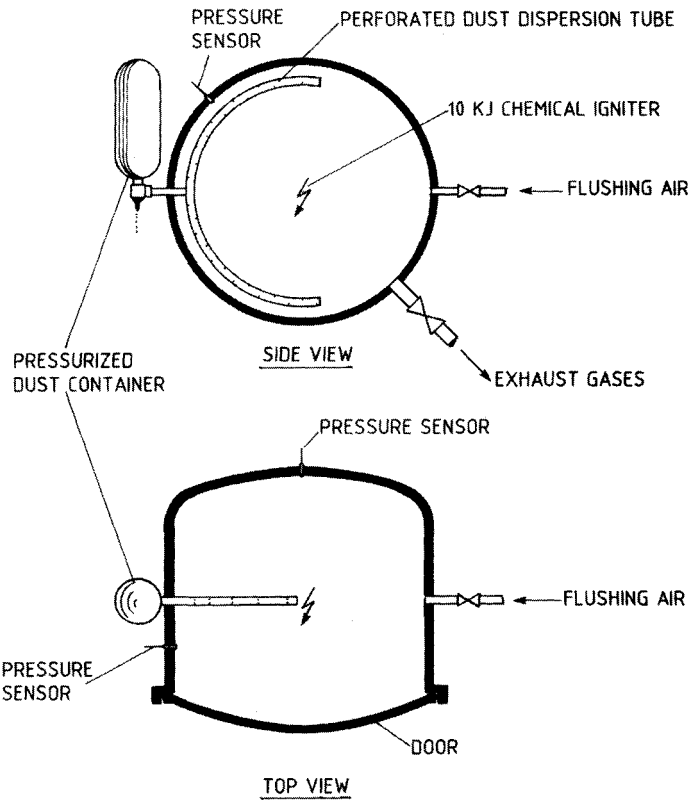


Figure 5–18 1 m^3 closed vessel specified by the International Standardization Organization (1985) for determination of maximum explosion pressures and maximum rates of pressure rise of dust clouds in air.

Based on the comprehensive studies by Bartknecht and Siwek, smaller vessels, most often of 20 liter volumeare now used for determining so-called K_{St} values according to the 1 m^3 ISO standard. K_{St} is defined as the product of the maximum rate of pressure rise determined in a given closed vessel and the cube root of the vessel volume. However, this "cube-root" law is only valid for idealized theoretical conditions. This implies that in practice, in order to satisfy the "cube-root" law, the dust dispersion system, the ignition source strength and the ignition delay in smaller vessels must be tuned in such a way that the products of the measured maximum rates of pressure rise (bar/s) and the cube roots of the vessel volumes (m) equal the K_{St} values that would have been measured for the same dusts in the 1 m^3 ISO standard test.

Figure 5–19 shows the 20 liter closed vessel designed by U.S. Bureau of Mines. This design gives easy access to the interior of the vessel for cleaning etc. The original dust dispersion system shown in the figure is not compatible with the ISO standard. However, as shown by Skjold (2003), compatibility can be obtained for this chamber by using an alternative dust dispersion system developed in Switzerland.

5.2.5.3 Acceleration of Dust Explosion in Elongated Systems

Coal mines essentially consist of long galleries of large length-to-diameter ratio (L/D). In the extensive systematic research on the propagation of coal dust explosions in coal mines, large-scale experimental galleries have been a main tool of investigation. The tests in experimental coal mines in UK about 1890 were probably the first of this kind.

Fischer (1957) reported results from coal dust explosion experiments in a 260 m long experimental coal mine gallery of equivalent-circle cross-sectional diameter of 3.2 m, i.e. L/D of about 80. The main purpose of these experiments was to investigate whether deposits of stone dust on shelves in the upper part of the gallery cross-section would prevent the propagation of coal dust explosions in the gallery. However, it appeared that under certain circumstances this stone dust had little effect, and very violent flame acceleration phenomena were observed, as shown in Figure 5–20.

The coal dust explosion was initiated by an explosion of 40 m^3 methane/air at the upstream closed end of the gallery. The gas was ignited by black powder probably ensuring violent combustion of the gas. The blast from the gas explosion in turn swept up the coal dust layer of 4 kg per m

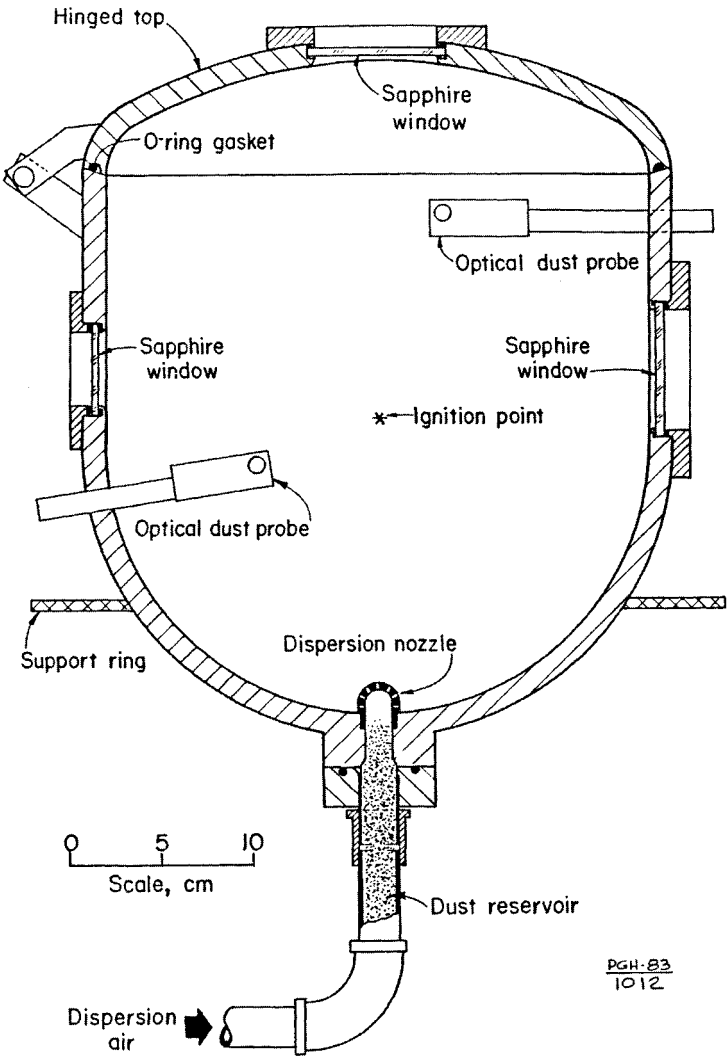


Figure 5–19 Cross-section of U.S. Bureau of Mines' 20 liter explosion vessel for determination of maximum explosion pressure, maximum rate of rise of explosion pressure, minimum explosive dust concentration, and other parameters of explosive dust clouds. From Cashdollar and Hertzberg (1985).

length of gallery on the floor and initiated the self sustained dust explosion down the entire length of the gallery. The most striking feature of Figure 5–20 is the very constant flame speed of 1040 m/s measured from about 50 m from the closed end right to the open tube end 200 m further

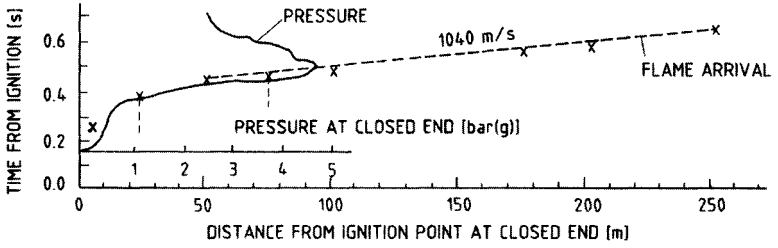


Figure 5–20 Time of arrival of bituminous coal dust/air flame as a function of distance from ignition point at closed end of gallery of length 260 m and diameter 3.2 m. Pressure at closed end as a function of time. Nominal average dust concentration 500 g/m^3. From Fischer (1957).

down. Fischer associated this with "some kind of detonation" (see Section 5.2.6). The pressure versus time was recorded only at the upstream closed end of the gallery, because the explosion was so violent that all the measurement stations further down the gallery were destroyed. As can be seen, the peak pressure at the closed end was about 5 bar(g). It would be anticipated that the pressures further down the gallery were considerably higher.

Jost and Wagner (in Freytag, 1965) have illustrated the various characteristic phenomena occurring during acceleration of premixed gas flames in long one-end-open tubes. There are good reasons for assuming that their overall picture, as reproduced in Figure 5–21, also applies to dust clouds.

The only major difference is that a dust cloud needs to be generated by raising dust deposits into suspension. This means that stage one and possibly also stage two in Figure 5–21, the ignition and laminar propagation of the initial flame, may not be relevant for dust flames. As already discussed, Fischer (1957) used a turbulent gas flame for initiating the dust entrainment and explosion. However, once the dust explosion gets under way, the blast wave generated by it will entrain dust further downstream as already discussed. Therefore all stages of Figure 5–21, from stage three and downwards, will apply even to dust clouds. The essential reason for the flame acceleration is turbulence generated in the unburned cloud ahead of the flame due to wall friction when the cloud is pushed towards the open tube end by the expansion of the part of the cloud that has burned. The cloud combustion rate increases as soon as the flame front reaches the turbulent region of the unburned cloud. This in turn increases

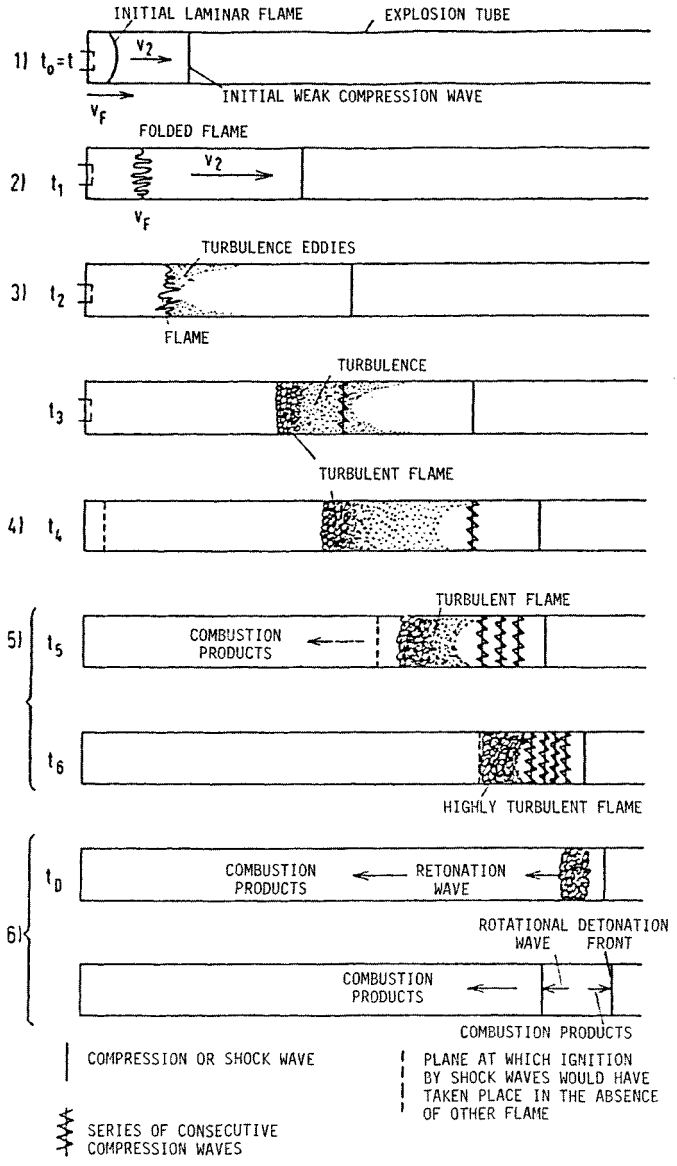


Figure 5–21 Characteristic phenomena during acceleration of gas or dust flames in one-end-open long tubes, from laminar combustion via turbulent combustion to detonation. V_F is the flame speed; V_2 is the velocity of the unburned gas or dust cloud ahead of the flame. From Jost and Wagner, in Freytag (1965).

the expansion rate of the combustion products and therefore also the flow rate of the unburned cloud further ahead. The result is an even higher turbulence level and further increase of the combustion rate.

During all these stages, compression waves will be emitted and propagate towards the open tube end. Because of heating of the cloud ahead of the flame due to adiabatic compression, each wave will propagate at slightly higher velocity than the previous one. Ultimately, therefore, they will all catch up with the initial wave and form a strong leading shock front. The turbulent flame front will also, due to the positive feedback mechanism of combustion rate flow rate turbulence enhanced combustion rate, eventually catch up with the leading shock wave. If the leading shock is sufficiently strong, a switch can occur in the mechanism of flame propagation. Instead of heat being transferred by turbulent diffusion behind the leading shock wave, the dust cloud may become ignited in the highly compressed state inside the leading shock. If the induction time of ignition is sufficiently short, the chemical reaction zone and the propagating shock wave then become closely coupled and propagate through the cloud at constant velocity. This is *detonation*. However, as already mentioned, flame propagation at a constant high speed will not necessarily have to be a classical detonation, but can also be a high-speed turbulent deflagration supported by wall friction or repeated-obstacle-induced turbulence.

5.2.6 Detonation of Dust Clouds

5.2.6.1 Qualitative Description of Detonation

As for detonation in premixed gases (see Section 2.1.7), detonation in an explosive dust cloud is a singular, extreme mode of flame propagation through the cloud. The transfer of heat from the burning to the unburned cloud is then no longer by diffusion, which is characteristic of the deflagration mode of explosion propagation discussed so far. In stead heat is transferred by extreme, very fast compression of the unburned cloud captured in a shock wave driven through the cloud at supersonic speed by the explosion itself. The detailed mechanism of ignition and combustion inside the shocked detonation front is still a subject of research.

Detonation in a dust cloud can only by initiated by a sufficiently strong shock wave. Such a wave can either be supplied by an explosive charge

detonated inside the cloud, or by gradual build-up of a strong shock by turbulent acceleration of the explosion itself as illustrated in Figure 5–21. (Deflagration-to-Detonation-Transition, DDT.)

5.2.6.2 Experiments in Small Ducts and Large-Scale Galleries

Various workers have found clear evidence of proper dust detonations in ducts of smaller diameters. For example, Kauffman et al. (1982, 1984) demonstrated that a self-sustained steady detonation wave propagated through clouds of oats and wheat grain dust in air, inside a vertical laboratory scale duct of 6.35 cm × 6.35 cm square cross-section and 6 m length. The dust was charged into the duct at the top at a mass rate giving the desired dust concentration during gravity settling down the tube. The dust explosion was initiated by a local hydrogen/oxygen explosion at the bottom tube end. At an oats dust concentration of 250–270 g/m^3 (slightly below stoichiometric) the measured detonation front velocity was 1540 m/s, which is somewhat lower than the theoretical Chapman-Jouguet velocity at stoichiometric concentration of about 1800 m/s. This is in accordance with expectations because inevitable energy losses in a real dust detonation would cause the real detonation velocity to be somewhat lower than the ideal C-J velocity. The highest measured peak pressure was, however, about 24 bar, quite close to the theoretical C-J pressure at stoichiometric concentration of 22.4 bar.

Kauffman et al. (1984) also investigated the upper and lower dust concentration limit for detonations of oats dust in air in their laboratory-scale vertical tube. They found that detonations could only be initiated, even with very vigorous ignition sources, within the narrow concentration range of approximately 200–450 g/m^3.

Gardner et al. (1986) provided further conclusive evidence of proper detonations being possible in clouds of high-volatility coal dusts in air, and that such detonations can be brought about by turbulent flame acceleration (DDT), as illustrated in Figure 5–21. The experimental arrangement consisted of a 20 m^3 ignition chamber connected to a 42 m long straight cylindrical test duct of diameter 0.6 m, which was essentially open at the downstream end. To start the experiments, air was blown through the system at a volumetric rate giving an average velocity of 20–30 m/s in the duct. Dust was then fed into the air stream just upstream of the 20 m^3 chamber to give the desired dust concentration, ranging from 30 g/m^3 to

850 g/m^3, in the explosion chamber as well as in the 42 m long duct. The dust cloud was ignited in the 20 m^3 chamber by a flame jet or a chemical ignitor. An extreme experimental transient peak pressure value of 81 bar(g) was measured in the duct when DDT occurred. This agrees with the general experience that at DDT there is always a regime in which the detonation wave is overdriven before settling down to the C-J conditions. During this transient period, the detonation pressure can exceed the C-J value considerably.

5.2.6.3 Unconfined Dust Cloud Detonations

Tulis (1984) provided direct experimental evidence of a self-sustained detonation wave being able to travel through an unconfined dust cloud in air. In these experiments a very fine aluminum flake powder of specific surface area 3–4 m^2/g was used. The unconfined cloud was about 10 m long and about 1 m high. When this cloud was initiated at one end with a 2.3 kg high-explosive charge, stable, self-sustained detonation was achieved. The average velocity of the detonation wave was 1750 m/s, and the peak pressures in excess of 28 bar. The corresponding calculated C-J values were 1850 m/s and 26 bar. This close agreement between experiment and theory seems to confirm that the phenomenon observed was in fact a proper, unconfined dust cloud detonation.

5.2.7 Primary and Secondary Dust Explosions

One important objective of dust explosion mitigation is to limit dust explosions in process equipment to the process unit in which they were initiated. This first explosion inside a process unit is called the primary explosion. A central concern in all dust explosions protection efforts is to avoid *secondary* explosions outside process equipment caused by entrainment and ignition of dust layers accumulated there by the blast wave and flame from the primary explosion, as illustrated in Figure 5–22.

As illustrated in Figure 5–2, there is a gap of about two orders of magnitude between the maximum explosive dust concentration and the bulk density of dust layers and heaps. The consequence of this with regard to the volume of dust cloud that can be produced by a given volume of dust layer, is illustrated in Figure 5–23.

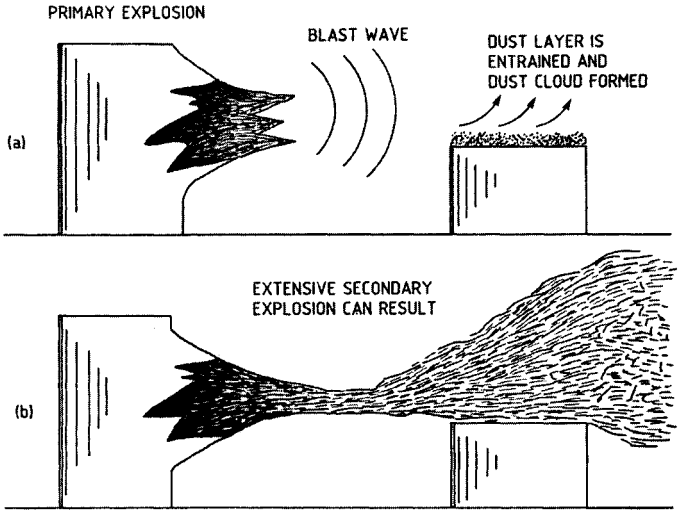


Figure 5–22 Illustration of how the blast wave from a primary explosion entrains and disperses a dust layer, which is subsequently ignited by the primary dust flame. From Eckhoff (2003).

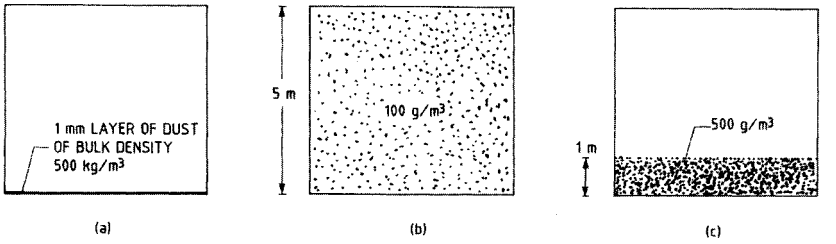


Figure 5–23 Illustration of the potential hazard of even thin dust layers. A 1 mm layer of a dust of bulk density 500 kg/m^3 will generate a cloud of average concentration 100 g/m^3 if dispersed in a room of 5 m height. Partial dispersion up to only 1 m gives 500 g/m^3. From Eckhoff (2003).

The simple relationship between the dust concentration c and the bulk density of the dust layer ρ_{bulk}, the layer thickness h, and the height H of the dust cloud produced from the layer, is:

$$c = \rho_{bulk}\,(h/H) \qquad (5.1)$$

Next Page

If a dust layer of thickness h on the internal wall of a cylindrical duct of diameter D, is dispersed homogeneously over the whole tube cross-section, one has

$$c = \rho_{bulk} (4h/D) \quad \textbf{(5.2)}$$

In the case of a tube diameter of 0.2 m, typical of many dust extraction ducts in industry, a layer thickness of only 0.1 mm is sufficient for generating a dust concentration of 1000 g/m^3 with a dust of bulk density 500 kg/m^3. In general, dispersible dust layers in process plants represent a potential hazard of extensive secondary dust explosions, which must be reduced to the extent possible.

5.3 Ignition of Dust Clouds in Air

5.3.1 Introduction

A combustible dust cloud will not start to burn unless it becomes ignited by a source of heat of sufficient strength. The most common ignition sources are:

- smoldering or burning dust
- open flames (welding, cutting, matches, etc.)
- hot surfaces (hot bearings, dryers, heaters, etc.)
- heat from mechanical impacts
- electrical discharges and arcs

In addition, there are some more sophisticated potential ignition sources including:

- laser light
- adiabatic compression and shock waves
- ultrasonic waves

There is considerable variation in the ignition sensitivity of clouds in air of various types of dusts, not least due to variations chemical composition including moisture content, and particle size. In order to quantify the ignition sensitivity of dust clouds and dust deposits, when exposed to various

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kinds of ignition sources, a range of laboratory-scale test methods have been developed.

5.3.2 Smoldering or Burning "Nests"

As discussed in Chapter 4, experience has shown that combustible dusts, when deposited in heaps or layers, may under certain circumstances develop internal smoldering nests of high temperatures. This is due to the porous structure of dust deposits, which gives oxygen access to the particle surface throughout the deposit, and also makes the heat conductivity of the deposit low. The initial oxidation inside the deposit may sometimes be due to the deposited dust or powder having initially a higher temperature than intended or permissible. However, some natural vegetable materials may develop initial spontaneous combustion even at normal ambient temperatures due to biochemical activity, if the content of fat and/or moisture is high.

If a dust deposit containing such a hot reaction zone, often called a 'smoldering nest', is disturbed and dispersed by an air blast or some mechanical action, the burning dust may initiate a dust explosion if brought in contact with a combustible dust cloud. In other cases the dust deposit or layer rests on a heated surface, which supplies the heat needed to trigger self-ignition in the dust. Such surfaces can be overheated bearings, heaters in workrooms, light bulbs, walls in dryers, etc. Dust deposits on heated surfaces may prevent normal cooling by forming an insulating layer. This may cause a further temperature rise in the surface, which will, in turn, increase the likelihood of ignition of the dust further. In general the minimum temperature of a hot surface for a layer of a given dust to self-ignite decreases with increasing thickness of the dust layer.

Pinkwasser (1985, 1986) studied the possibility of dust explosions being initiated by smoldering lumps ('nests') of powdered material that was conveyed through a process system. The object of the first investigation (1985) was to disclose the conditions under which smoldering material that had entered a pneumatic conveying line would be extinguished, i.e. cooled to a temperature range in which the risk of ignition in the downstream equipment was no longer present. In the case of > 1 kg/m^3 pneumatic transport of screenings, low-grade flour and C3 patent flour, it was impossible to transmit a 10 g smoldering nest through the conveying line

any significant distance. After only a few m, the temperature of the smoldering lump had dropped to a safe level. In the case of lower dust concentrations, between 0.1 and 0.9 kg/m^3, i.e. within the most explosive range, the smoldering nest could be conveyed for an appreciable distance, but no ignition was ever observed in the conveying line.

In the second investigation (1986) smoldering nests of 700°C were allowed to fall freely through a 1 m tall column containing dust clouds in air of 100–1000 g/m^3 of wheat flour or wheat starch. Ignition was never observed during free fall. However, in some tests with nests of at least 25 mm diameter and weight at least 15 g, ignition occurred immediately after the nest had come to rest at the bottom of the test column. This indicates that a smoldering nest falling freely through a dust cloud in a silo without disintegrating during the fall, has a higher probability of igniting the dust cloud at the bottom of the silo than during the fall. This result agrees with the findings of Alfert, Eckhoff and Fuhre (1989), who studied the ignition of dust clouds in a 22 m tall silo of diameter 3.7 m by falling smoldering nests. It was found that nests of low mechanical strength disintegrated during the fall and generated a large fire ball that ignited the dust cloud. With mechanically stable nests, ignition of the dust cloud occurred either some time after the nest had come to rest at the silo bottom, or when the best became disintegrated during the impact with the silo bottom. However, as soon as the nest had come to rest at the silo bottom, it could also become covered with dust before ignition of the dust cloud got under way.

5.3.3 Open Flames

The flames of welding and cutting burners are sufficiently powerful to initiate explosions in any dust cloud that is at all able to propagate a self-sustained flame. The cutting burner flame is particularly hazardous because it supplies excess oxygen to the working zone. All codes and regulations for preventing dust explosions contain strict requirements to the safety precautions that have to be taken when performing hot work in areas containing dust. Smoking should be prohibited in areas where combustible dusts exist. A burning wooden match develops about 100 J of thermal energy per second. This is more than sufficient for initiating explosions in most combustible dust clouds.

A pragmatic test method for assessing whether a dust cloud in air can be ignited by an open flame is illustrated in Figure 5–24. The test is based on the consideration that a welding burner flame is amongst the strongest ignition sources encountered in industrial practice. A vertical tube of length 40 cm and diameter 14 cm, open at both ends, was fitted with a U-shaped dust dispersion tube and an acetylene/oxygen welding burner. A quantity of the powder to be tested was placed at the bottom of the dispersion tube, and a controlled blast from a compressed air reservoir dispersed the dust into a cloud in the main tube where it was immediately exposed to the hot flame from the welding burner. The amount of powder and the dispersion air pressure was varied to produce optimal conditions for ignition. When a dust flame occurred, its maximum height, color and apparent violence were assessed by the observer.

5.3.4 Hot Surfaces

Besides igniting dust layers, hot surfaces can initiate dust explosions by direct contact between the dust cloud and the hot surface. However, the minimum hot surface temperatures needed for this are generally considerably higher (typically 400–500°C for organic dusts) than for ignition of dust layers.

In the U.S.A., the ignition temperature of dust clouds in contact with a hot surface was traditionally determined in the Godbert-Greenwald furnace, which is illustrated in Figure 5–25.

In this apparatus the internal surface of a vertical cylindrical ceramic tube, open at the lower end, is kept at a known, constant temperature, and a sample of the powder is dispersed as a dust cloud into the tube from above by means of a blast of air. The automatically controlled temperature of the internal wall of the tube is changed in steps and the experiment repeated until the minimum temperature for ignition has been identified. The International Electrotechnical Commission (IEC) investigated the performance of the Godbert-Greenwald furnace through several round-robin test series involving several central test laboratories in Europe and the U.S.A. The influences of a number of details of the apparatus itself and of the experimental procedure were studied and details of apparatus and procedure specified more closely. The resulting, improved Godbert-Greenwald furnace has now been adopted by IEC as a standard method for determining minimum ignition temperatures of dust clouds.

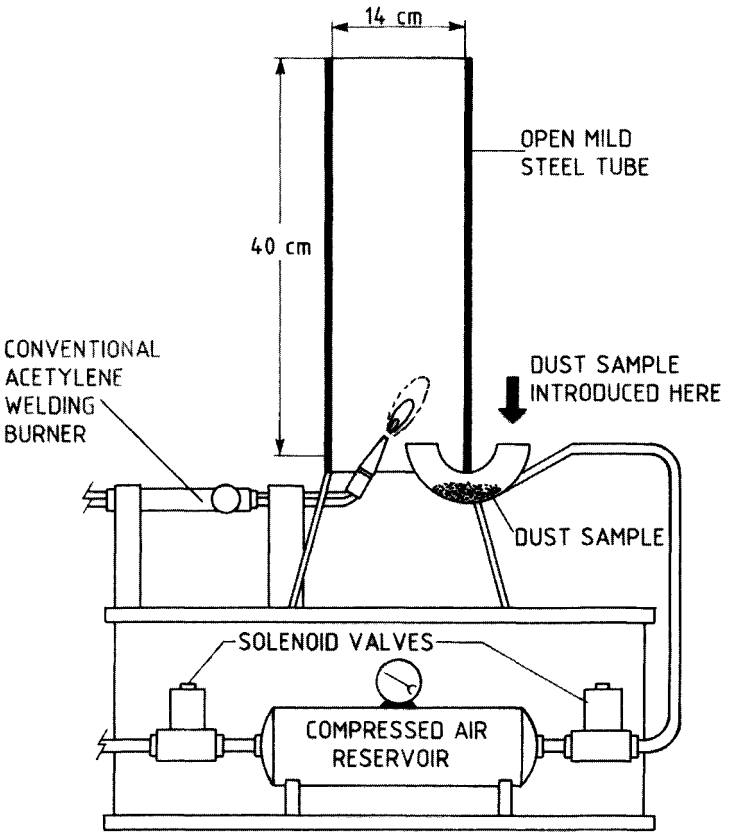


Figure 5–24 Welding torch ignition test apparatus used in Norway for assessing whether or not a dust cloud can be ignited by an open flame. From Eckhoff (2003).

5.3.5 Heat from Mechanical Impacts

The literature on dust explosions is sometimes confusing when discussing ignition of dust clouds by heat from mechanical impacts. This is reflected in the use of terms such as "friction" or "friction sparks" when categorizing ignition sources. In order to clarify the situation, it seems useful to distinguish between friction and impact.

Friction is a process of fairly long duration whereby objects are rubbed against each other and heat is gradually accumulated. This produces hot

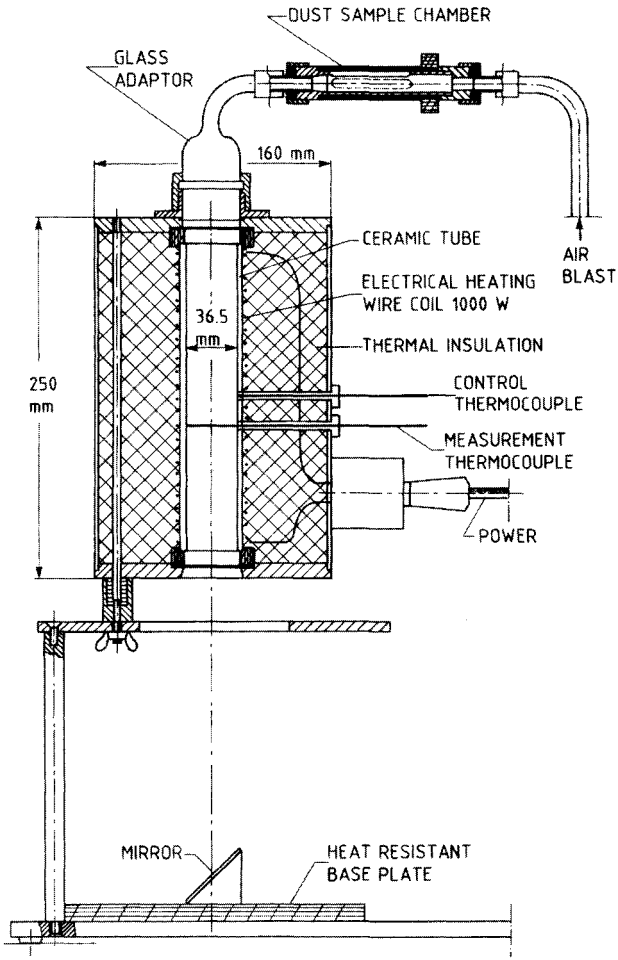


Figure 5–25 Godbert-Greenwald furnace for determination of the minimum ignition temperature of dust clouds. From Eckhoff (2003).

surfaces, and in some cases inflammation, for example when an elevator or conveyor belt is slipping.

Impact is a short-duration interaction between two solid bodies under conditions of large transient mechanical forces. Small fragments of solid material may be torn off, and if made of metal, they may start burning in air due to the initial heat absorbed in the impact process. In addition, local *hot-spots* may be generated at the points of impact. In some cases the

impact may occur repeatedly at one specific point, for example when some misplaced stationary object inside a bucket elevator is repeatedly hit by the buckets. This may gradually generate a hot spot of sufficient size and temperature to ignite the dust cloud directly.

A practical mechanical impact situation may be as follows: A steel bolt is accidentally entering the top of a large concrete silo during filling of the silo with maize starch. The bolt falls down into the nearly empty silo and hits the concrete wall near the silo bottom at a velocity of 25–30 m/s. Visible sparks are generated. A dense, explosive cloud of maize starch occupies the region where the impact occurs. Is ignition of the cloud probable? A test method for investigating this problem is shown in Figure 5–26.

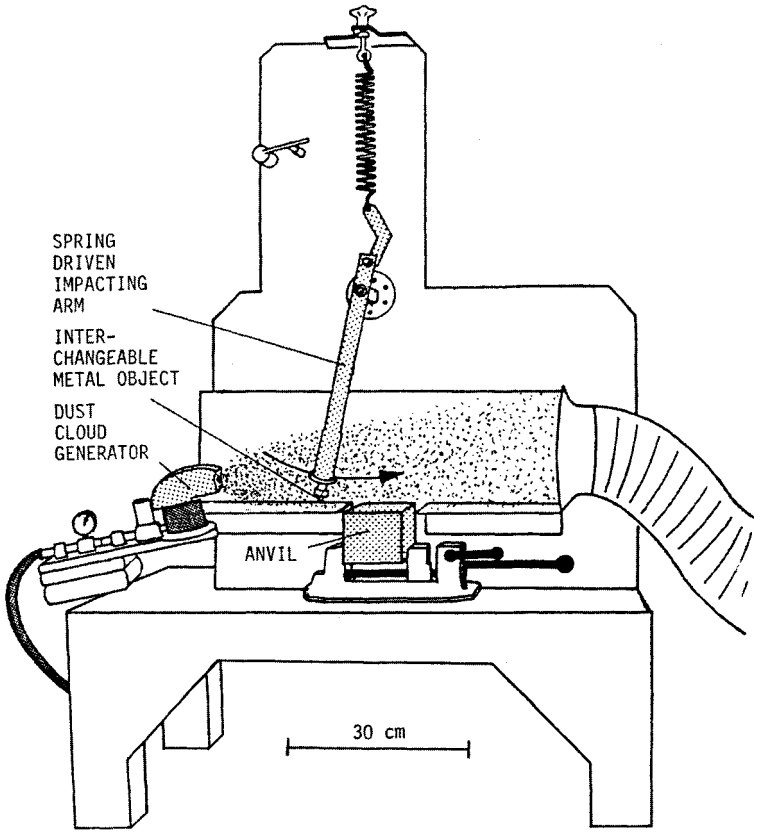


Figure 5–26 Apparatus for determining the sensitivity of dust clouds to ignition by single accidental mechanical impacts. From Pedersen and Eckhoff (1987).

Experiments using this apparatus indicated that ignition of clouds of natural organic materials by metal sparks from single impacts, where steel is the metal component, seems less likely than believed by many in the past. However, if the metal is titanium or zirconium, ignition can occur quite readily.

The thermite reaction ($2Al + Fe_2O_3 \rightarrow Al_2O_3 + 2Fe + heat$) is often mentioned as a potential ignition source from impacts involving aluminum and rust. However, if a lump of normal soft aluminum collides with a rusty steel surface, a thermite reaction will not necessarily take place. In fact, due to the softness of the aluminum, the result is often just a thin smear of aluminum on top of the rust. However, if this sandwich of aluminum and rust is given a blow by a third hard object, a thermite flash capable of igniting dust clouds can easily be produced. The same applies to a rusty surface that has been painted with aluminum paint, if the pigment content of the paint is comparatively high.

5.3.6 Electric Sparks and Arcs; Electrostatic Discharges

5.3.6.1 Introduction

It has been known for more than 100 years that electric sparks and arcs can initiate dust explosions. The minimum spark energy required for ignition varies with the type of dust, the effective particle size distribution in the dust cloud, the dust concentration and turbulence, and the spatial and temporal distribution of the energy in the electric discharge or arc. For many decades it was thought that the electric spark energies needed for igniting dust clouds in air were generally much higher, by one or two orders of magnitude, than the minimum ignition energies for gases and vapors in air. However, it has now become generally accepted that many dusts can be ignited by spark energies in the range 1–10 mJ, and even below 1 mJ, i.e. in the range typical of gases and vapors.

5.3.6.2 Inductive and Capacitive Spark Discharges

5.3.6.2.1 General

One distinguishes between inductive sparks or arcs generated when live electric circuits are broken, either accidentally or intentionally (e.g. in switches), and discharges caused by release of accumulated electrostatic charge.

In the former case, if the points of rupture are separated at high speed, transient inductive sparks are formed across the gap as illustrated in Figure 5–27. If the current in the circuit prior to rupture is i and the circuit inductance L, the theoretical spark energy, neglecting external circuit losses, will be $1/2Li^2$. As an example, a current of 10 A and L equal to 10^{-5} H gives a theoretical spark energy of 0.5 mJ. This is too low for igniting most dust clouds in air. However, larger currents and/or inductances can easily give incendiary sparks. Sometimes rupture only results in a small gap of permanent distance. This may result in a hazardous stationary arc if the circuit stays live.

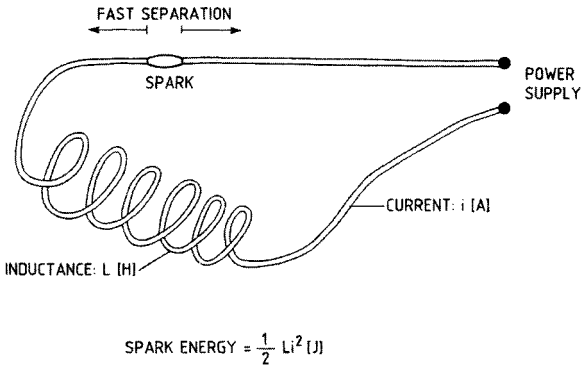


Figure 5–27 Inductive spark or "break flash" generated when a live electric circuit is suddenly broken and the points of rupture are separated at high speed. From Eckhoff (2003).

Capacitive spark discharges occur when charge that has accumulated on an electrically conducting, unearthed, object is discharged to earth across a small air gap. In the process industries producing and handling powders, electrostatic charging is generally tribo-electric, which implies transfer of electrons between objects of different electron affinity during contact and subsequent separation. In a process plant this occurs during handling and transport of powders and dusts whenever the powder/dust and the process equipment make contact and separates.

An electrostatic spark discharge results if the discharge of the accumulated charge occurs between two electrically conducting electrodes. The spark gap distance must then be sufficiently short to allow breakdown and spark channel formation at the actual voltage difference between the charged object and earth. On the other hand, in order for the spark to become incendiary, the gap distance must be sufficiently long to permit

the required voltage difference to build up before break-down of the gap occurs. The theoretical spark energy, neglecting external circuit losses, equals $1/2CU^2$, where C is the capacitance of the un-earthed, charged process item with respect to earth, and V is the voltage difference. Figure 5–28 illustrates a practical situation that could lead to a dust explosion initiated by an electrostatic spark discharge.

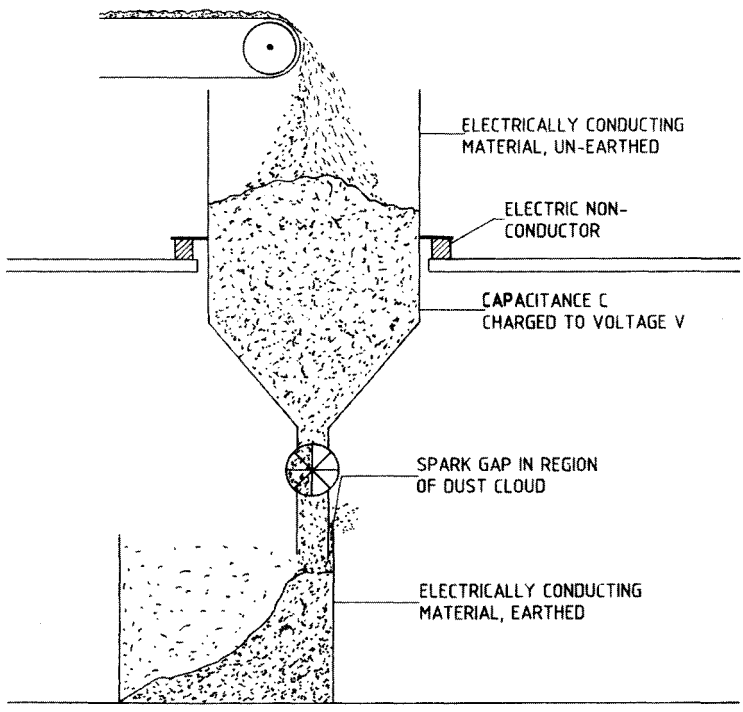


Figure 5–28 Illustration of a practical situation that could lead to a dust explosion initiated by an electrostatic spark discharge. From Eckhoff (2003).

Glor (1988) estimated typical approximate capacitance-to-earth values for objects encountered in the process industry. The values given in Table 2–11 in Section 2.2 (ignition of gases) are equally valid in the context of dust explosions. Such capacitance values are used for estimating the maximum theoretical spark energies $1/2CU^2$ when discharging a non-earthed metal object of capacitance C and at a voltage U, to earth.

5.3.6.2.2 The Minimum Ignition Energy (MIE) of a Dust Cloud

The MIE of a given dust in air is a commonly used measure of the ease with which the cloud is ignited by electric sparks and electrostatic discharges. A typical apparatus for laboratory determination of MIE of dust clouds is illustrated in Figure 5–29. An appropriate quantity of the dust is placed in the dispersion cup at the bottom of the 1.2 liter plastic or glass cylinder and dispersed by a blast of air deflected by a conical "hat" as indicated. A spark of the desired energy is discharged across the electrodes synchronously with the transient appearance of the dust cloud in the spark gap region, and it is recorded visually whether or not ignition takes place. MIE is often defined as the lowest spark energy that gives at least one ignition in ten trials at the same spark energy.

As discussed by Eckhoff (2003) there can be a strong influence of the spark discharge time on MIE for dust clouds, in particular for spark energies exceeding 1 J. In standard testing for MIE optimal discharge times for ignition is achieved by introducing a 1–2 mH inductance in the discharge circuit, and this feature is compulsory in the current international (IEC) and European standard methods for MIE determination for dust clouds. This inductance is to be removed, however, when the purpose of the test is to assess the sensitivity to ignition by electrostatic spark discharges. A major limitation of the standard test apparatuses commonly used, is that they cannot generate sparks of energies significantly lower than 2–3 mJ. However, recently Randeberg et al. (2005) developed a spark generator capable of producing synchronized capacitive sparks of energies down to less than 0.1 mJ, suitable for duct cloud ignition tests.

Figure 5–30 illustrates various circuits that have been used to generate the electric spark discharges in MIE tests. In the original circuit of U.S. Bureau of Mines, a substantial part of the capacitor energy was lost in the transformer. In the case of the direct high voltage discharge circuit the switch element needed for synchronization of dust dispersion and spark discharge requires special considerations. In the CMI circuit a trigger transformer feeding 2–3 mJ into the main discharge is used for triggering the discharge. Other synchronization principles are also in use, including a fast-moving earthed electrode.

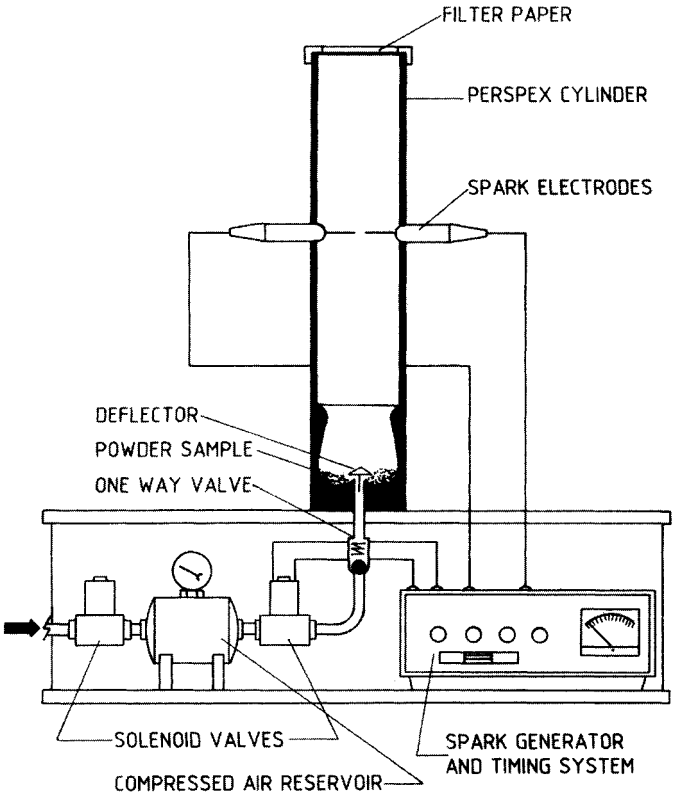
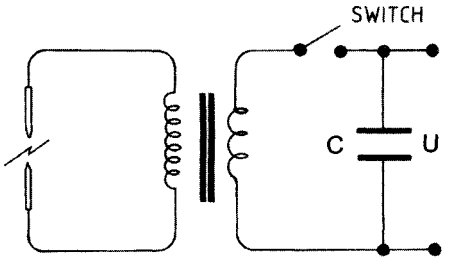


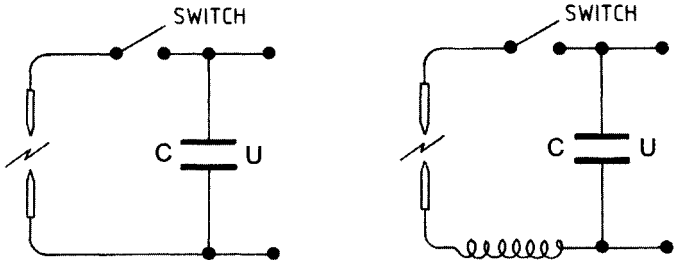
Figure 5–29 Illustration of an apparatus type commonly used for experimental determination of minimum electric spark ignition energies (MIE) of explosive dust clouds in air. From Eckhoff (2003).

5.3.6.2.3 Range of MIEs for Dust Clouds

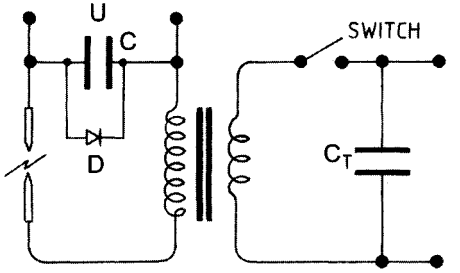
Minimum electric spark energies (MIE) for ignition of dust clouds vary, as already mentioned, with dust type, particle size, and other factors. In the past it was thought that the absolute lower limit for MIEs of dust clouds in air was of the order of 10 mJ, i.e. nearly two orders of magnitude above typical MIEs of IIA gases (see Chapter 2). However, it is now accepted that MIEs of clouds of dusts in air span over at least 8 decades, from perhaps as low as 0.01 mJ in the lower end to beyond 1kJ in the upper.



(a) LOW-VOLTAGE CAPACITOR DISCHARGED THROUGH TRANSFORMER (ORIGINAL US BUREAU OF MINES CIRCUIT)



(b) DIRECT DISCHARGE OF HIGH-VOLTAGE CAPACITOR WITHOUT AND WITH INDUCTANCE



(c) CMI-DISCHARGE CIRCUIT

Figure 5–30 Three different electric circuits used in experimental determination of minimum electric spark ignition energies (MIE) of explosive dust clouds in air. From Eckhoff (2003).

5.3.6.2.4 Conservative Ignition Curves for Dusts Clouds Based on MIE

For capacitive circuits, conservative ignition curves are obtained directly from the equation $1/2CU^2$ = MIE. Charts showing this graphically, for a range of MIE values, are given in Figure 5–31. In the case of inductive circuits, conservative ignition curves are obtained directly from the equation $1/2Li^2$ = MIE for L > 1 mH. For smaller L, the ignition current will be independent of L, and the circuit will essentially be resistive. Charts showing conservative ignition curves for inductive circuits, for a range of MIE values, are given in Figure 5–32. For resistive ignition circuits Eckhoff (2002) has suggested that an equation of the form $I = A \cdot MIE^B/U^2$, where A and B are empirical constants, may provide a first order conservative estimate. Graphs representing this equation for preliminary values of A and B are given by Eckhoff (2003). More research is needed to establish more reliable resistive circuit ignition data in the range of higher MIEs.

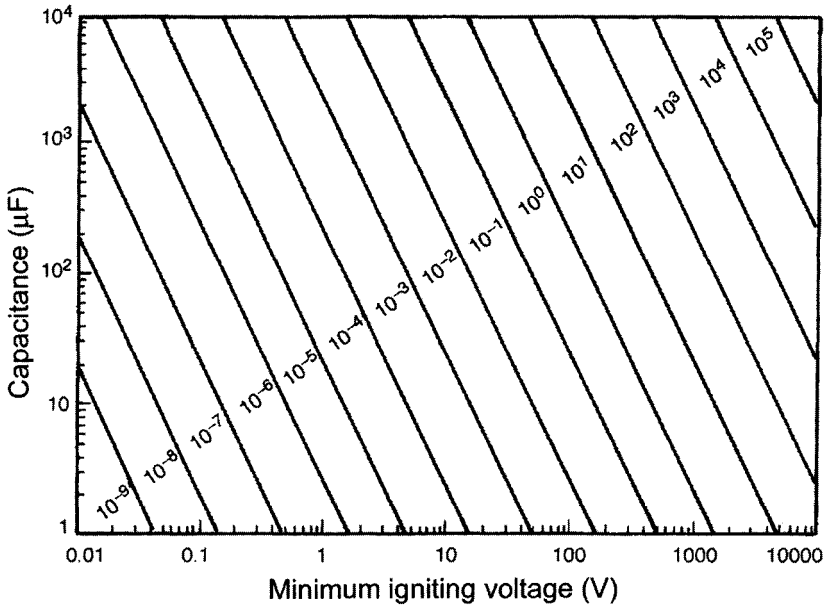


Figure 5–31 Theoretical conservative capacitive ignition curves for dust clouds based on the equation $1/2CU^2$ = MIE. The numbers attached to the straight lines are the respective MIE values in J. From Eckhoff (2003).

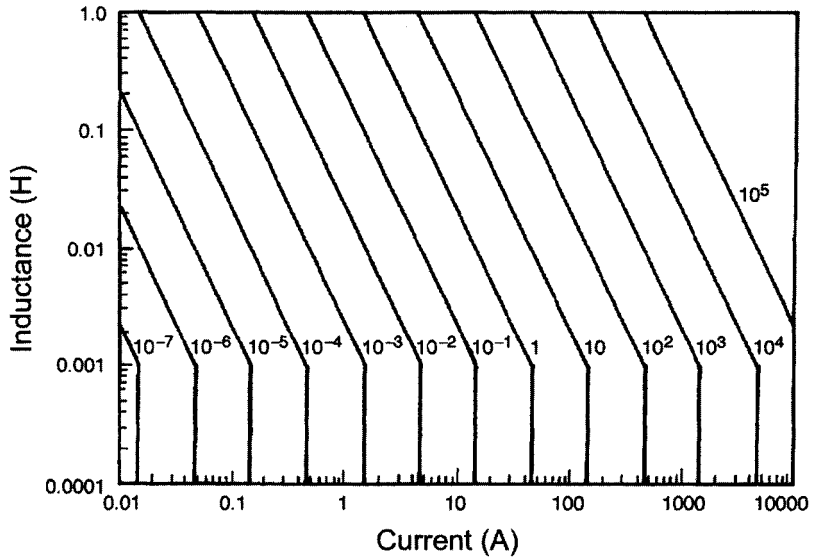


Figure 5–32 Theoretical conservative inductive ignition curves for dust clouds based on the equation $1/2Li^2$ = MIE, valid for L > 1 mH. The numbers attached to the straight lines are the respective MIE values in J. From Eckhoff (2003).

5.3.6.2.5 Parameters Influencing measured MIEs of Dust Clouds

Turbulence in the dust cloud raises the effective MIE and therefore provides a safety factor as far as ignition is concerned. This is illustrated by the work of Yong Fan Yu (1985), who was unable to ignite turbulent explosive clouds of wheat grain dust in a container at the exit of a pneumatic transport pipe, even with highly incendiary "soft" electric sparks of energies in excess of 1 J. In the type of test illustrated in Figure 5–29 the turbulence of the dust cloud at the moment of ignition can be varied by varying the delay between dust cloud generation and ignition. The larger the delay, the lower the turbulence. Figure 5–33 gives some results of MIE as a function of ignition delay.

Figure 5–34 shows how particle size influences the minimum ignition energy for three different dusts. The vertical scale is logarithmic, and it is seen that the effect is very strong. Kalkert and Schecker (1979) developed a theory indicating that MIE is proportional to the cube of the particle

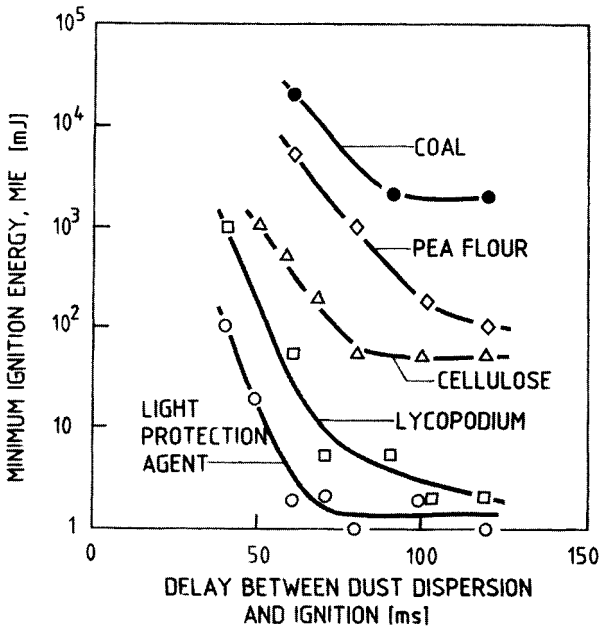


Figure 5–33 Illustration of the influence of initial turbulence of explosive dust clouds on the minimum electric spark energies required for ignition. Data from Glarner (1984).

diameter, as also illustrated in Figure 5–34 by their theoretical prediction of the relationship for polyethylene.

Moisture in the dust reduces both ignition sensitivity and explosion violence of dust clouds. Figure 5–35 illustrates the influence of dust moisture on MIE. The vertical axis is logarithmic, and it is seen that the effect is quite significant. If safety measures against electric spark ignition are based on MIE data for a given dust moisture content, it is essential that this content is not subsided in practice.

5.3.6.3 Electrostatic Discharges Other Than Sparks

5.3.6.3.1 Overview

Glor (1988) and Lüttgens and Glor (1989) distinguished between five different types of electrostatic discharges in addition to sparks, namely:

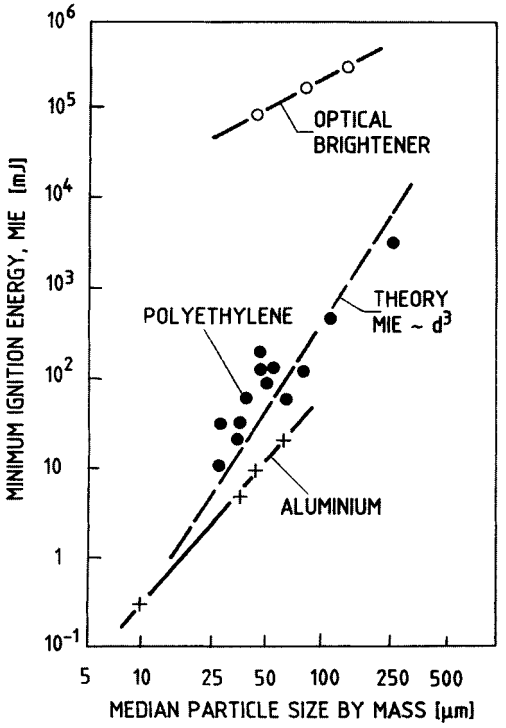


Figure 5–34 Minimum experimental electric spark ignition energies (MIE) of clouds in air of an optical brightener, polyethylene and aluminum, as functions of median particle size, from Bartknecht (1987), and theoretical line for polyethylene. From Kalkert and Schecker (1979).

- brush discharge
- corona discharge
- propagating brush discharge
- discharge along the surface of powder/dust in bulk
- lightning-like discharge

5.3.6.3.2 Corona Discharges

Corona discharges, illustrated in Figure 2–33 in Section 2.2, occur under the same conditions as brush discharges, but are associated with earthed electrodes of much smaller radii of curvature, such as sharp edges and needle tips. For this reason such discharges will occur at much lower field

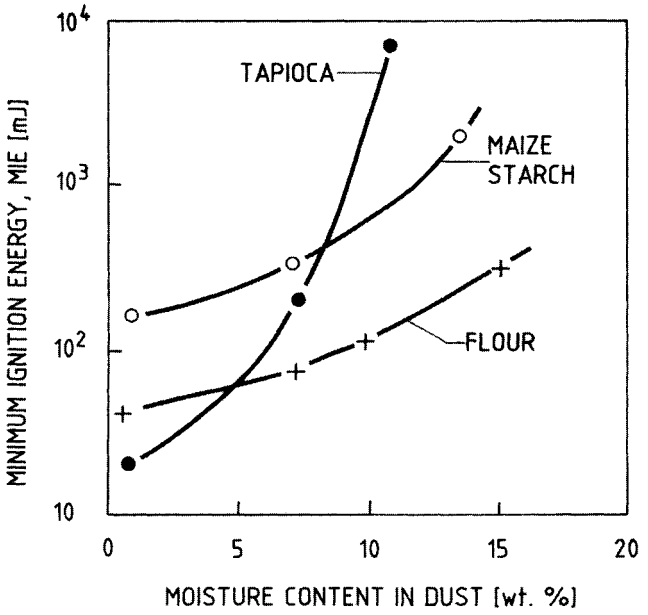


Figure 5–35 Influence of dust moisture content on minimum electric spark ignition energy (MIE) for three dusts. From van Laar and Zeeuwen (1985).

strengths than the brush discharges, and the discharge energies will therefore also be much lower. Consequently, the possibility of igniting dust clouds by corona discharges can be ruled out.

5.3.6.3.3 Brush Discharges

Brush discharges, illustrated in Figure 2–34 in Section 2.2, occur between a single curved, earthed metal electrode (radius of curvature 5–50 mm) and a charged non-conducting surface (plastic, rubber, dust). Brush discharges can ignite explosive gas mixtures. However, according to Glor (1988), no ignition of a dust cloud by a brush discharge had yet been demonstrated, not even in sophisticated laboratory tests using very ignition sensitive dusts. The conclusion that brush discharges cannot ignite dust clouds in air is strongly supported by the more recent experiments by Larsen et al. (2003). It must be emphasized, however, that this does not apply if the powder/dust contains significant quantities of combustible solvents, which can increase the ignition sensitivity of the cloud substantially, even when the vapor concentration in the air is well below the LEL

of the vapor alone. Such clouds of a combustible dust suspended in a mixture of a combustible gas or vapor and air are called "hybrid" mixtures.

5.3.6.3.4 Propagating Brush Discharges

Propagating brush discharges, illustrated in Figure 2–35 and Figure 2–36 in Section 2.2, can, however, initiate dust explosions. Such discharges, which will normally have much higher energies than ordinary brush discharges, occur if a double layer of charges of opposite polarity is generated across a relatively thin sheet (< 8 mm thickness) of a non-conducting material (Glor, 1988). The reason for the high discharge energy is that the opposite charges allow the non-conductor surfaces to accumulate much higher charge densities than if the sheet had been charged on only one of the faces. Glor pointed out that in principle close contact of one of the faces of the sheet with earth is not necessary for obtaining a charged double layer. By adding electrons to one side of a plastic sheet and removing electrons from the other, the same result can be obtained. However, in practice earth on one side is the most common configuration. In the context of powders and dusts Figure 2–36 may illustrate pneumatic transport of a powder/dust in a steel pipe with an internal electrically insulating plastic coating. Due to the rubbing of the powder against the plastic, charge is accumulated on the internal face of the plastic coating. The high mobility of the electrons in the steel causes build-up of a corresponding charge of opposite polarity on the outer face of coating in contact with the steel. If a short passage between the two oppositely charged faces of the plastic coating is provided, either via a perforation of the coating or at the pipe exit, a propagating brush discharge can result. Lüttgens and Glor (1989) give an example of a dust explosion that was initiated by a propagating brush discharge. Acrylic powder was transported pneumatically in a 50 mm diameter plastic pipe outdoors, and rainwater and snow provided the earthed electrically conducting shield on the outer surface of the pipe.

Glor (1988) identified five typical situations which may lead to propagating brush discharges during transport and handling of powders:

- high-velocity pneumatic transport of powder through an electrically insulating pipe, or a conductive pipe with an insulating internal coating

- use of inspection windows of glass or Plexi glass in pneumatic transport pipes
- continuous impact of powder particles onto an insulating surface (e.g. a coated dust deflector plate in the cyclone of a dust separator)
- fast movement of conveyor or transmission belts made of an insulating material, or of a conductive material coated with an insulating layer of high dielectric strength
- filling of large containers or silos made of insulating material (e.g. flexible intermediate bulk containers) or of metallic containers or silos coated internally with an insulating layer of high dielectric strength

5.3.6.3.5 Electrostatic Discharge Along the Surface of Powder/Dust in Bulk

This type of discharge may occur if strongly non-conducting powders are blown or poured into a large container or silo. When the charged particles settle in a heap in the container, very high space charge densities may be generated and luminous discharges may propagate along the surface of the powder heap from its base to its top. However, under realistic industrial conditions only very large particles, of 1–10 mm diameter, are likely to generate spark discharges due to this process. It further seems that very high specific electrical resistivity of the powder is also a requirement ($>10^{10}$ ohm•m) which probably limits this type of discharge to coarse plastic powders and granulates. The maximum equivalent spark energy for this type of discharge has been estimated to the order of 10 mJ, but little is known about the exact nature and incendivity of these discharges. Because of the large particle size required to generate the charge, these particles are unlikely to give dust explosions, and therefore a possible explosion hazard must be associated with the simultaneous presence of an explosive cloud of an additional, fine dust fraction. Glor (1988) pointed out that the probability of occurrence of this discharge type increases with increasing charge-to-mass ratio in the powder, and increasing mass filling rate.

5.3.6.3.6 Lightning Type Discharge

Lightning type discharge, which may in principle occur within an electrically insulating container with no conductive connection from the interior

to the earth is the last type of discharge mentioned by Glor (1988) and Lüttgens and Glor (1989). However, as Glor stated, there is no evidence that lightning discharges have occurred in dust clouds generated in industrial operations.

5.3.7 Jets of Hot Combustion Products

The basic process is the same as for explosive gas mixtures, as described in Section 2.2.7. In general terms, the maximum experimental safe gap (MESG) may be defined as the largest width of a slot that will just prevent transmission of a flame in a gas or dust cloud inside an enclosure to a similar gas or dust cloud on the outside. This definition is somewhat vague and raises several questions. It neither defines the length of the slot, nor the explosion pressure inside or the volume of the enclosure. Therefore, MESG is not a fixed constant for a given explosive cloud, but depends on the actual circumstances. MESG for dust clouds is of limited relevance in practice and has no relevance, in the context of electrical equipment enclosure design.

Schuber (1988, 1989) investigated the influence of various parameters on MESG for dust clouds. The situation addressed was possible transmission of a dust flame from one side of a rotary lock to the other, as illustrated in Figure 5–36.

The research apparatus used by Schuber is illustrated in Figure 5–37. Explosive dust clouds of desired concentrations were generated simultaneously from compressed dust reservoirs in both the large vessel (1 m^3) and the smaller vessel (40 liters) mounted inside the large one. The cloud in the smaller vessel was subsequently ignited, and it was observed whether the cloud in the large vessel was ignited by the flame jet being transmitted through the annular gap in the wall of the smaller vessel.

Schuber found that MESG increased with increasing initial turbulence in the dust clouds. This is in harmony with the increase of the minimum electric spark energy for ignition of both gases and dust clouds with increasing turbulence. In order to ensure conservative results, Schuber conducted his experiments with comparatively low initial turbulence in the dust clouds. He correlated his experimental MESG values with the product of minimum electric spark ignition energy and the dimensionless minimum ignition temperature, as shown in Figure 5–38. Because (TI + 273)/273 is in the range between two and three for most of the dusts tested, the double-logarithmic correlation in Figure 5–38 is essentially between MESG and MIE.

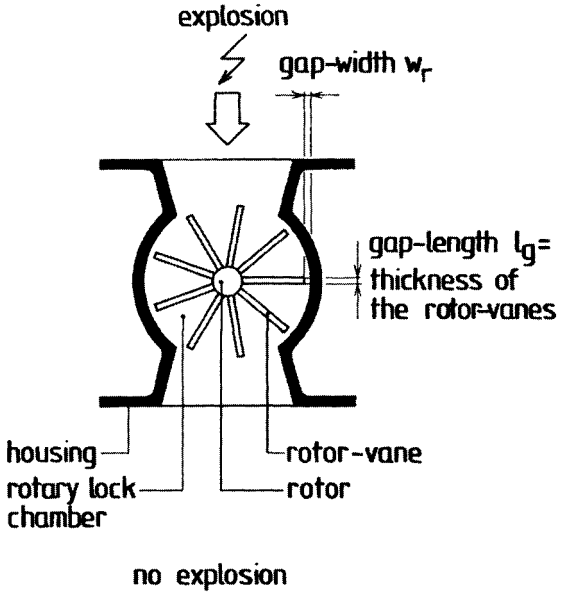


Figure 5–36 Illustration of a practical flame jet transmission situation for burning dust clouds: a rotary lock with a dust explosion occurring on one side. From Schuber (1988).

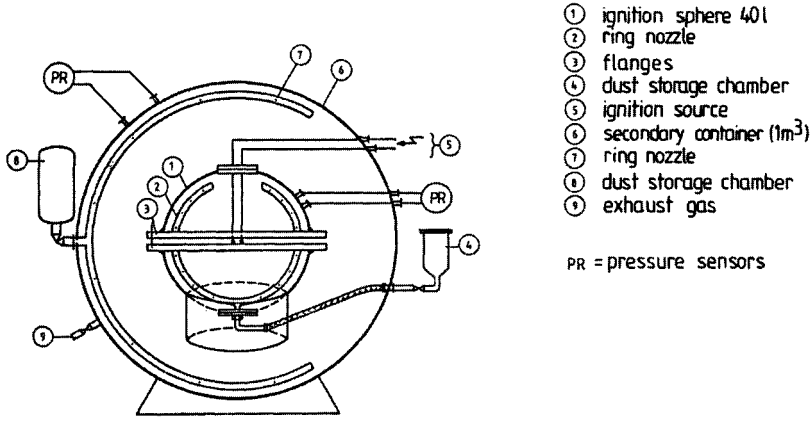


Figure 5–37 Illustration of the experimental set-up comprising a split and flanged 40 liter explosion vessel mounted inside a 1 m^3 explosion vessel. The annular gap for possible flame transmissions between the two flanges on the two halves of the 40 liter sphere. From Schuber (1989).

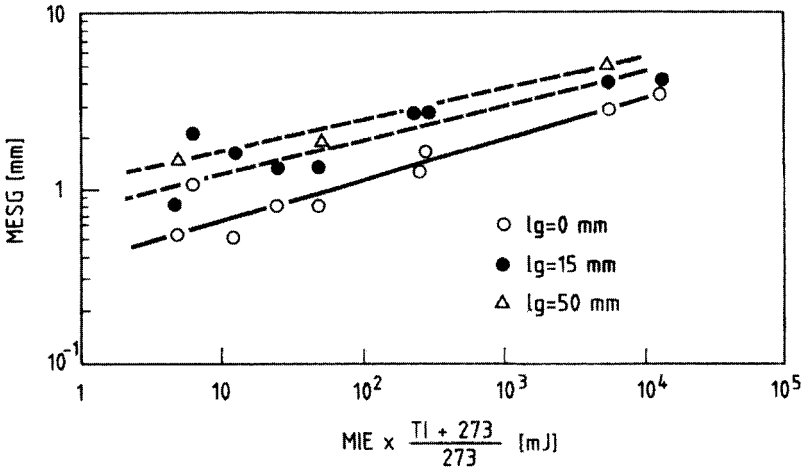


Figure 5–38 Correlations between MESG and ignition sensitivity of clouds in air of various dusts, for various gap lengths. From Schuber (1989).

As already pointed out the original motivation for Schuber's work was the uncertainty related to the ability of rotary locks to prevent transmission of dust explosions. He also carried out industrial scale experiments with a set-up as illustrated in Figure 5–36, with a rotary lock mounted between two vessels in which explosive dust clouds were generated simultaneously. The dust cloud on the one side was then ignited and it was observed if transmission of flame occurred to the extent that the dust cloud on the other side was also ignited. On the basis of numerous experiments, Schuber (1989) constructed a nomograph for predicting maximum permissible gaps between the edges of the rotary lock blades and the housing for preventing flame transmission. (See Eckhoff, 2003).

5.3.8 Shock Waves

Shock waves, i.e very steep and strong pressure wave fronts traveling at speeds larger than the speed of sound, can be generated by industrial accidents, e.g. by pressure vessel failure. An informative analysis of shock wave ignition of dust clouds was given by Wolanski (1990). His illustration of the particle interaction with the convective flow behind the incident shock wave is shown in Figure 5–39.

At the first moment of interaction, the stationary particle is subjected to the supersonic flow. A bow shock is formed near the particle; and since the tem-

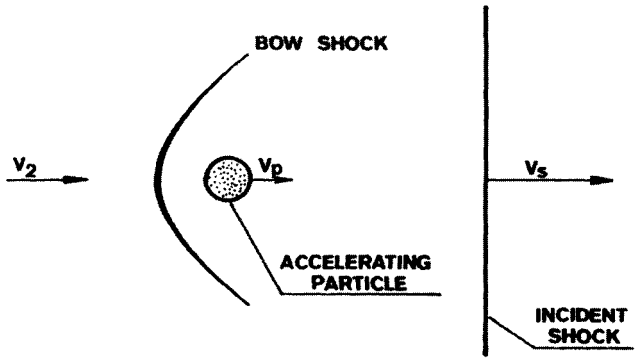


Figure 5–39 Schematic illustration of the interaction between a stationary dust particle and the convective supersonic gas flow behind the incident shock front. From Wolanski (1990).

perature of the gas between the particle and the bow shock is close to the stagnation temperature, the particle is heated rapidly. At the same time, the drag force causes the particle to accelerate and the temperature to decrease. The rate of particle acceleration depends on particle size and other properties. Because very small particles accelerate rapidly to supersonic speeds, the stagnation temperature at the particle front exists only for an extremely short time. Larger particles accelerate more slowly, thus remaining exposed to the higher stagnation temperature between the bow shock and the particle for a longer time. However, the larger particles require more heat for ignition. Wolanski suggested, therefore, that depending on the dust type, there is an optimal particle diameter range, which provides the most favorable conditions for ignition. In fact, in incident shock waves some dust particles can be more easily ignited than can even a hydrogen/oxygen mixture. According to Wolanski, experimental data do confirm that the ignition delay of dust particles behind the incident shock depends on particle diameter; and the optimum diameter for ignition varies with the nature of the particle. For organic particles the optimal diameter is of the order of 10 μm. Particles of diameters of only a few μm, or larger than 100 μm are usually more difficult to ignite in an incident shock than particles of intermediate sizes.

5.3.9 Light Radiation

Ignition experiments using laser light can provide basic information not only about dust cloud ignition processes, but also about flame propagation

processes in dust clouds. Wel et al. (1994) used a simple, modified Semenov model for auto-ignition (no temperature gradients inside initially heated volume), for using experimental laser-light-pulse ignition data to predicted minimum ignition temperatures and energies of dust clouds. The predicted values were in approximate agreement with minimum ignition temperatures and energies measured directly. Zevenbergen (2002) reviewed various aspects of ignition of dust clouds by laser light.

Proust (1996), using a Nd-YAG CW laser, investigated experimentally the ability of a beam of laser light to ignite an explosive dust cloud. He was unable to find a generally valid correlation between the ease with which a dust cloud could be ignited directly by a laser beam, and the minimum electrical spark ignition energy of the same dust cloud. Proust also conducted experiments where ignition occurred indirectly, via a small solid target in the dust cloud that had first been heated by the laser beam. He found that the probability of a given laser beam igniting a given dust cloud was considerably higher with this set-up than with direct laser beam ignition. Proust (2002) extended these experiments by using both continuous and pulsed radiation, and both targets of layers of non-combustible material, e.g. iron oxide, and of layers of the same material as in the dust cloud to be ignited. He was able to quantify minimum laser beam powers impinging on a solid target required for igniting dust clouds by the heat absorbed by the target. The parameters included the laser beam diameter, the duration of the irradiation, the target material (combustible/non-combustible), and the type of dust (starch, lycopodium, lignite, sulphur, ABS, and aluminum). An inherent complicating element in this kind of experiment is that the laser light has to travel through some length of dust cloud before impinging on the target. The part of the beam power absorbed by the dust cloud is a function of both the length of the light path through the dust cloud, the dust concentration, and the material and geometry of the dust particles. Figure 5–40 shows a set of Proust's results for pulsed laser radiation.

The minimum ignition energies in Figure 5–40 were deduced from experimental graphs of the minimum radiated laser energy received by the target required for igniting the dust cloud, as a function of the time from onset of radiation to ignition. By extrapolating the almost linear relationships found to zero time, an estimate of the minimum ignition energy with very short laser pulses hitting an inert iron oxide target, could be deduced. The minimum ignition energies found in this way, ranging from 4–14 J

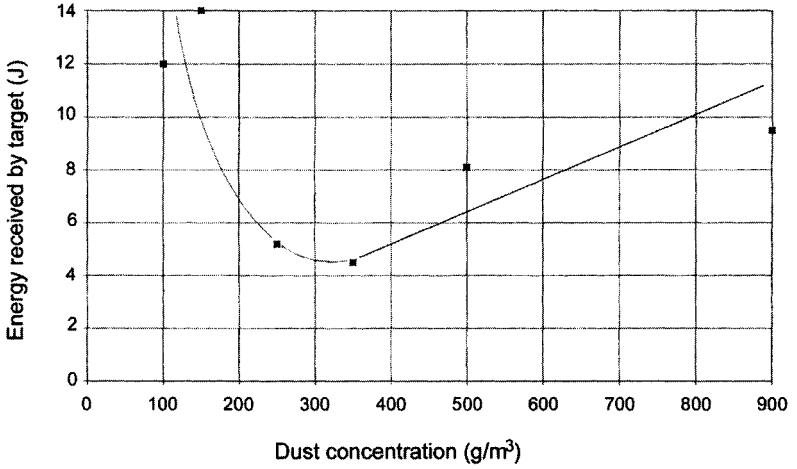


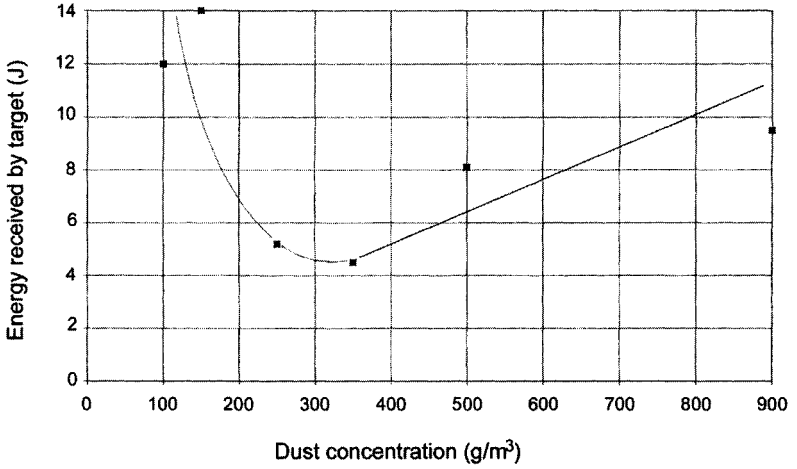
Figure 5–40 Deduced relationship for the minimum energy of a very short laser pulse, hitting a solid inert iron oxide target inside an explosive cloud of starch particles in air, required for igniting the cloud, as a function of the starch concentration in the cloud. From Proust (2002).

are substantially higher than the minimum electric spark ignition energy of 10 mJ found for the actual starch in a standard MIE test.

5.4 Case Histories of Dust Explosions

5.4.1 Motivation for Selection

Most of the case histories given in the following are taken from the book by Eckhoff (2003). Because of his close co-operation with Norwegian industry in investigating accidental dust explosions that have occured there. He has had access to detailed information on several such explosions. It was natural, therefore, to include some of this information even in the present book. However, the case history from China and the very recent one provided by CSB (2003) constitute most valuable additional examples illustrating that the dust explosion hazard continues to threaten a wide range or process industries in many countries.



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Some well known dust explosions in other countries, which have been described extensively elsewhere in the open literature, have not been included. This, for example, applies to the catastrophic wheat flour explosion in the Roland Mill in Bremen, Germany, which has been discussed in detail by the Fire and Police Authorities of Bremen (1979). It also applies to many of the major dust explosions in the U.S.A. in the 1970s and 1980s, which have been discussed in detail by Kauffman (1982, 1989) and Kauffman and Hubbard (1984).

5.4.2 Historical Perspective— Wheat Flour Explosion in Turin (1785)

The dust explosion hazard has probably been recognized in Europe for several centuries, but the flour explosion in Turin in 1785 seems to be the first accident of this kind that was investigated extensively. When the Academy of Science of Turin heard about Morozzo's investigations, they asked him to prepare a written account of his findings. Only very rarely are details of Count Morozzo's (1795) fascinating account mentioned in modern literature. It is considered appropriate, therefore, to start this sequence of case histories by quoting some selected sections of the full original account (see Eckhoff, 2003). The wheat flour explosion in Mr. Giacomelli's bakery in Turin was a comparatively minor one, but there is still much to learn from Count Morozzo's analysis. The considerations related to the low moisture content of the flour due to dry weather are still highly relevant. The same applies to the observation of a primary explosion causing a secondary explosion by entrainment of dust deposits.

> On the 14th of December, 1785, about six o'clock in the evening, there took place in the house of Mr. Giacomelli, baker in this city, an explosion which threw down the windows and window-frames of his shop, which looked into the street; the noise was as loud as that of a large cracker, and was heard at a considerable distance. At the moment of the explosion, a very bright flame, which lasted only a few seconds, was seen in the shop; and it was immediately observed, that the inflammation proceeded from the flour warehouse, which was situated over the back shop, and where a boy was employed in stirring some flour by the light of a lamp. The boy had his face and arms scorched by the explosion; his hair was burned, and it was more than a fortnight before his burns were healed. He was not the only victim of this event; another boy, who happened to be upon a scaffold, in a little room on the other side of the warehouse, seeing the flame, which had made its passage that way, and

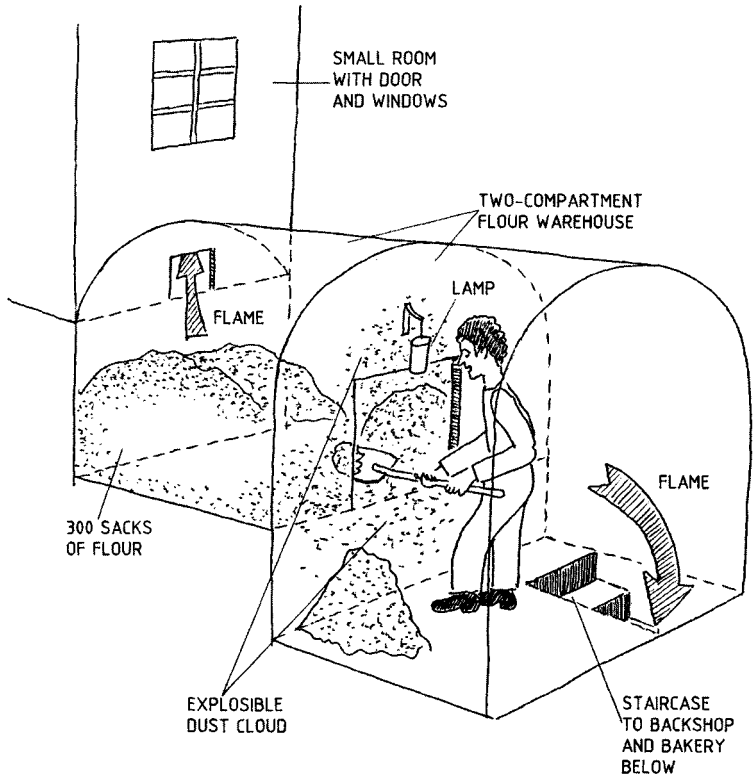

Figure 5–41 Reconstruction of possible scene of wheat flour explosion in Mr. Giacomelli's bakery in Turin, Italy, on 14th December, 1785, as described by Count Morozzo (1795). (From Eckhoff, 2003).

thinking the house was on fire, jumped down from the scaffold, and broke his leg.

The flour warehouse, which is situated above the back shop, is six feet high, six feet wide, and about eight feet long. It is divided into two parts, by a wall; and arched ceiling extends over both, but the pavement of one part is raised about two feet higher than that of the other. In the middle of the wall is an opening of communication, two feet and a half wide, and three feet high; through it the flour is conveyed from the upper chamber into the lower one.

The boy, who was employed in the lower chamber, in collecting flour to supply the bolter below, dug about the sides of the opening, in order to make the flour fall from the upper chamber into that in which he was; and, as he was digging, rather deeply, a sudden fall of a great quantity

took place, followed by a thick cloud, which immediately caught fire, from the lamp hanging to the wall, and caused the violent explosion here treated of.

The flame pushed itself in two directions; it penetrated, by a little opening, from the upper chamber of the warehouse, into a very small room above it, where, the door and window frames being well closed and very strong, it produced no explosion; here the poor boy, already mentioned, broke his leg. The greatest inflammation, on the contrary, took place in the smaller chamber, and, taking the direction of a small staircase, which leads into the back shop, caused a violent explosion, which threw down the frames of the windows which looked into the street. The baker himself, who happened then to be in his shop, saw the room all on fire some moments before he felt the shock of the explosion.

The baker told me that he had never had flour so dry as in that year (1785), during which the weather had been remarkably dry, there having been no rain in Piedmont for the space of five or six months: indeed, he attributed the accident which had happened in his warehouse to the extraordinary dryness of the corn.

The phenomenon, however, striking at the time it happened, was not entirely new to the baker, who told me that he had, when he was a boy, witnessed a similar inflammation; it took place in a flour warehouse, where they were pouring flour through a long wooden trough, into a bolter, while there was a light on one side; but, in this case, the inflammation was not followed by an explosion.

5.4.3 Three Grain Dust Explosions in Norway (1970-1988)

5.4.3.1 Wheat Grain Dust, Stavanger Port Silo, June 1970

The explosion occurred in Norway's largest and newly built import grain silo in Stavanger on a hot and dry summer day. Fortunately, no persons were killed, but some workers suffered first degree burns. Although the extent of flame propagation was considerable, the material damage was moderate due to the comparatively strong reinforced concrete structure of the buildings and the venting through existing openings.

The entire event lasted for a period of about 25–30 seconds, during which a sequence of six or seven distinct explosions were heard. In the middle of

this sequence, there was an interval of 10–12 seconds. The flame propagated a total distance of about 1,500 meters, through a number of bucket-elevators, horizontal conveyors, ducting, filters and rooms in the building. Dust explosions occurred in six of the large, cylindrical storage silos of 2000 m^3 volume each, in one slightly smaller silo, in seven of the slimmer, intermediate silos of capacities 400 or 51000 m^3, in one 150 m^3 silo, and in seven loading-out silos of capacities 50 m^3 each. The six largest silos had no venting, whereas the explosions in the single, slightly smaller silo, and in all the intermediate and loading-out silos, were vented through 0.4 m^2 manholes, which had their covers flung open. It is interesting to note that only one silo was damaged in this incident, namely one of the six unvented, large storage silos, which had its roof blown up, as shown in Figure 5–42. It is thus clear that the maximum explosion pressures in all the other twenty-one silos, vented and unvented, were lower than about 0.2 bar(g), which would be required to blow up the actual type of silo roof.

Almost all the windows, except those in the office department, were blown out, as was a large provisional light wall at the top of the head house. The legs of all of the five bucket elevators of 0.65 m × 0.44 m cross-section were torn open from bottom to top. The dust extraction ducts were also in part torn open.

The ignition source and its location were never fully identified. However, two hypotheses were put forward. The first was self-ignition of dust deposits in the boot of a bucket elevator in which the explosion was supposed to have started. The self-ignition process was initiated by a bucket that had been heated by repeated impacts until it finally loosened and fell into the dust deposit in the elevator boot. The second hypothesis was that the chain of events leading to ignition started with welding on the outside of the grain feeding duct leading to one of the elevator boots. The situation is illustrated in Figure 4–8. Due to efficient heat transfer through the duct wall, self-heating could have been initiated in a possible dust deposit on the inside of the duct wall. Lumps of the smoldering deposit could then have loosened and subsequently become conveyed into the elevator boot, initiating an explosion in the dust cloud there.

Figure 5–42 Damaged silo roof after the wheat grain dust explosion in Stavanger in June 1970. From Eckhoff (2003).

5.4.3.2 Wheat Grain Dust in New Part of Stavanger Port Silo, October 1988

The explosion was described by Olsen (1989). Because of effective explosion mitigation by venting and automatic explosion suppression, the extent of and damage caused by the explosion were minor. There were neither fatalities nor injuries. The incident deserves attention, however, because the chain of events leading to explosion initiation was identified, and because the incident illustrates that properly designed measures for explosion mitigation are effective.

The explosion occurred in a bucket elevator head immediately after termination of transfer of Norwegian wheat grain between two silo cells. At the moment of explosion the transport system was free of grain. In this new part of Stavanger Port Silo, the bucket elevator legs are cylindrical and mounted outdoors, along the wall of the head house. A number of vents are located along the length of the legs. In the explosion incident the vent covers on the elevator leg involved were blown out, which undoubtedly contributed to reducing the extent of the explosion. There was no significant material damage, either by pressure or by heat. Figure 5–43 illustrates the head of the bucket elevator in which the explosion occurred.

Because of a slight offset, the steel cover plate for the felt dust seal for the pulley shaft touched the shaft and became heated by friction during operation of the elevator. The hot steel plate in turn ignited the felt seal, from which one or more glowing fragments dropped into the wheat grain dust deposit on the inclined surface below, initiating smoldering combustion in the deposit. Just after the elevator had stopped, there was presumably still enough dust in the air to be ignitable by the smoldering dust, and to be able to propagate a flame. Alternatively, some of the smoldering dust may have slid down the inclined surface and become dispersed and transformed into an exploding dust cloud. Just after the explosion, some smoldering dust was still left on the inclined plate below the elevator pulley.

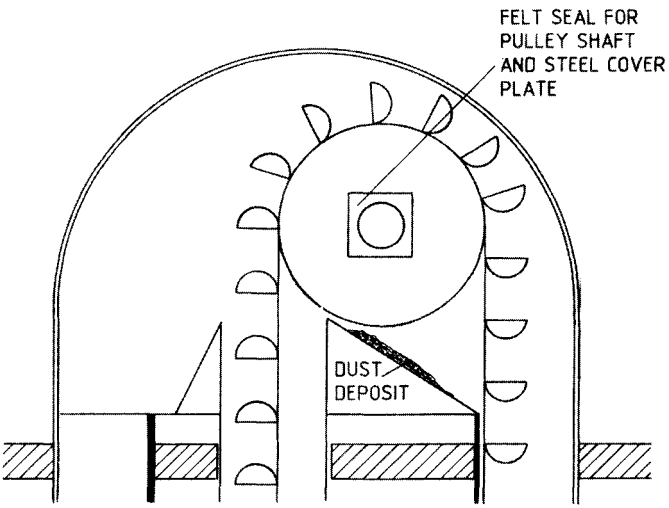

Figure 5–43 Schematic illustration of head of a bucket elevator in the new part of Stavanger Port Silo, where the minor 1988 wheat grain dust explosion was initiated. Courtesy of O. Olsen, Stavanger Port Silo, Norway.

5.4.3.3 Barley/Oats Dust Explosion in Head House of Silo Plant at Kambo, June (1976)

This explosion, described by Storli (1976), caused considerable material damage, but due to fortunate circumstances there were neither fatalities nor significant injuries. The dust involved was from Norwegian barley or oats.

The explosion probably started in a bucket elevator, initiated by burning/glowing material from an overheated hammer mill. The primary explosion developed into a secondary explosion in the head house itself, which pushed out most of its front wall, as shown in Figure 5–44.

Two of the bucket elevators had bulged out along the entire length and the dust extraction ducting had become torn apart, and this gave rise to the secondary explosion. Because the floors were supported by the wall, and the connections between wall and floors were weak, the entire wall sheet was pushed out at a quite low explosion pressure, leaving the floors unsupported at the front.

After the explosion, the head house was reconstructed. The floors were supported by a rigid framework, and should an explosion occur again, the lightweight wall elements can serve as vent covers, without weakening the support of the floors.

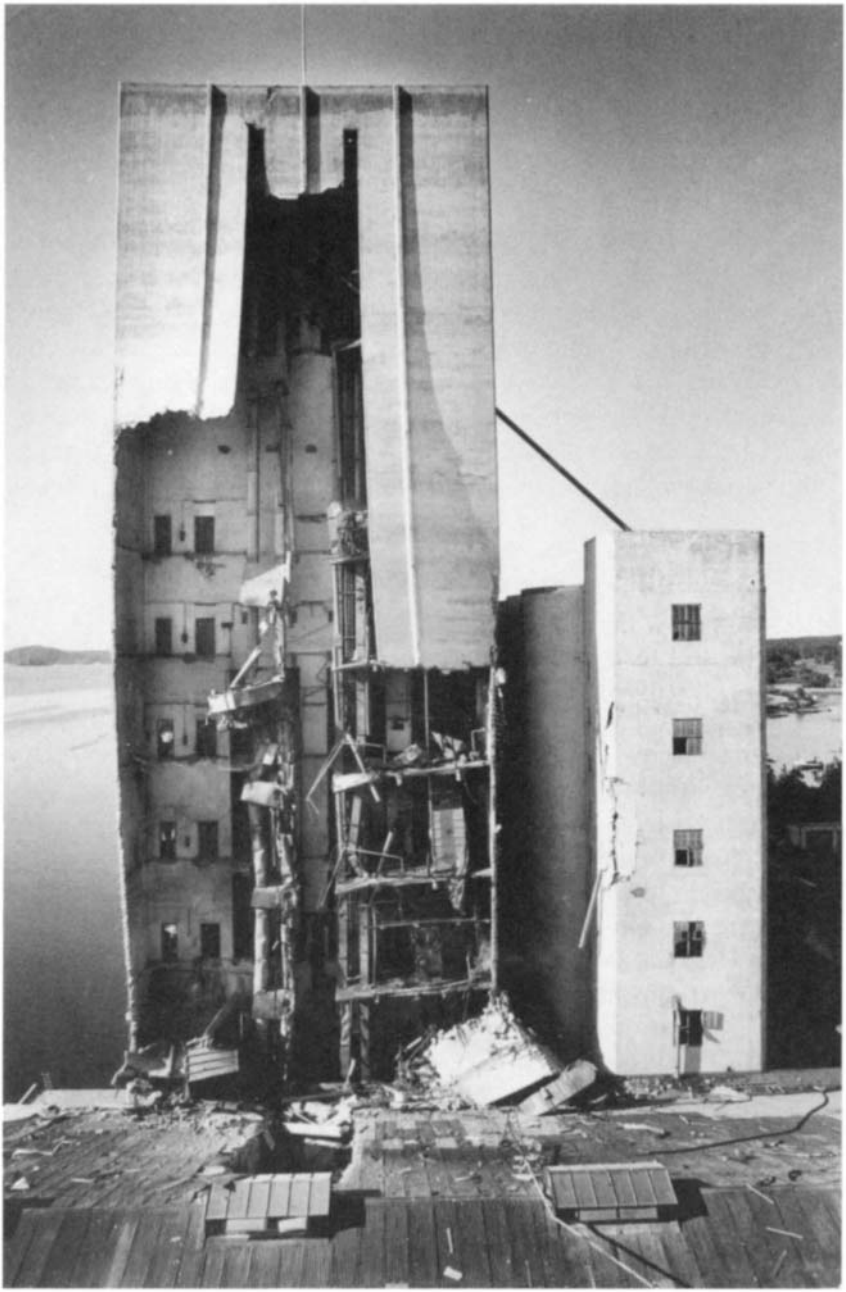

Figure 5–44 Damaged silo head house after a grain dust explosion at Kambo, Norway, in June 1976. Courtesy of Scan Foto, Oslo, Norway.

5.4.4 Major Linen Dust Explosion in Harbin, China (1987)

5.4.4.1 General Outline

In the middle of the night, at 0239 on 15 March, 1987, the spinning section of the large linen textile plant in Harbin, P.R. China, was afflicted with a catastrophic dust explosion. The losses were substantial. Out of the 327 women and men working night shift in the spinning section when the explosion occurred, fifty-eight lost their lives and 177 were injured. 13,000 m^2 of factory area was demolished. The explosion accident has been discussed in detail by Xu Bowen (1988) and Zhu Hailin (1988). Xu Bowen et al. (1988) reconstructed a possible course of the explosion development on the basis of a seismic recording of the explosion by the State Station of Seismology, located only 17 km. from the linen textile plant.

5.4.4.2 Explosion Initiation and Development, Scenario 1

Figure 5–45 illustrates the 13,000 m^2 spinning section through which the explosion swept, and the possible locations and sequence of the nine successive explosions that comprised the event according to Xu Bowen (1988) and Xu Bowen et al. (1988). These workers based their reconstruction of the explosion on three independent elements of evidence. First, they identified the location of the various explosion sites throughout the damaged plant. Secondly, they ranked the relative strengths of the local explosions by studying the extent and nature of the damage. Thirdly, they arranged the various local explosions in time by means of the relative strengths of the nine successive explosions, identified by decoding the seismic recording of the event.

Figure 5–46(a) shows a direct tracing of the amplitude-modulated seismic signal actually recorded 17 km from the explosion site. Figure 5–46(b) shows the sequence of nine energy pulses impacting on the earth at the location of Harbin Linen Textile Plant, deduced from the signal in (a). Finally, Figure 5–46(c) shows the theoretical prediction of the seismic signal to be expected from the sequence of explosions in Figure 5–46(b). The agreement between the (a) and (c) signals is striking, which supports the validity of the postulated energy impact pulse train (b).

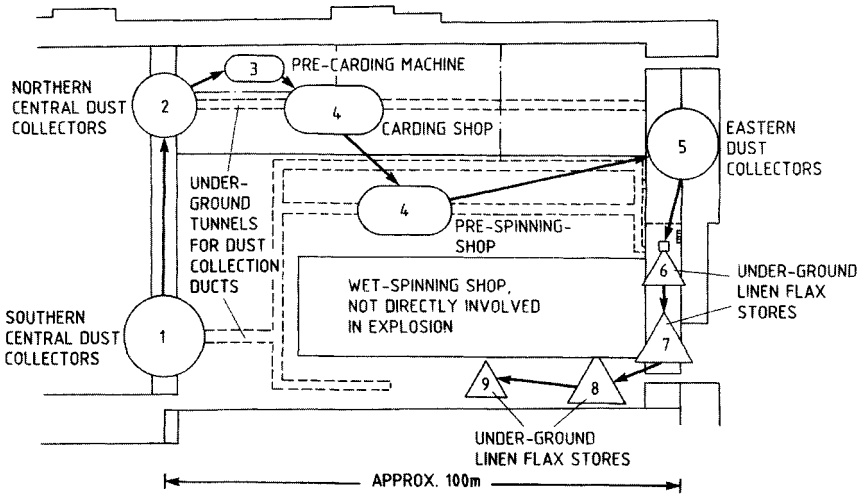

Figure 5–45 Schematic illustration of the 13, 000 m^2 spinning section of the Harbin Linen Textile Plant, P.R. China, that was afflicted with a catastrophic dust explosion on 15 March 1987. Numbered circles, ovals and triangles indicate location and sequence of a postulated series of nine successive explosions. From Eckhoff (2002).

According to Xu Bowen et al. (1988), the explosion was initiated in one of the nine units in the central dust collector system. All nine units were connected by ducting. The ignition sources were not identified, but an electrostatic spark was considered as one possibility, a local fire or glow as another. The initial flame was transmitted immediately to the next dust collecting unit, and both units (1) indicated in Figure 5–45 exploded almost simultaneously, giving rise to the first major impact pulse in Figure 5–46(b). The explosion then propagated through the other seven dust collecting units in the central collecting plant (2) indicated in Figure 5–45, and into the pre-carding area. Here the blast wave preceding the flame had generated an explosive dust cloud in the room, which was ignited by the flame jet from the dust collectors (3). The room explosion propagated further to the carding and pre-spinning shops (4), and right up to the eastern dust collectors, where another distinct explosion (5) occurred. The final four explosion pulses were generated as the explosion propagated further into the underground linen flax stores, where it finally terminated after having traveled a total distance of about 300 m. The chain of nine explosions lasted for about eight seconds.

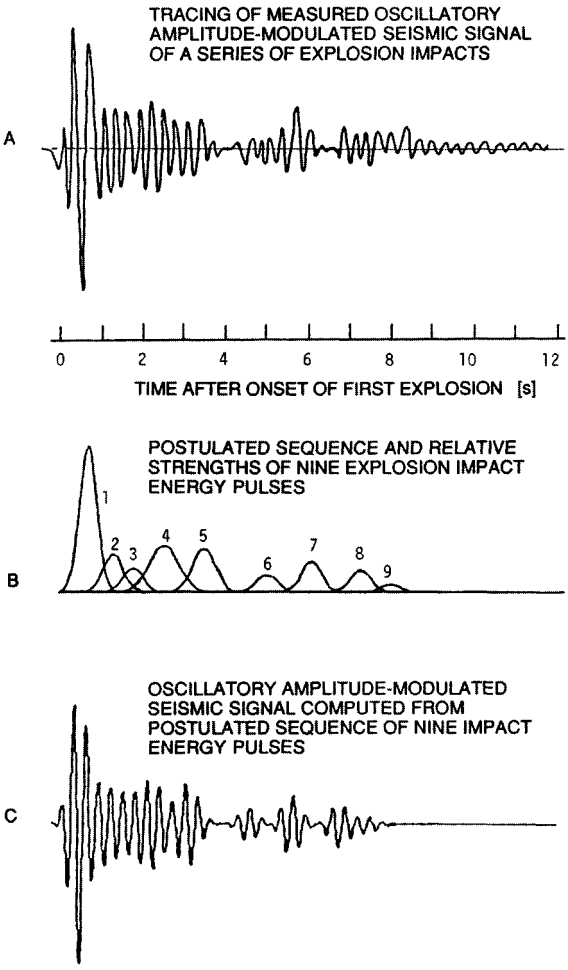

Figure 5–46 Sequence of nine impact energy pulses from nine successive dust explosions in the Harbin Linen Textile Plant, Harbin, P.R. China, 15 March 1987, postulated on the basis of a seismic record of the event. From Eckhoff (2000).

5.4.4.3 Explosion Initiation and Development, Scenario 2

This alternative scenario originates from the investigation of Zhu Hailin (1988), who found evidence of an initial smoldering dust fire caused by a live 40 W electrical portable light lamp lying in a flax dust layer of 6–8 cm thickness in a ventilation room. He also found evidence of flame propagation through the underground tunnels for the dust collection

ducting. On the basis of his analysis, Zhu suggested that the explosion was initiated in the eastern dust collectors (5) indicated in Figure 5–45, from which it transmitted to the nine units of the central dust collecting plant (1) and (2) via the ducting in the underground tunnels. Severe room explosions were initiated when the ducting in the tunnel ruptured, and the resulting blast dispersed large quantities of dust in the workrooms into explosive clouds that were subsequently ignited. From the eastern dust collectors the explosion also propagated into the underground flax stores.

5.4.4.4 Final Remark

The investigation of the Harbin disaster exposed the great difficulties in identifying the exact course of events of major explosions creating massive damage. In addition to causing pain and grief, loss of life also means loss of eyewitnesses. Besides, the immediate need for fire fighting and rescue operations changes the scene before the investigators can make their observations. Also, the explosion itself often erases evidence, e.g. of the ignition source. However, the Harbin disaster unequivocally demonstrated the possible dramatic consequences of inadequate housekeeping in industrial plants where fine combustible dust is generated.

5.4.5 Major Silicon Dust Explosion in Bremanger, Norway (1972)

In this serious explosion accident, five workers lost their lives and four were severely injured. The explosion 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. Figure 5–47 shows a view of the extensive damage. Eyewitnesses reported that the flame was very bright, almost white. This is in accordance with the fact that the temperature of silicon dust flames, as of flames of aluminum and magnesium dust, is very high due to the large amounts of heat released in the combustion process per mole of oxygen consumed. Because of the high temperature, the thermal radiation from the flame is intense, which was a main reason for the very severe burns that nine of the workers suffered.

The investigation after the accident disclosed a small hole in a steel pipe for conveying silicon powder from one of the mechanical sieves to a silo below. An oxygen/acetylene cutting torch with both valves open was

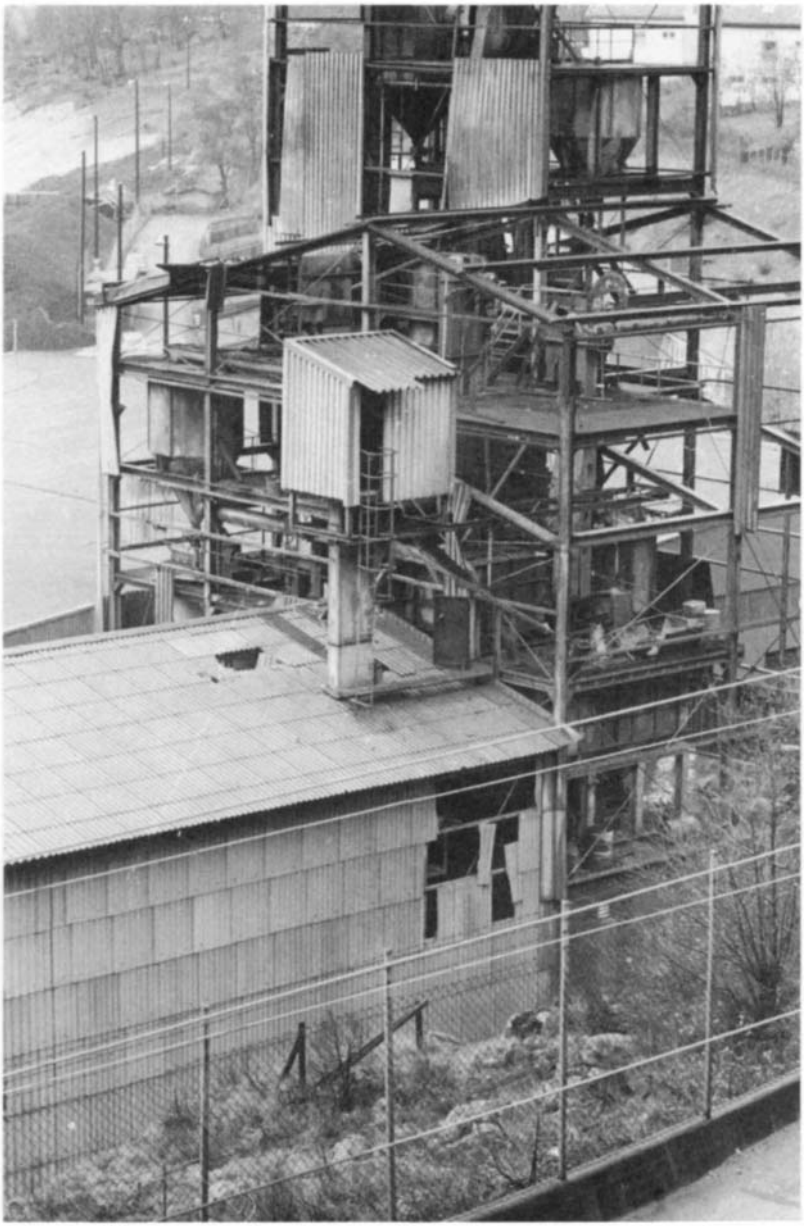

Figure 5–47 View of the extensive demolition of the silicon grinding plant caused by the silicon dust explosion at Bremanger, Norway, October 1972.

found lying on the floor about 1 m from the pipe with the hole. According to Kjerpeseth (1990) there was strong evidence of the small hole having been made by means of the cutting torch just at the time when the explosion occurred. The interior of the pipe that was perforated had probably not been cleaned prior to the perforation. At the moment of the explosion, part of the plant was closed down due to various repair work. However, the dust extraction system was operating. In view of the high temperature and excessive thermal power of the cutting torch, and not least the fact that it supplied pure oxygen to the working zone, a layer of fine dust on the internal pipe wall may well have become dispersed and ignited as soon as the torch had burned its way through the pipe wall. The blast from the resulting primary silicon dust explosion then raised dust deposits in other parts of the plant into suspension and allowed the explosion to propagate further until it eventually involved the entire silicon grinding building. The grinding plant was not rebuilt after the explosion.

5.4.6 Major Aluminum Dust Explosion at Gullhaug, Norway (1973)

The main source of information concerning the original investigation of the accident is Berg (1989). The explosion occurred during the working hours, just before lunch, while ten workers were in the same building. Five of these lost their lives, two were seriously injured, two suffered minor injuries, whereas only one escaped unhurt. A substantial part of the plant was totally demolished, as illustrated by Figure 5–48.

The premix preparation plant building was completely destroyed. Debris was found up to 75 m from the explosion site. The explosion was followed by a violent fire in the powders left in the ruins of the plant and in adjacent storehouse for raw materials. The explosion occurred when charging the 5.2 m^3 batch mixer, illustrated in Figure 5–49.

About 200 kg of very fine aluminum flake, sulphur, and some other ingredients had been charged at the moment of the explosion. The total normal charge of the formulation in question was 1,200 kg. The upper part of the closed vertical mixing vessel was cylindrical, and the lower part conical. The feed chute was at the bottom. The internal mixing device consisted of a vertical rubber-lined screw surrounded by a rubber-lined earthed steel tube. The powders to be mixed were transported upwards by the screw, and when emerging from the top outlet of the tube, they dropped to the surface of the powder heap in the lower part of the vessel. There they

Figure 5–48 Scene of total demolition after aluminum dust explosion in the premix plant of a slurry explosives factory at Gullaug, Norway, in August 1973. From Eckhoff (2003).

became mixed with other powder elements and eventually re-transported to the top.

The construction materials of the mixer had been selected so as to eliminate the formation of mechanical sparks. This was probably why both the screw and the internal wall of the surrounding earthed steel tube were lined with rubber.

During operation the 5.2 m^3 vessel was flushed with nitrogen, the concentration of oxygen in the vessel being controlled by a direct-reading oxygen analyzer at the vessel top. According to the foreman's statement, the oxygen content at the moment of explosion was within the specified limit. After the explosion, the heavy central screw part of the mixer, with the mixer end cap, was retrieved about 12 m away from the location of the mixer prior to the explosion. More detailed investigation of the part of the screw that was shielded by the steel tube, revealed that the screw wings had been deformed bi-directionally as if an explosion in the central part had expanded violently both upwards and downwards. This evidence was considered as a strong indication of the explosion having in fact been initiated inside the steel tube surrounding the screw. The blast and

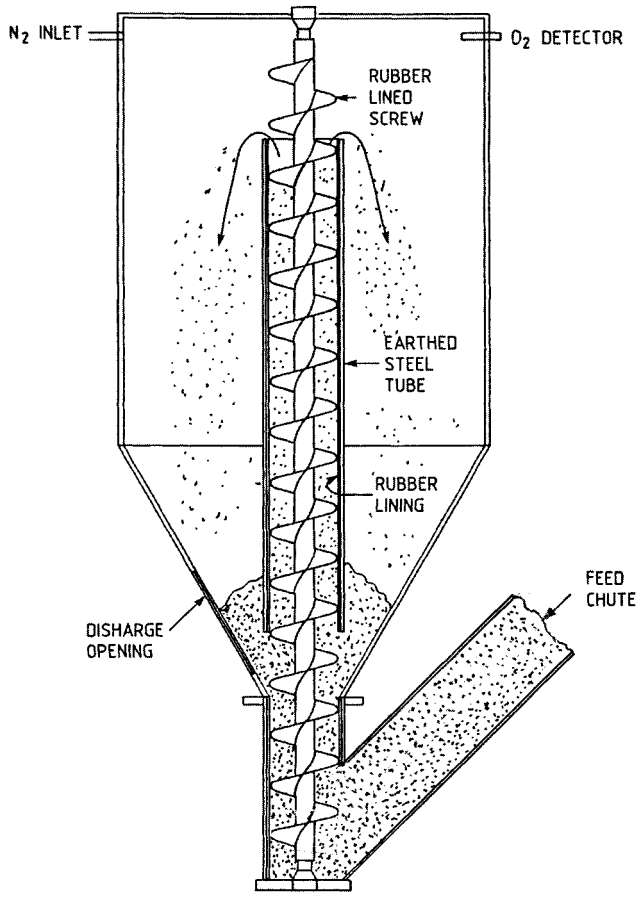

Figure 5–49 Cross-section of the mixer used for production of dry premix for slurry explosives at Gullaug, Norway, in 1973, From Eckhoff (2003).

flame from this primary explosion, in turn, generated and ignited the dust cloud in the main space inside the mixer. Finally the main bulk of the powder in the mixer was thrown into suspension and brought to ignition when the mixer ruptured, giving rise to a major dust explosion in the workrooms.

Subsequent investigations revealed that clouds in air of the fine aluminum flake powder involved were both extremely sensitive to ignition and exploded extremely violently. The minimum electric spark ignition

energy was of the order of 1 mJ, and the maximum rate of pressure rise in the Hartmann bomb 2,600 bar/s. Both these values are extreme. The thickness of the aluminum flakes was about 0.1 μm, which corresponds to a specific surface area of about 7.5 m^2/g.

The investigation further disclosed that the design of the nitrogen inerting system of the mixer was inadequate. First, the nitrogen flow was insufficient to enable reduction of the average oxygen concentration to the specified maximum level of 10 vol.% within the time allocated. Secondly, even if the flow had been adequate, both the nitrogen inlet and the oxygen concentration probe were located in the upper part of the vessel (see Figure 5–49) which rendered the measured oxygen concentration unreliable as an indicator of the general oxygen level in the mixer. It is highly probable that the oxygen concentration in the lower part of the mixer, and in particular in the space inside the tube surrounding the screw, was considerably higher than the measured value. This explains why a dust explosion could occur in spite of low measured oxygen concentration.

The final central concern of the investigators was identification of the probable ignition source. In the reports from 1973, it was concluded that the primary explosion in the tube surrounding the screw was probably initiated by an electrostatic discharge. However, this conclusion was not qualified in any detail. In more recent years the knowledge about various kinds of electrostatic discharges has increased considerably. It now seems highly probable that the ignition source in the 1973 Gullaug explosion was a propagating brush discharge, brought about by the high charge density that could accumulate on the internal rubber lining of the steel screw and of the steel tube surrounding the screw, because of the earthed electrically conducting backing provided by the screw and the tube.

5.4.7 Major Polyethylene Dust Explosion, Kinston, North Carolina, U.S.A. (2003)

5.4.7.1 Introduction

On 29 January 2003, a dust explosion occurred at the West Pharmaceutical Services, Inc. plant in Kinston, North Carolina, U.S.A. Six workers

lost their lives and thirty-eight were injured, including two fire fighters. Because of the number of deaths and injuries, the U.S. Chemical Safety and Hazard Investigation Board (CSB) launched an investigation to determine the root and contributing causes of the explosion and to make recommendations to prevent similar occurrences The present account is based on the comprehensive CSB (2003) report.

5.4.7.2 Company and Process

Founded in 1923, West is one of the world's largest manufacturers of closures and components for sealing drug vials and pre-filled syringes. The headquarters are in Lionville, Pennsylvania, U.S.A. The company has approximately 4,000 employees working in eight facilities in North America and ten in Europe and Asia.

The West Kinston plant had manufactured rubber drug delivery components for syringe plungers, septums, and vial seals since 1975. The rubber compounding process in use at the time of the dust explosion was started up in 1987, following a major expansion and automation project. This process was basically similar to other rubber manufacturing processes, such as tire production. Production operations included rubber compounding, molding, and extrusion. Raw materials were prepared in another area of the plant. The production was semi-continuous, producing sequential batches and operating twenty-four hours per day, five or six days per week. At the time of the explosion, 264 employees and thirty-five full time contract workers were employed at the Kinston plant.

5.4.7.3 Location and Layout of the Kinston Plant

The Kinston plant was located in a light industrial business park adjacent to the regional airport. Two private residences and the local Humane Society shelter were each located about 300 m from the facility. The plant area illustrated in Figure 5–50 was approximately 13500 m^2, and primarily single story. However; some of the rubber compounding equipment was located in a 18 m high, three storey area, which is hatched in Figure 5–50. The plant housed two main operations, viz. rubber compounding and product finishing. In the finishing process, the compounded rubber was molded and pressed into stoppers and plungers.

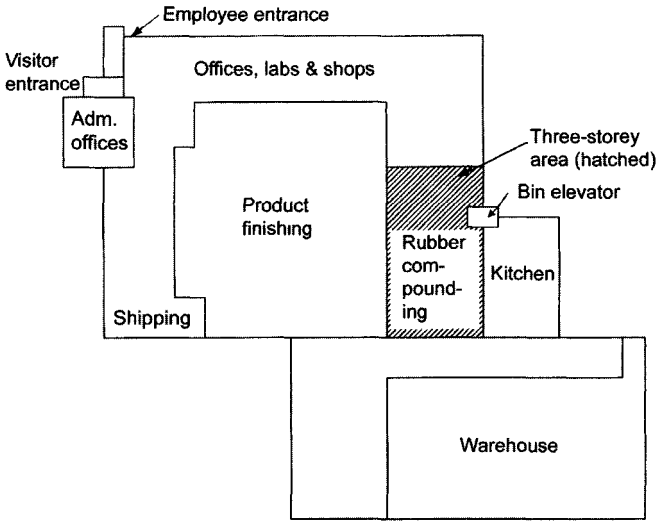

Figure 5–50 Plan of the West Pharmaceutical Services Inc., production plant in Kinston, North Carolina, U.S.A., prior to the dust explosion 29 January, 2003. From CSB (2003).

5.4.7.4 Rubber Compounding Process

5.4.7.4.1 Overview

The location of the compounding area in relation to the other facilities, is given in Figure 5–50. The automated rubber compounding system consisted of two separate production lines, each with a mixer, a roller, and a "batchoff" machine (see Figure 5–52). The compounding was done in batches, and the purpose of the batchoff machine was to cool, coat, and fold the strips of rubber from a compounded batch. The batchoff machines used were of a common design utilized in various industries.

5.4.7.4.2 Raw Material Preparation

The raw material preparation for the rubber compounding process was taking place in the "kitchen." This was a process area to the side of the

rubber compounding area and separated from it by a concrete masonry firewall. Solid materials were weighed and loaded into bins. A roller conveyer transported the bins to an elevator, where they were lifted to the second floor of the compounding structure.

5.4.7.4.3 Mixing/Kneading

A conveyor carried the bins further to the mixers, where the ingredients were compounded. Ingredients were generally loaded into the mixers through an open hatch on the side. However, bulk powders used in large portions, such as calcined clay, were pneumatically transferred to weighing hoppers and automatically dropped into the mixers. Each mixer had two opposing rotors that meshed, pulled, and sheared the components to create a uniform mix. The frictional heat generated by the mixer facilitated this process. Mineral oil was used as a plasticizer for the rubber blends and was piped directly to the mixers. After the batch was loaded, the operator closed the feed door and engaged the mixer with automatic controls.

The kneading action of the mixer caused frictional heating of the rubber. Although chilled cooling water flowed through the kneading rotors, the varying speed of the rotors and the duration of the mixing phase largely controlled the temperature of the rubber. The process temperature was held by automatic controls to below the onset temperature for vulcanization. (The rubber was vulcanized later during the forming process, when the finished products were shaped by molding).

5.4.7.4.4 Rolling, Trimming, Cooling, and Drying

Once compounded in the mixers, the rubber was dropped through chutes to the ground floor, where rollers smoothed it into sheets of roughly uniform thickness. The sheets were then cut into strips, which entered the batchoff machine, where they were cooled, coated, and folded. Then the strips were trimmed, i.e. they were dipped into a tank containing a slurry of very fine polyethylene powder in water. The polyethylene powder acted as an anti-tack agent and had an average particle size of 12 μm, i.e. it was very fine. The water slurry also cooled the product to prevent premature vulcanization.

5.4.7.4.5 Air Drying of Trimmed Strips

After leaving the dip tank, the rubber passed in front of a series of air fans. The fans drew air from the room and blew it across the rubber strips to enhance drying. At the exit of the batchoff, the rubber was folded. Finally the dried strips were stacked for shipment, or for molding in the finishing area of the plant. After drying almost all of the polyethylene powder in the slurry coating adhered to the strips, but a small amount became airborne.

5.4.7.5 Housekeeping Standards

5.4.7.5.1 General

The plant management knew that the compounding process could create dusty conditions. Therefore, local exhaust ventilation (LEV) ducts had been installed at the compound mixers and in certain areas of the kitchen, primarily to limit employee exposure to airborne nuisance dusts. The LEV ducts transported the captured dust to collectors located outdoors. Efforts were also made to prevent dust accumulation in work areas by a having a continuous house keeping program. A cleaning staff worked around the clock vacuuming and wiping up dust to minimize visible accumulation on exposed surfaces. Because the plant manufactured products for pharmaceutical use, keeping the facility free of dust was given high priority.

5.4.7.5.2 Batchoff Machines

The batchoff machines were sources of fugitive emissions of combustible dust. Twelve fans blew air across the rubber strip to cool and dry it as it passed through each machine. Some portion of the anti tack agent was carried by air currents from the machine into the room, where it tended to settle on surfaces. The cleaning crew continuously wiped and vacuumed the dust from surfaces so that the area was generally free of visible accumulation. However, there was no organized cleaning program for surfaces of beams, conduits, and other features above the ceiling, where dust was known to accumulate due to the design of the dust extraction system of the machines. Partition walls partially enclosed

the batchoff machines to separate them from other areas. Regular housekeeping was conducted around the machines, and de-humidifiers/filters associated with the enclosures removed some dust from the air. Witness statements and photographs submitted by the Kinston plant indicated that visible accumulation of dust in the milling area, even around the batchoff machines, was minimal. Management focused on the extent and effectiveness of housekeeping in working areas, and the effort was a matter of facility pride.

5.4.7.5.3 Hidden Areas not Covered by the Housekeeping Program

However, although the cleaning crew continuously cleaned the areas around the equipment, several employees told CSB investigators that there was a layer of dust on top of the suspended ceiling, above the room where the rolling mills and batchoff machines were located. Accumulation was reported to be widespread but heaviest in the areas directly above these machines. Accounts on the thickness of dust layers varied. Some employees claimed that dust accumulations of 6 mm were common, but other witnesses described heavier accumulations, such as 13 mm or more. One individual who had performed a maintenance job above the ceiling in the months prior to the incident recalled seeing as much as 50 mm of powder in some areas. Another person, who had been above the ceiling two weeks before the explosion, estimated an accumulation of up to 13 mm across 90 percent of the ceiling area. However, the investigation by the company management concluded that the overall thickness of dust accumulations ranged from 3 to 6 mm.

The area above the ceiling also contained pneumatic conveying lines for the calcined clay and other high-volume non-combustible powders used in the mixers. Because these lines were reported by employees to have leaked on at least one occasion, it is possible that some of the dust accumulation above the ceiling was non-combustible.

5.4.7.6 Outline of the Accidental Explosion

Interviews conducted by CSB investigators indicate that the operations on the day of the explosion were as usual. No one recalled to have noticed any sights, sounds, or odors that would have indicated a problem. The explosion occurred abruptly at 1328 on 29 January. Employees throughout the

plant heard the explosion, which some of them described as sounding like "rolling thunder." After seeing the exterior side panels being blown off the second story of the compounding structure witnesses outside saw a fireball and a rising smoke cloud.

Inside the facility, employees had made different observations. Those most distant from the compounding area saw lights flickering off, and ceiling tiles and debris being blown about. Some workers saw a bright flash and felt either a pressure wave or a vacuum effect that knocked them off their feet. The entire facility was affected to some extent, though explosion damage was most severe in the rubber compounding and milling areas. Figure 5–51 shows the plant after the explosion. The elevated part of the structure in the background is what was left of the three-storey rubber compounding section (see Figure 5–50 and Figure 5–52). As can be seen, the damage was extensive.

A Kinston police officer on patrol about 1 km south of the plant noticed smoke above the tree lines that surrounded the facility. He immediately contacted his dispatch to inquire if a controlled burn was taking place at the airport. Seconds later, he observed the smoke rising into a cloud about 100 m into the air, and he reported the explosion, which was heard as far as 40 km away. He immediately proceeded to the scene and began helping victims out of the facility to safety.

5.4.7.7 Fires Following the Explosion

Fires began to develop throughout the facility. The sprinkler system designed to mitigate incipient fires within the plant was rendered inoperable from the outset of the incident because the explosion broke feeder lines to the system. Emergency responders reported hearing water freely flowing into the structure. The largest and most persistent fire, lasting for two days, developed in the warehouse (Figure 5–50 and Figure 5–52). Rubber and other raw materials were stored in the warehouse, and the thermal effect from the explosion most probably reached this area and initiated the fire. Eventually, the entire warehouse was fully engulfed in flames involving the large volume of stored baled and strip rubber. Heat from the fire caused most of the steel framing to yield and collapse. Some of the rubber continued to smolder and flare up for about a week.

Mineral oil was stored in two 28 m^3 plastic tanks located between the kitchen and the warehouse (Figure 5–50). These tanks failed, spilled their

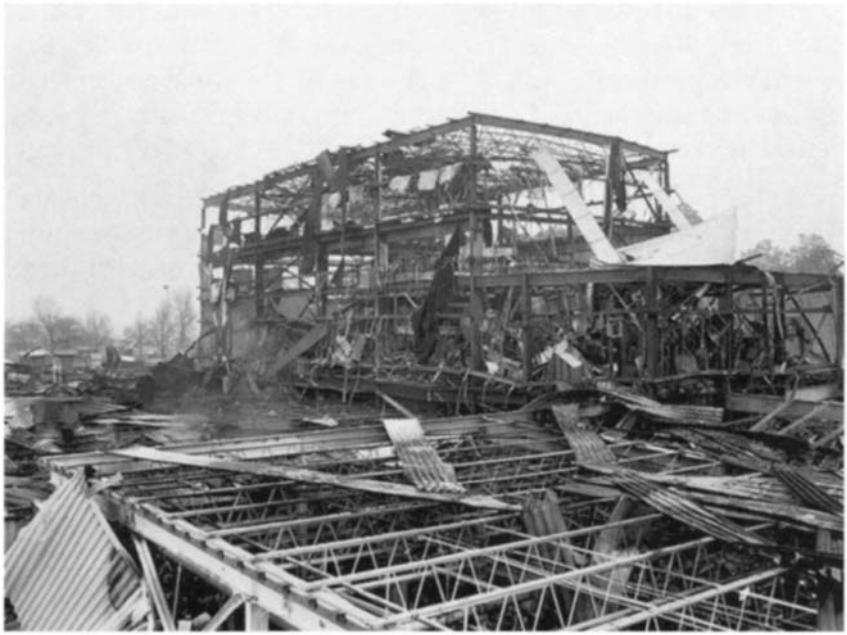

Figure 5–51 Photo of the West Pharmaceutical Services Inc., production plant in Kinston, North Carolina, USA, after the dust explosion 29 January 2003. From CSB (2003).

contents, and burned to the ground. Two additional but smaller plastic tanks containing mineral oil, located near the warehouse, also failed and contributed fuel for the fire. The concrete masonry retention walls around the tanks failed and did not prevent the burning oil from spreading.

5.4.7.8 Fatalities and Injuries

The six people who were killed were working on the ground level of the plant. Three were near one of the mills and its batchoff machine. The fourth, who died several weeks after the incident, was working at another batchoff machine. The force of the blast pushed the fifth victim east into the kitchen, and falling objects on the finishing side of the plant fatally injured the sixth person. The majority of the fatal injuries were either by thermal burns or by ejected objects or collapsing walls. When police officers entered the facility, one of the victims was trapped under a fallen girder near the end of a batchoff machine. However, because of the advancing fire, attempts at rescuing this person failed, and he died at the scene.

Immediately after the explosion, many employees were dazed or buried under debris. Responders and other employees equipped with flashlights assisted them out of the plant to triage areas. A few workers clung to the exposed frame of the building's second story and were later rescued by firefighters.

One student was injured when windows were broken at a school about 1 km away. Businesses located in the same industrial park as the Kinston plant were damaged, and windborne burning debris initiated fires in wooded areas as far as 3 km away. One home located nearby was damaged slightly, and at least two families were evacuated as a precautionary measure.

5.4.7.9 Facility Damage and Relocation of Production

The explosion and ensuing fire heavily damaged the compounding section of the Kinston facility. Figure 5–51 shows the extent of the damage. All exterior sheathing on the compounding structure was destroyed and masonry block walls were knocked down. The warehouse collapsed, and the remaining building structure was rendered mostly unusable.

Fourteen months after the explosion, the company relocated their production to an industrial facility several km south of the destroyed plant. Some equipment that was not used in the compounding process was salvaged from the original plant and is in use at the new location, and much of the workforce was rehired. When the CSB report was written, however, the destroyed facility and the compounding machinery was not in use.

The relocated production plant is not compounding rubber. Instead, rubber strips are being produced by contract manufacturers or at other West facilities and are shipped to Kinston for molding.

5.4.7.10 Analysis of Explosion

5.4.7.10.1 Fuel For Explosion

Because no other material capable of producing such a large explosion was present or used at the plant, CSB concluded that accumulated fine polyethylene dust above the suspended ceiling tiles was the main fuel in the explosion. Other possibilities were investigated, but were found to be

not credible. However, CSB were unable to conclusively determine what specific mechanism dispersed the polyethylene dust to create the primary explosive cloud.

5.4.7.10.2 Center of Explosion

CSB further concluded that the primary, strong explosion occurred in the compounding section of the plant, as illustrated in Figure 5–52. Figure 5–53 shows a photograph of the extensive damage in what was assumed by CSB to be the center of the explosion (zone of maximum blast pressure).

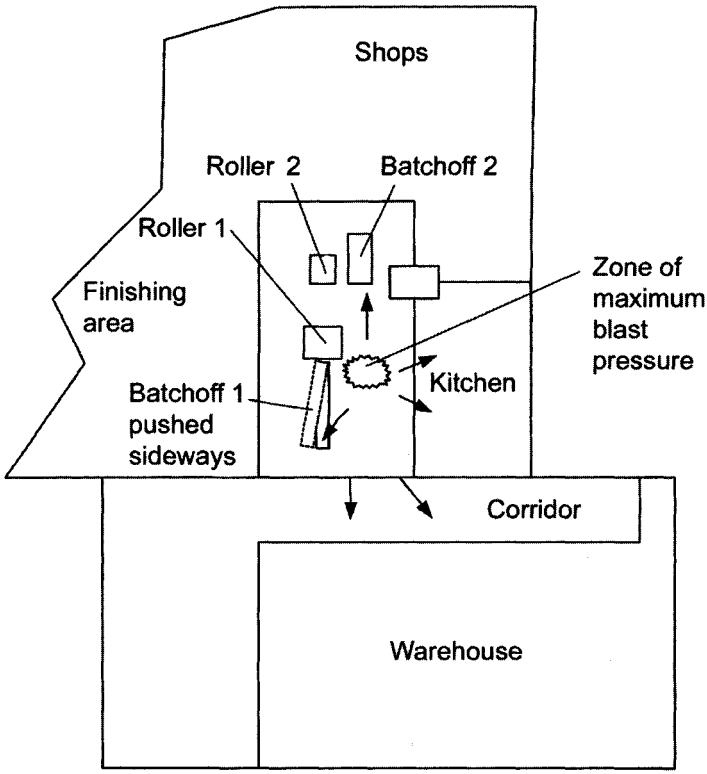

Figure 5–52 Plan of part of the West Pharmaceutical Services Inc., production plant in Kinston, North Carolina, U.S.A., showing the centre of the dust explosion 29 January 2003. From CSB (2003).

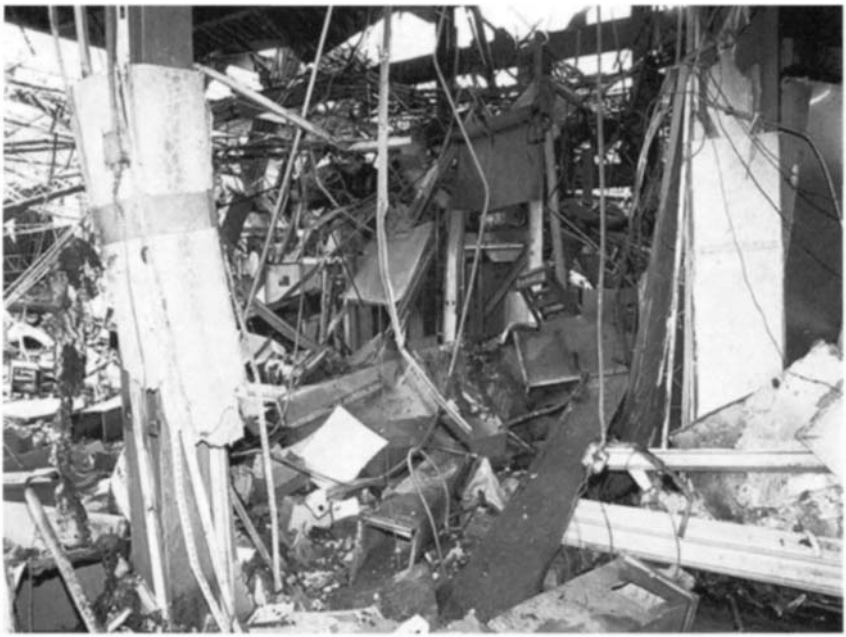

Figure 5–53 Photograph of what was assumed to be the centre of the dust explosion in the West Pharmaceutical Services Inc., production plant in Kinston, North Carolina, U.S.A., 29 January 2003. From CSB (2003).

The location of the highest pressure was determined to be close to Roller 1 on the first floor of the three-story compounding section, as indicated in Figure 5–52. Force vectors derived from observed damage to the building and surrounding equipment indicate that the largest pressure developed in the compounding section and emanated spherically outward. CSB investigators also recovered ceiling tiles debris from the suspended roof above the ground floor of the compounding section. Nearly all of the tiles appeared to be burned and splattered on top, but not on the bottom, which had faced the ground floor room below. Furthermore, some of the fluorescent light fixture pans recovered from the Roller 1 area had been flattened from above, as if they had been forcefully driven downward to the concrete floor. These two pieces of evidence give further support to the theory that the explosion occurred within the confined space above the suspended ceiling. Independent quantitative CFD simulations also concluded that the explosion occurred on the first floor of the compounding section, and that the explosion pressure from this area was the source of the extensive blast damage throughout the facility.

5.4.7.11 Initiating Events Giving Rise to the Explosive Dust Cloud and Its Ignition

CSB concluded that the accumulation of combustible dust, mainly fine polyethylene dust, above the suspended ceiling, was the most important safety issue in the West incident. The extent of damage to the Kinston facility made it extremely difficult to definitively determine the initial events that dispersed the polyethylene dust and ignited it. CSB was unable to determine whether any of the following scenarios may have been the actual initiating event:

- overheating of a batch of rubber and subsequent ignition of the vapors produced by thermal decomposition
- ignition of the dust layer by an overheated electrical ballast or light fixture
- ignition of the dust layer by an electrical spark from an unidentified electrical fault
- unsettling of dust in a cooling air duct for an electric motor and subsequent ignition of the dust by the motor

Instead the investigators focused on the most pertinent hazard, i.e. the possibility of accumulation of combustible dust in spite of a systematic housekeeping program, and considered the initiating event as a matter of secondary importance.

5.4.7.12 Previous Incident

CSB were informed that in an earlier maintenance operation involving welding, polyethylene powder in proximity to the batchoff machine had ignited, but the fire self extinguished. This incident demonstrated that the powder was ignitable. However, there was no documented investigation of this incident.

5.4.7.13 Root Causes

CSB concluded that the following root causes were responsible for the explosion and subsequent fire in the Kinston plant of West Pharmaceutical Services Inc. in 2003:

- The company did not perform an adequate safety assessment of the use of powdered zinc stearate and polyethylene as anti tack agents in the rubber batchoff process.
- The company's engineering management systems did not ensure that relevant industrial fire safety standards were consulted.
- The company's management systems for Material Safety Data Sheets (MSDS) did not identify combustible dust hazards.
- The hazard communication program at the Kinston plant did not identify combustible dust hazards or make the employees aware of such.

5.4.7.14 Some Recommendations Given by CSB to West Pharmaceuticals

- Revise policies and procedures for new material safety reviews. In particular: Use the most recent versions of MSDSs and other technical hazard information.
- Fully identify the hazardous characteristics of new materials, including relevant physical and chemical properties, to ensure that those characteristics are incorporated into safety practices, as appropriate. Include an engineering element that identifies and addresses the potential safety implications of new materials on manufacturing processes.
- Develop and implement policies and procedures for safety reviews of engineering projects. In particular: Address the hazards of individual materials and equipment and their effect on entire processes and facilities, and consider hazards during the conceptual design phase, as well as during engineering and construction phases.
- Cover all phases of the project, including engineering and construction performed by outside firms.
- Identify and consider applicable codes and standards in the design.
- Identify other production plants within the company that use combustible dusts. Ensure that they incorporate applicable recognized safety precautions. In particular: Ensure that penetrations of partitions, floors, walls, and ceilings are sealed dust tight, and ensure that spaces inaccessible to housekeeping are sealed to prevent dust accumulation there.

- Improve hazard communication programs so that the hazards of combustible dust are clearly identified and communicated to the employees. In particular: Ensure that the most current codes of practice are in use and that employees receive training on the revised/updated information.

5.5 Means of Preventing and Mitigating Dust Explosions in the Process Industries

5.6.1 Overview

Table 5–2 gives an overview of the various means that are presently known and in use. They can be divided in two main groups, namely means for preventing explosions and means for their mitigation. The preventive means can again be split in the two categories prevention of ignition sources and prevention of explosive/combustible clouds. Quite often one has to accept the occurrence of explosive dust clouds inside process equipment as an inherent feature of the process. One central issue is then whether only preventing ignition sources can give sufficient safety, or whether it is also necessary to employ additional means of explosion mitigation. The general answer is that preventing ignition sources is not sufficient. In the following sections the various means listed in Table 5–2 will be discussed separately.

Table 5–2 Overview of Means for Preventing and Mitigating Dust Explosions in the Process Industries

Prevention		Mitigation
Preventing explosive dust clouds	**Preventing ignition sources**	
Inerting by N_2, CO_2 and rare gases	Smoldering combustion in dust, dust flames	Reduce expl. cloud size
		Partial inerting
Intrinsic inerting	Other types of open flames (e.g. hot work)	Isolation (sectioning)
Inerting by adding inert dust	Hot surfaces	Venting
Dust concentration outside explosive range	Electric sparks and arcs, electrostatic discharges	Pressure resistant construction
	Heat from mechanical impact (metal sparks and hot spots)	Automatic suppression
		Good housekeeping (dust removal/cleaning)

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5.6.2 Preventing Explosive Dust Clouds

5.6.2.1 Inerting by Adding Inert Gas to the Air

For any type of combustible dust, and a given type of inert gas added to the air, there is a limiting oxygen content below which the dust cloud is unable to propagate a self-sustained flame. By keeping the oxygen content below this limit throughout the process system, dust explosions are effectively excluded. As the oxygen content in the gas is gradually reduced from that of air, the ignitability and explosivity of the dust cloud is also gradually reduced, until ultimately flame propagation becomes impossible.

Four types of inert gases are in common use for this purpose:

- nitrogen
- carbon dioxide
- water vapor
- rare gases

In some cases the special protective method called "*intrinsic inerting*" can be a good solution. This method implies that the required quantity of inert gas is produced in the plant itself, e.g. by controlled combustion in a hot-gas generator and recirculation of the gas. Such hot-gases mainly consist of nitrogen, carbon dioxide and water vapor. The residual concentration of oxygen is kept at a sufficiently low level to ensure inert conditions. However, normally such combustion gases are not clean enough for being used in connection with food and feed materials, phamarceuticals etc. In the past also *halogenated hydrocarbons* (halons) were used for inerting. However, due to the environmental problems caused by these substances, they are no longer used for protecting against explosions and fires. The choice of inert gas depends on several considerations, such as availability and cost, possible contaminating effects on products, and effectiveness. In the case of dusts of light metals, such as aluminum and magnesium, exothermic reactions with CO_2 and also in some situations with N_2 are known, and the use of rare gases may have to be considered in certain cases.

The design of gas inerting systems depends on whether the process is continuous or of the batch type, the strength of the process equipment and

type and source of inert gas. Two main principles are used for establishing the desired atmosphere in the process:

- pressure variation method
- flushing method

The pressure variation method operates either above or below atmospheric pressure. In the former case, the process equipment, initially filled with air at atmospheric pressure, is pressurized to a given overpressure by inert gas. When good mixing of air and inert gas has been obtained, the process equipment is vented to the atmosphere and the cycle repeated until a sufficiently low oxygen content has been reached. The alternative is to first evacuate the process equipment to a certain under-pressure, and fill up with inert gas to atmospheric pressure, mix, and repeat the cycle the required number of times.

The flushing method is used if the process equipment has not been designed for the significant pressure increase or vacuum demanded by the pressure variation method. There are two extreme cases of the flushing method, viz. the replacement method (plug flow) and the through-mixing method (stirred tank). In order to maintain plug flow, the flow velocity of inert gas into the system must be low (< 1 m/s) and the geometry must be favorable for avoiding mixing. In practice this is very difficult to achieve, and the stirred tank method, using high gas velocities and turbulent mixing, is normally employed. It is essential that the instantaneous through-mixing is complete over the entire volume, otherwise pockets of unacceptably high, oxygen concentrations may form.

5.6.2.2 Dust Concentration outside Explosive Range

In principle one could avoid dust explosions by running the process in such a way that explosive dust concentrations were avoided. In practice, however, this is difficult in most cases, because the dust concentration inside process equipment in normal operation often varies in unpredictable and uncontrollable ways. On the other hand, keeping the powder/dust in the settled state by avoiding generation of dust clouds should be attempted whenever possible. Good process design can significantly reduce the regions in which explosive dust concentrations occur, as well

as the frequencies of their occurrence. One example is the use of mass flow silos instead of the traditional funnel flow type.

Nevertheless there are some special situations where it may be possible to actively keep the dust concentration below the lower explosive limit. One such situation is in dust extraction ducts, another is in cabinets for electrostatic powder coating, and a third is spray dryers. One essential condition for control of dust concentration is that the concentration can be adequately measured. In-situ methods based on light attenuation or backscattering of light have been found most suitable. The use of dust control in dust extraction systems is more likely to be successful in cases where a small dust fraction is to be removed from a coarse main product, e.g. grain dust from grain, or plastic dust from pellets. By monitoring dust concentrations and controlling air flows the desired low level of dust concentration can be maintained. However, if the air velocities are too low to prevent dust deposition on the internal walls of the ducting over time, dust explosions may nevertheless be able to propagate through the ducts.

Dust entrainment and formation of explosive dust clouds by the air blast from a primary dust explosion, may also occur in mixers, conveyors, etc. containing fine dust present as layers/deposits. This means that explosion control by controlling the concentration of suspended dust is only feasible for preventing primary explosion initiation, not for preventing secondary explosions.

5.6.2.3 Adding Inert Dust

This principle is used in coal mines, by providing sufficient quantities of stone dust either as a layer on the mine gallery floor, or on shelves, etc. The blast that will always precede the flame in a dust explosion will then entrain the stone dust and coal dust simultaneously and form a mixture that is non-combustible in air, and the flame, when arriving, will become quenched. In other industries than mining, adding inert dust is seldom applicable due to contamination and other problems.

5.6.3 Preventing Ignition Sources

5.6.3.1 Introduction

A question asked frequently is whether preventing ignition sources can be relied upon as the only means of protection against dust explosions. The general consensus is that this is not possible. Relying on preventing ignition sources only is definitely inadequate if the minimum electric spark ignition energy of the dust is in the region of vapors and gases, i.e. < 10 mJ. However, for dusts of very high MIEs it may be argued that several types of process plants could be satisfactorily protected against dust explosions solely by eliminating ignition sources.

According to Scholl (1989) one may distinguish between two categories of ignition sources. Organizational ignition sources, which can be prevented by enforcing adequate working routines, include:

- smoking
- open flames
- open light (bulbs)
- welding (gas/electric)
- cutting (gas/rotating disc)
- grinding

Operational ignition sources arise within the process itself and include:

- open flames
- hot surfaces
- smoldering nests
- exothermic decomposition
- heat from mechanical impact between solid bodies (metal sparks/ hot-spots)
- electric sparks/arcs, electrostatic discharges

5.6.3.2 Preventing Self-Heating, Smoldering and Burning in Dust Deposits

The risk of self-heating in powder/dust deposits depends on the inherent properties of the material. Therefore, possible disposition to self-heat should be known for any combustible material before admitting it to storage silos or other part of an industrial plant where conditions could favor self-heating and further temperature rise to smoldering and burning.

Possible means of preventing self-heating and self-ignition in powders with a disposition to self-heat include:

- control of temperature, moisture content and other relevant powder/dust properties before admitting powder/dust to e.g. storage silos.
- adjustment of powder/dust properties to acceptable levels by cooling, drying etc. before storage, whenever required
- ensuring that hot solid bodies do not become embedded in the powder/dust mass
- continuous monitoring of temperature in powder mass at several points by thermometer chains
- rolling of bulk material from one silo to another, or within the same silo, whenever onset of self-heating is detected, or as a routine after certain periods of storage, depending on the powder/dust type
- inerting of bulk material in silo by suitable inert gas, e.g. nitrogen
- monitoring of possible development of gaseous decomposition/oxidation products, e.g. CO and methane, for early detection of self-heating

Means for preventing and controling self-heating/self-igniting and means for extinction of smoldering combustion inside large dust deposits e.g. in silos is discussed in Chapter 4.

Some synthetic organic chemicals, in particular cyclic compounds, can decompose exothermally and become ignited by a hot surface, a smoldering nest, frictional heat or other ignition source. Such decomposition does not require oxygen, and therefore inerting has no effect. Adiabatic exothermal decomposition of bulk powder at constant volume can, due to the very high powder concentration, generate much higher pressures than a dust explosion in air. Zwahlen (1989) suggested the following alternative safety measures:

- Process the hazardous powder in the wet state, as a slurry or suspension.
- If wet processing is impossible, avoid processes involving moving mechanical parts in contact with the powder that can give rise to ignition.
- Keep the processed batches of the powder as small as feasible.
- Keep strict control to prevent foreign bodies from entering the process.
- Detectors for observing early temperature and pressure rise, and sprinkler systems must be provided.
- Use of additives that suppress the decomposition tendency may be helpful in some cases.

5.6.3.3 Preventing Ignition by Open Flames/Hot Gases

- Most potential ignition sources of the open flame type can be avoided by enforcing adequate organizational procedures and routines. This in particular applies to prohibition of smoking and other use of lighters and matches, and to enforcement of strict rules for performing hot work. Hot work must not be carried out unless the entire area that can come in contact with the heat generated by the hot work, indirectly as well as directly, is free of dust, and hazardous connections through which this heat may transmit to other areas, have been blocked. It is important to note that hot work also includes disc-cutting and grinding operations. Gas cutting torches are particularly hazardous because they work with excess oxygen. This gives rise to ignition and primary explosion development where explosions in air would be unlikely (see Section 5.4.5). Factory inspectorates in most industrialized countries have issued detailed regulations for hot work in factories containing combustible powders or dusts.

In certain situations in the process industry, hot gaseous reaction products may entrain combustible dust and initiate dust explosions. Each such case has to be investigated separately and the required set of precautions tailored to serve the purpose in question.

5.6.3.4 Preventing Ignition by Hot Surfaces

Hot surfaces may occur in industrial plants both intentionally and unintentionally. The first category includes external surfaces of hot process equipment, heaters, dryers, steam pipes and electrical equipment. The equipment where hot surfaces may be generated unintentionally include engines, blowers and fans, mechanical conveyors, mills, mixers, bearings, and unprotected light bulbs. A further category of hot surfaces arises from hot work. During grinding and disc-cutting, glowing hot surfaces are often generated, in addition to the luminous spark showers typical of these operations. A hot surface may ignite an explosive dust cloud directly, or via ignition of a dust layer that subsequently ignites the dust cloud. Parts of glowing or burning dust layers may loosen and be conveyed to other parts of the process where they may initiate explosions.

The hot surface temperature various apparatuses can, if covered by a dust layer, be significantly higher than it would normally be without dust, due to thermal insulation by the dust. This both increases the ignition hazard and may cause failure of equipment due to increased working temperature. The measures taken to prevent ignition by hot surfaces must cover both layer and cloud ignition. The measures include:

- removal of all combustible dust before performing hot work
- prevention/removal of dust accumulations on hot surfaces (electrical apparatuses etc.)
- isolation or shielding of hot surfaces
- use of electrical apparatus approved for use in the presence of the combustible dust of concern
- use of equipment with minimal risk of overheating
- inspection and maintenance procedures that minimize the risk of overheating

5.6.3.5 Preventing Ignition of Dust Clouds by Smoldering Nests

Infrared radiation detection and subsequent extinction of smoldering nests and their fragments during pneumatic transport in dust extraction ducts, has proven to be an effective means of preventing fire and explosions in downstream equipment, for example dust filters. Normally the transport

velocity in the duct is known, and this allows effective extinction by precise injection of a small amount of extinguishing agent at a convenient distance just when the smoldering/burning nest or fragment passes the nozzles. Water is the most commonly used extinguishing agent, and it is applied as a fine mist. Such systems are mostly used in the wood industries, but also to some extent in the food and feed and some other industries.

5.6.3.6 Preventing Ignition by Heat from Accidental Mechanical Impacts

Mechanical impacts produce two different kinds of potential ignition sources, viz. small flying burning fragments of solid material and a pair of hot-spots where the impacting bodies touch. Sometimes, e.g. in rotating machinery, impacts may occur repeatedly at the same points on one or both of the impacting bodies, and this may give rise to hot-spots of appreciable size and temperature. The hazardous source of ignition will then be a hot surface.

With regard to single accidental impacts, research has revealed that in general the ignition hazard associated with single accidental impacts is smaller than often believed in the past. This in particular applies to dusts of natural organic materials such as grain and feedstuffs, when being exposed to accidental sparking from impacts between steel hand tools like spades or scrapers, and other steel objects or concrete. In such cases the ignition hazard is probably non-existent.

However, if more sophisticated metals are involved, such as titanium or some aluminum alloys, energetic spark showers can be generated. In the presence of rust, luminous, incendiary thermite flashes can result. Thermite flashes may also result if a rusty steel surface covered with aluminum paint or a thin smear of aluminum, is struck with a steel hammer or another hard object. However, impact of ordinary soft unalloyed aluminum on rust seldom results in thermite flashes, but just in a smear of aluminum on the rust. For a given combination of impacting materials, the incendivity of the resulting sparks or flash depend on the sliding velocity and contact pressure between the colliding bodies.

Although the risk of initiation of dust explosions by accidental single impacts is probably smaller than believed by many in the past, there are special combinations of impacting materials where the ignition hazard is real. It would in any case seem to be good engineering practice to:

- Remove foreign objects from the process stream as early as possible.
- Avoid construction and tool materials that can give incendiary metal sparks or thermite flashes (titanium, magnesium, aluminum etc.).
- Inspect process and remove cause of impact immediately in a safe way whenever special noise signals indicating accidental impact(s) in process stream are observed.

5.6.3.7 Electric Sparks and Arcs: Electrostatic Discharges

The various types of electric sparks and arcs and electrostatic discharges are described in Section 2.2.6 and Section 5.3.6. Sparks or arcs due to breakage of live circuits can occur when fuses blow, in rotating electric machinery and when live leads are accidentally broken. The main rule for minimizing the risk of dust explosions due to such sparks and arcs is to:

- Obey regulations for electrical installations and apparatuses in areas containing combustible dust. (see Chapter 7)

The electrostatic hazard is more complex and it has not always been straightforward to specify clearly defined design guidelines. However, Glor (1988), who has contributed substantially to developing a unified approach, recommends the following measures:

- Use conductive materials or materials of low dielectric strength, including coatings, (breakdown voltage across dielectric layer or wall < 4 kV) for all plant items that may accumulate very high charge densities (pneumatic transport pipes, dust deflector plates, and walls of large containers that may become charged due to ionization during gravitational compaction of powders). This prevents propagating brush discharges.
- Earth all conductive parts of equipment that may become charged. This prevents capacitive spark discharges from equipment.
- Earth personnel if powders of minimum ignition energies (MIE) < 100 mJ are handled. This prevents capacitive spark discharges from humans.
- Earth electrically conductive powders (metals etc.) by using earthed conductive equipment without non-conductive coatings. This prevents capacitive discharges from conductive powder.

- If highly insulating material (resistivity of powder in bulk > 10^{10} Ωm) in the form of coarse particles (particle diameter > 1 mm) is accumulated in large volumes in silos, containers, hoppers, etc., electrostatic discharges from the material in bulk may occur. These discharges can be hazardous when a fine combustible dust fraction of minimum ignition energy < 10–100 mJ is present simultaneously. So far, no reliable measure is known to avoid this type of discharge in all cases, but an earthed metallic rod introduced into the bulk powder will most probably drain away the charges safely. It is, however, not yet clear whether this measure will always be successful. Therefore the use of explosion venting, suppression or inerting should be considered under these circumstances.
- If highly insulating, fine powders (resistivity of powder in bulk > 10^{10} Ωm) with a minimum ignition energy < 10 mJ as determined with a low-inductance capacitive discharge circuit, is accumulated in large volumes in silos, containers, hoppers, etc., measures of explosion protection should be considered. There is no experimental evidence that fine powders without any coarse particles will generate discharges from powder heaps, but several explosions have been reported with such powders in situations where all possible ignition sources, other than electrostatics have been effectively eliminated.

If combustible powders are handled or processed in the presence of a flammable gas or vapor (hybrid mixtures), the use of electrically conductive and earthed equipment is absolutely essential. Insulating coatings on earthed metallic surfaces may be tolerated provided that the thickness is less than 2 mm, the breakdown voltage less than 4 kV at locations where high surface charge densities have to be expected, and conductive powder cannot become isolated from earth by the coating. If the powder is non-conducting (resistivity of the powder in bulk > 10^6 Ωm), measures of explosion prevention (e.g. inert gas blanketing) are strongly recommended. If the resistivity of the powder in bulk is less than 10^6 Ωm, brush discharges, which would be incendiary for flammable gases or vapors, can also be excluded.

However, experience has shown that even in the case of powders of resistivities in bulk < 10^6 Ωm it is very difficult in practice to exclude all kinds of effective ignition sources when flammable gases or vapors are present. In such cases large amounts of powders should only be handled and processed in closed systems blanketed with an inert gas.

Glor also emphasized that, due to increasing use of non-conducting construction parts in modern industrial plants, the chance of overlooking unearthed conducting items is high. Therefore the effort to ensure proper earthing of all conducting parts must be maintained, in particular in plants handling dusts of low MIE. Adequate earthing is maintained as long as the leak resistance to earth does not exceed 10^6 Ω for process equipment and 10^8 Ω for personnel. However, in practice, one should aim for considerably lower resistances to earth.

5.6.4 Mitigating Dust Explosions that are Initiated in Spite of Preventive Measures

5.6.4.1 Reducing Sizes of Explosive Dust Clouds by Good Process Design (Inherently Safe Design)

5.6.4.1.1 Minimize Volumes of Process Equipment

A general rule is that volumes of process equipment should not be larger than the volumes required by the process. Nevertheless one sometimes finds industrial plants with e.g. silos that are considerably larger than required by the process. This can either be due to inadequate design in the first place, or due to the plant being used for another purpose than originally designed for.

5.6.4.1.2 Minimize Volumes of Dust Clouds Generated at Transfer Points

Undesired dust clouds are practically always generated when powder/dust/pelletized material etc. is falling freely under gravity. Whenever possible, therefore, efforts should be made to design transfer points in such a way that the material is flowing smoothly in bulk, rather than being dispersed as a cloud. For example, by having an inclined chute at transfer points between chain or belt conveyors, dusting can be reduced considerably. Another example is the very smooth discharge of material from a silo on to a chain/belt conveyor, which can be obtained if the silo hopper is designed to produce mass flow rather than funnel flow.

5.6.4.2 Partial Inerting by Inert Gas

In Table 5–2 partial inerting, as opposed to complete inerting discussed in Section 5.6.2.1, has been included as a possible means of mitigating dust explosions. The concept, discussed by Eckhoff (2004), implies that a smaller fraction of inert gas than that required for complete inerting, is added to the air. In this way both the ignition sensitivity, the explosion violence and the maximum constant-volume explosion pressure will reduced, in some cases appreciably. This offers a new possibility for applying mitigatory measures such as explosion venting or automatic explosion suppression in situations where the explosion violence of the dust in air only is too severe to permit the use of such techniques. More research is needed to establish correlations between the oxygen content in the gas phase and various ignitability and explosibility parameters of various dusts.

5.6.4.3 Isolation (sectioning)

In Section 2.4.5.2, three main reasons are given for trying to prevent a gas explosion in one process unit from spreading to others via pipes and ducts. This also applies to dust explosions. Firstly, there is always a desire to limit the extent of the explosion as far as possible. Secondly, a dust flame propagating in a duct between two process units can give rise to violent flame jet ignition of the dust cloud in the second volume. The third main reason is pressure piling. The effect of pressure piling towards generation of very high transient explosion pressures is enhanced by flame jet ignition in the second chamber.

As for gases basically two categories of methods are used for obtaining explosion isolation, viz. passive methods activated by the propagating explosion itself, and active ones, which require a separate flame/pressure sensor system, which triggers a separately powered system for activating the isolation mechanism. For obvious reasons, the passive systems are generally preferable, as long as they function as intended and are otherwise suitable for the actual purpose.

Passive isolation systems include the concept of flame propagation interruption in ducts by providing a vented 180° bend system, as illustrated in Figure 5–54. This concept is used quite frequently to interrupt dust explosions in pipes and ducts.

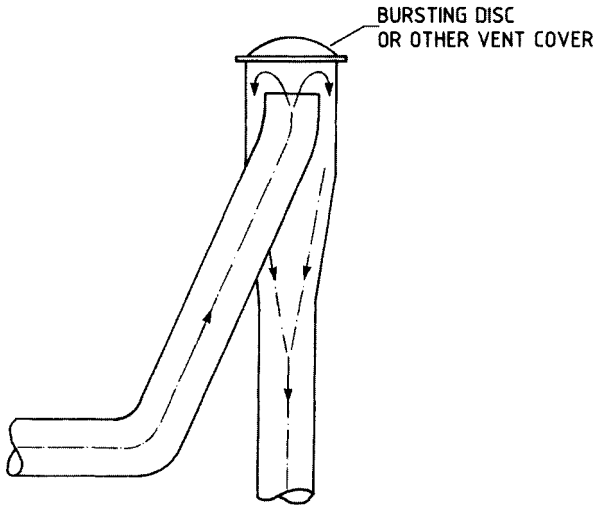


Figure 5–54 Passive device for interrupting dust explosions in pipes and ducts by combining change of flow direction and venting. Flow direction may also be opposite to that indicated by the arrows. From Eckhoff (2003).

The basic principle is that the explosion is vented at a point where the flow direction is changed by 180°. Due to the inertia of the fast flow caused by the explosion, the flow will tend to maintain its direction rather than making a 180° turn. However, the boundaries for the applicability of the principle have not been fully explored. Parameters that may influence performance include explosion properties of dusts, velocity of flame entering the device, direction of flame propagation, and direction, velocity and pressure of initial flow in duct. The use of two passive explosion interrupters of the type shown in Figure 5–54 in series in ducts between two process units, probably is a satisfactory solution in most cases. One interrupter should then be located close to each of the two process units. Screw conveyors can also be used for interrupting dust explosions. The removal of part of the screw will ensure that a plug of bulk powder/dust will always remain as a choke that will prevent transmission of a dust explosion through the screw. Specially designed rotary locks are also used for preventing explosion transfer between process units or a process unit and a duct.

Active isolation methods also include various kinds of fast-response mechanical valves. The required closing time of an automatic explosion isolation valve depends on the distance between the remote pressure or

flame sensor, and the valve, and on the type of dust. Often closing times as short as 50 ms, or even shorter, are required. This may be 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 peak pressures may be even higher than this).

Another active explosion isolation method is flame interruption by fast automatic injection of extinguishing chemicals ahead of the flame in pipes connecting process units. This is a special application of automatic explosion suppression, which will be described in Section 5.6.4.5. However, there is a possibility of the inert plug being pushed by the explosion in the pipe into the downstream process unit where its flame-stopping effect may be destroyed. Important design parameters for this type of barrier are type of dust, initial turbulence in primary explosion, duct diameter, distance from vessel where primary explosion occurs, method used for detecting onset of primary explosion, and type, quantity and rate of release of extinguishing agent.

5.6.4.4 Dust Explosion Venting

5.6.4.4.1 Main Principle

The main principle of dust explosion venting is the same as for venting of gas explosions and outlined in Section 2.4.5.6 and illustrated in Figure 2–68.

5.6.4.4.2 Sizing of Dust Explosions Vents

Several parameters have an influence on the required area for venting of dust explosions:

- enclosure volume
- length/diameter ratio of enclosure
- maximum over-pressure P_{red} that the enclosure can withstand
- static opening over-pressure P_{stat} of vent cover
- mass of vent cover
- burning rate of the dust cloud

For some time it was thought by many that the burning rate of a cloud of dust in air was a constant property of a given dust, which could be determined once and for all e.g. in the standard 1 m^3 closed vessel test (see Figure 5–18). However, a cloud in air of a given dust can burn with widely different combustion rates, depending on the dust concentration, the turbulence and the degree of dust dispersion in the actual industrial situation. This means that the required vent area also depends markedly on the specific industrial situation of dust cloud generation and flame propagation. During the last few decades, further experimental evidence in support of this fact has been produced. As a result a differentiated view on dust explosion vent sizing has gradually evolved, which has also been taken into account in the latest European Union dust explosion venting standard issued by CEN (2002a). Experimental evidence supporting a differentiated vent sizing approach is given in Chapter 6 of Eckhoff (2003). It is forseen that in the future CFD-based numerical codes will be used even for vent sizing. for simulating turbulent dust explosions in complex geometries.

5.6.4.4.3 Vent Covers

A wide range of vent cover designs are in use. One classical and simple type of vent cover is a light but rigid panel, e.g. an aluminum plate, held in position by a rubber clamping profile as used for mounting windows in cars. The profile must then remain unlocked. Other methods for keeping the vent cover in place include various types of clips. When choosing a method for securing the panel, it is important to make sure that the pressure, P_{stat}, needed to release the vent panel is small compared with the maximum tolerable explosion pressure, P_{red}. It is further important to anchor the vent panel to the enclosure to be vented, e.g. by means of a wire or a chain. Otherwise the panel may become a hazardous projectile in the event of an explosion. Finally, it is also important to make sure that rust formation or other processes do not increase the static opening pressure of the vent cover over time.

Bursting panels constitute a second type of vent covers. In the past, such panels were often "home made," and adequate data for the performance of the panels were lacking. A primary requirement is that P_{stat}, the static bursting pressure of the panel, is considerably lower than the maximum permissible explosion pressure, P_{red}. Today, high quality bursting panels are manufactured in several countries. Figure 5–55 shows one example.

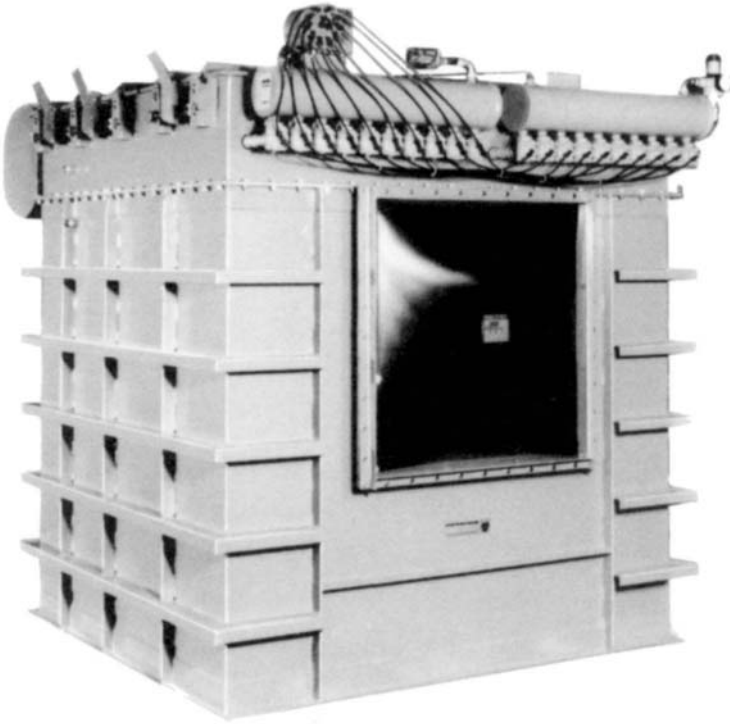


Figure 5–55 Reinforced 6 m^3 vented bag filter enclosure fitted with a modern 0.85 m^2, 3-layer bursting panel. P_{red} is 0.4 bar(g). From Eckhoff (2003).

Modern explosion vent panels burst reliably at the P_{stat} values for which they are certified, and are manufactured in a wide range of sizes and shapes, and coatings may be provided that allow permanent contact with various types of chemically aggressive atmospheres. Often a backing film of Teflon is used as an environmental protection to prevent the vent panel from contaminating the product inside the enclosure that is equipped with the vent. However, the upper working temperature limit of Teflon is about 230°C.

Hinged explosion doors constitute a third category of vent covers. Such doors may take a variety of different forms, depending on the equipment to be vented and other circumstances. Various kinds of calibrated locking mechanisms to ensure release at the predetermined P_{stat} have been developed. Hinged doors may be preferable if explosions are relatively frequent.

The final category of vent covers to be mentioned are the reversible ones, i.e. covers that close as soon as the pressure has been relieved. The purpose of such covers is to prevent secondary air from being sucked into the enclosure after the primary explosion has terminated, and giving rise to secondary explosions and fires. However, there is a risk of implosion that must be kept under control. The reversible vent covers include counterbalanced hinged doors and spring-loaded, axially traversing vent covers.

5.6.4.4.4 Potential Hazards Caused by Venting

Explosion venting prevents rupture of the enclosure in which the explosion takes place. However, significant hazards still remain. These include:

- ejection of strong flame jets from the vent opening
- emission of blast waves from the vent opening
- reaction forces on the equipment, induced by the venting process
- emission of solid objects (vent panels and other possible objects)
- emission of toxic combustion products

In general, flame ejection will be more hazardous the larger the vent and lower the static opening pressure of the vent cover. This is because with a large vent and a weak cover, efficient venting will start at an early stage of the combustion process inside the enclosure. Then large clouds of unburned explosive mixture will be pushed out through the vent and subsequently ignited when the flame passes through the vent. The resulting, secondary fire ball outside the vent opening can present a substantial hazard. If, on the other hand, the enclosure is strong, allowing the use of a small vent and a high P_{red}, mainly combustion products are vented, and the flame outside the vent is considerably smaller.

Reaction forces from explosion venting can significantly increase both the material damage and the extent of the explosion. Process equipment can tilt and ducts can become torn off, and secondary explosive clouds can be formed and ignited. Whenever an explosion vent is installed, it is therefore important to make an assessment of whether the equipment to be vented is able to withstand the reaction forces from explosion venting. A simple first order quasi-static consideration says that the maximum reaction force equals the maximum pressure difference between the interior of the vessel being vented and the outside atmosphere, times the vent

area. Experiments have confirmed that this simplified model in fact predicts reaction forces in fairly close agreement with the forces actually measured, as long as the duration of the pressure peak is not too short. However, for very fast explosions, dynamic (impulse) considerations may be required.

5.6.4.4.5 Vent Ducts

One traditional solution to the flame jet problem is the use of vent ducts. As illustrated in Figure 5–56, this means that a duct of cross-sectional area at least equal to the vent area is mounted between the vent and a place where a strong flame jet will not present any hazard. Vent ducts will generally increase the flow resistance, and therefore also the pressure difference to the atmosphere. Consequently, adding a vent duct increases the maximum explosion pressure in the vented vessel. Furthermore, the pressure increases with increasing duct length, increasing number of sharp bends and decreasing duct diameter.

5.6.4.4.6 The Quenching Tube

In some applications where venting ducts are difficult to implement, the quenching tube, invented by Alfert and Fuhre (1989) may provide a good solution. The principle of this device is illustrated in Figure 5–57.

If a dust explosion occurs in the enclosure to be vented, and the bursting panel, which constitutes an integral part of the quenching tube assembly, bursts, the explosion is vented through the comparatively large specially designed wall of the quenching tube. The wall is designed to yield low pressure drop, but high retention efficiency for dust particles and efficient cooling of combustion gases. This means that flame ejection from the vent is effectively prevented and the blast effects significantly reduced. Furthermore, burning lumps of powder and other smaller objects that could be ejected through an open vent, are retained inside the quenching tube. However, any toxic gaseous combustion products, e. g. carbon monoxide, will escape to the atmosphere. The increase of the maximum explosion pressure in the vented enclosure due to the flow resistance through the quenching tube wall is mostly moderate, and can normally be compensated for by a moderate increase of the vent area.

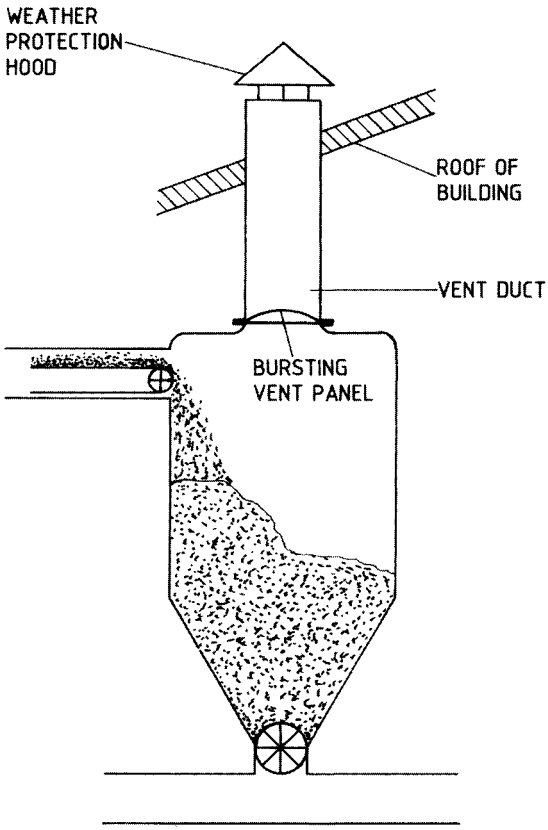


Figure 5–56 Illustration of the principle of vent ducts. From Eckhoff (2003).

5.6.4.5 Explosion-Pressure-Resistant Design

In most situations one can assume that the maximum pressure load from dust explosions is static. However, in some cases with very fast explosions, dynamic considerations may be recognized. The strength of some materials, including structural steels, is highly strain rate sensitive. This means that the stress at which plastic deformation starts, depends on the rate of loading. On the other hand, the damage to a structure also depends on how quickly the structure responds to the pressure loading. The natural period of vibration of the mechanical structure is normally used as a measure of the response time. If the duration of the pressure peak is long compared with the natural period of vibration, the loading can be considered as being

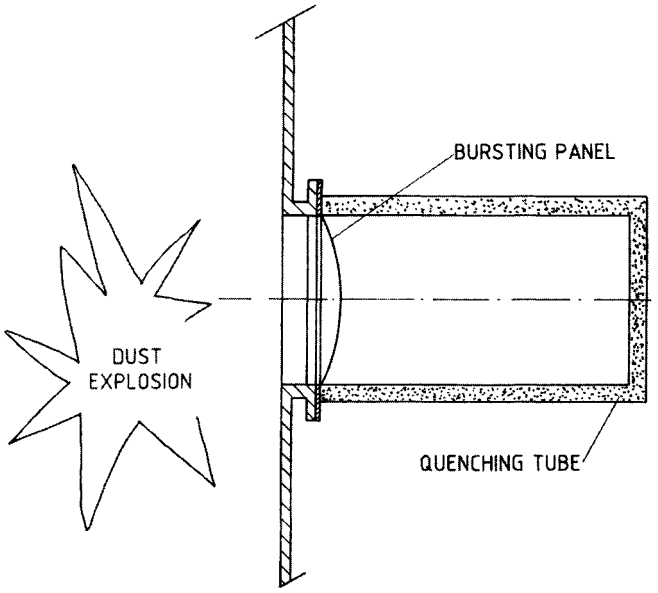


Figure 5–57 Illustration of the principle of the quenching tube for flame and dust free venting of dust explosions. From Eckhoff (2003).

essentially a static load. If, on the other hand, the pressure pulse is short compared with the response time of the structure, the damage is determined by the impulse, i.e. the time integral of pressure.

5.6.4.6 Automatic Explosion Suppression

The basic principle is described in Section 2.4.5.8. and illustrated in Figure 2–70.

Automatic suppression of dust explosions has been found to be feasible for organic dusts of maximum rate of pressure rise in the standard 1 m^3 closed ISO-vessel of up to 300 bar/s, i.e. K_{St} = 300 bar·m/s. It is somewhat unclear, however, whether the method can also be used for aluminum dusts of K_{St} in the range 300–600 bar·m/s. Moore and Cooke (1988) found that for aluminum flake of K_{St} = 320 bar·m/s it was difficult to ensure lower suppressed explosion pressures than about 2 bar(g), even under optimum conditions for suppression. In the case of dusts of natural organic materials and plastics of K_{St} up to 300 bar·m/s, the corresponding suppressed explosion pressures would typically been 0.2–0.4 bar(g). It

was therefore concluded that reliable suppression of the very violent aluminum flake explosions is difficult. However, they showed that a combination of explosion suppression and venting can reduce the maximum explosion pressure to a level significantly lower than the level for venting only. An alternative approach is to reduce K_{St} by partial inerting (see Section 5.6.4.2), but then one has to rely on two active systems, which is quite expensive.

Moore and Bartknecht (1987) conducted dust explosion suppression experiments in large vessels of volumes up to 250 m^3 and were able to show that successful suppression of explosions in clouds of organic dusts is possible even in such large volumes. However, as the vessel volume increases, more suppressant and faster injection are required for successful suppression. The actual design of suppression systems depends very much on the specific design of the suppressors, and other details which vary somewhat from supplier to supplier. Therefore generally applicable quantitative design criteria are difficult to specify.

Moore and Bartknecht employed three standardized types of suppressors. The smallest type, of volume 5.4 liters, was used for vessel volumes up to 5 m^3, whereas suppressors of 20 liters were used in the vessel volume range of 5–30 m^3. The largest suppressor type of 45 liters was used for the larger process volumes. The performance of the suppression system in large process volumes was verified experimentally in vessels of up to 250 m^3, for which ten of the 45 liter suppressors were required for successful suppression of St 2 dust explosions (organic dusts). For St 1 dusts, seven such suppressors were sufficient. However, these results refer to dust clouds of very high turbulence and homogeneity, and later investigations have shown that considerably smaller total suppressor volumes are required if the dust cloud is less turbulent and less homogeneous, which is often the case in industrial practice.

Moore (1989) compared venting and suppression and showed that the two explosion protection methods are to a great extent complementary. In practice, cost effective safety is achieved by using either one of the two methods, or a combination of both.

5.6.4.6.1 Influence of Type of Suppressant (Extinguishing Agent)

Traditionally halogenated hydrocarbons (halons) were used as suppressants in automatic dust explosion suppression systems. However, long

before the environmental problems caused by these chemicals became a major issue, Bartknecht (1978) showed that powder suppressants, such as $NH_4H_2PO_4$, were in general much more effective for suppressing dust explosions than halons. Therefore, powder suppressants have been used for suppressing dust explosions for many years. But powders differ in their suppressive power, and efforts have been made to identify the most effective ones. For example, addition of only 30 weight % of $NH_4H_2PO_4$ powder is required to prevent flame propagation in dust clouds in air of Pittsburgh bituminous coal, whereas with $CaCO_3$ dust (limestone) 70 weight % is needed. $NaHCO_3$ has proved to be an effective agent for suppressing some aluminum dust explosions. This material can also in some cases be used even in the food industry. It is soluble in water and can therefore be removed effectively by water only. Superheated steam (water at > 180°C) has also been used as a non-polluting suppressant.

5.6.4.7 Flexible Options for Explosion Prevention and Mitigation

Figure 5–58 and Figure 5–59, based on an analysis by Farber, illustrates how a given process plant can be protected against hazardous dust explosions by choosing quite different overall strategies. In Figure 5–58, the main strategy is explosion prevention by inerting using CO_2, whereas in the strategy adopted in Figure 5–59 explosion mitigation/control by venting and isolation plays a central role. Whenever a solution is developed for a given process plant, cost effectiveness is a major concern.

5.6.4.8 Good Housekeeping (Dust Removal/Cleaning)

5.6.4.8.1 General Outline

The main prerequisite for disastrous secondary explosions in factories is that significant quantities of combustible dust have accumulated outside the process equipment to permit development of explosive secondary dust clouds (see Section 5.2.7). Therefore, the possibility of extensive secondary explosions can be eliminated if the outside of process equipment, and shelves, beams, walls and floors of work rooms are kept free of dust.

Significant quantities of dust may accumulate accidentally outside process equipment due to discrete accidental events such as bursting of sacks

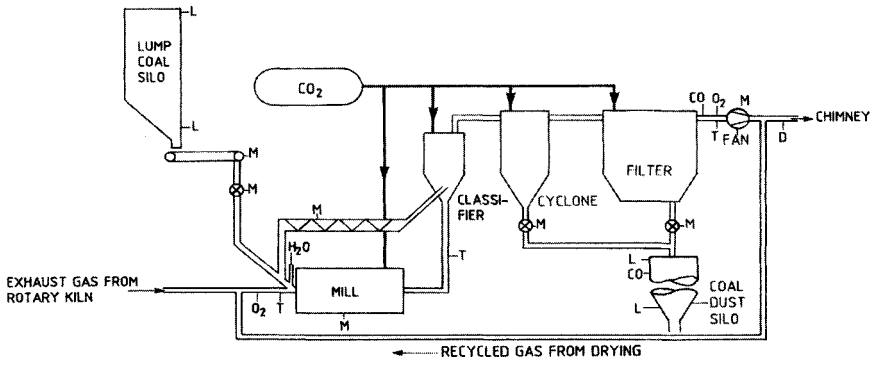


Figure 5–58 Comprehensive sensor system for monitoring, controlling and interlocking of a process for milling and drying of coal. Explosion protection based on inerting with CO_2.
CO = Carbon monoxide concentration sensors, D = Dust concentration sensor, L = Level sensors for coal and coal dust in silos, M = Movement sensors for mechanical components, O_2 = Oxygen concentration sensors, T = Temperature sensors. From Eckhoff (2003).

or bags or erratic discharge from silos or filters. In such cases it is important that the spilled dust be removed immediately. In case of large dust quantities the main bulk may be sacked by hand using spades or shovels, whereas industrial, explosion-proof vacuum cleaners should be used for the final cleaning. In the case of moderate spills, dust removal may be accomplished by vacuum cleaning only. Effective dust extraction should be provided in areas where dusting occurs as part of normal operation, e.g. at bagging machines.

Considerable quantities of dust can accumulate outside process equipment over a long time due to minor but steady leaks from process equipment. The risk of such leaks is comparatively large if the working pressure inside the process equipment is higher than ambient pressure, whereas running the process at slightly lower than ambient pressure reduces the leaks.

Process equipment should be inspected regularly for discovery and sealing of obvious accidental leak points as early as possible. However, often one has to accept a certain unavoidable level of dust leaks from process equipment. It is then important to enforce good housekeeping routines by which accumulations of combustible dust outside process

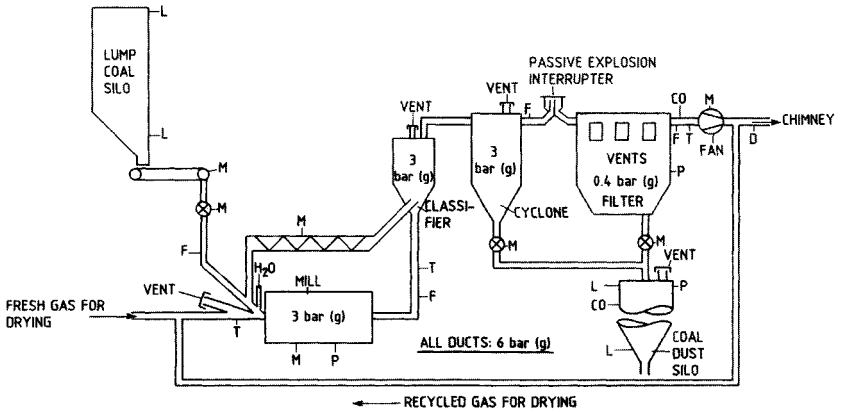


Figure 5–59 Comprehensive sensor system for monitoring, controlling and interlocking of a process for milling and drying of coal. Explosion protection based on venting and explosion shock resistant design.,
CO = Carbon monoxide concentration sensors,
D = Dust concentration sensor,
F = Flame sensor,
L = Level sensors for coal and coal dust in silos,
M = Movement sensors for mechanical components,
P = Pressure sensors
T = Temperature sensors.
From Eckhoff (2003).

equipment are removed at regular intervals, preferably by explosion-proof vacuum cleaning.

Use of compressed air for blowing spilled dust away should be prohibited. By this method dust is not removed, but only transferred to another location in the same room. Besides, dust explosions can result if the dust concentration in the cloud that is generated is in the explosive range and an ignition source exists in the same location.

5.6.4.8.2 Industrial Explosion-Proof Vacuum Cleaners

The subject was discussed by Beck and Jeske (1989) who listed the requirements to mobile type 1 vacuum cleaners recommended in F. R. Germany for removal of combustible dusts:

- The fan must be on the clean side and protected against impacts by foreign bodies.
- The electric motor and other electric components must satisfy the general requirements to such components that are to be used in areas containing combustible dusts. Motors must be protected against short-circuits and overheating.
- The exhaust from the vacuum cleaner must be guided in such a way that it does not hit dust deposits and generate dust clouds.
- All electrically conducting parts of the equipment, including hose and mouthpiece, must be earthed with a resistance to earth of less than 1 ΩM.
- Vacuum cleaner housings must be constructed of materials that are practically non-flammable. Aluminum and aluminum paints must not be used.
- A clearly visible sign saying "No suction of ignition sources" should be fitted to the housing of the vacuum cleaner.

Figure 5–60 shows an example of a large mobile vacuum cleaner for combustible dusts in industry.

Sometimes it is useful to install stationary vacuum cleaning systems rather than having mobile ones. Then a central dust collecting station with suction fan is connected to a permanent tube system with a number of plug-in points for vacuum cleaning hoses at strategic locations. Good housekeeping is essential because clean work rooms exclude the possibility of extensive secondary explosions. Cleanliness also improves the quality of the working environment in general.

Figure 5–60 Large mobile vacuum cleaner for collecting combustible/exposable dusts in industry. Both the main vessel and the connecting ducts are designed to withstand internal explosion pressures of 9 bar(g). Power requirement 45-55 kW. From Eckhoff (2003).

Chapter 6

Explosives, Pyrotechnics, and Propellants

6.1 Chemical Composition of Some Explosives

An explosive material is a material or a mixture of materials, which following activation liberates gases and heat due to a fast exothermal chemical reaction. A common feature of these materials, which makes them differ from explosive gases, sprays/mists, and dusts clouds, is that the oxygen of the atmosphere with which the material is mixed is not an essential participant in the heat generating chemical reaction. The different materials of this type used in common explosives may be grouped in the following categories:

(a) dry mixtures of powders of materials that can burn (e.g. metals, carbon, sulphur, starch, wood), and materials that can supply oxygen (e.g. nitrates)

(b) water-based suspensions (slurries) or gels of fuel powders (e.g. aluminum) in liquid solutions of solid materials that can supply oxygen (e.g. ammonium nitrate)

(c) nitrated organic compounds (e.g. nitro glycol, nitro glycerol, di-nitro-toluene, tri-nitro-toluene, tri-nitro-phenyl-methyl-nitramine). Such materials contain the fuel (carbon, C and hydrogen, H) and the source of oxygen for combustion (nitro-groups, NO_2) in the same molecule

(d) primary explosives (e.g. lead azide), which, when exposed to mechanical impact above a certain strength, will decompose spontaneously by detonation. Such primary explosives are

used to initiate detonation in other less reactive explosives, which would just deflagrate when exposed to less vigorous ignition sources.

Commercial explosives used in practices are often mixtures/combinations of materials belonging to the categories (a) through (d). One example is given in Table 6–1.

Table 6–1 Composition of a Typical Gelatine Dynamite

Chemical component	% by mass
Ammonium nitrate	60 - 65
Nitro-glycole	20 - 21
Nitro-glycerole	5 - 6
Di-nitro-toluen	6 - 8
Nitro-cellulose (providing gelatine)	1
Additives to produce desired consistency (e.g. starch, "kiselguhr", wood powder)	Balance to 100 %

Another example is AN/FO, which is simply a mixture of solid crystalline ammonium nitrate moistened with diesel oil. In practice, categories (a) and (b) may interchange, depending on the amount of water added. For example, a saturated solution of ammonium nitrate in water may exist side by side with solid crystalline ammonium nitrate.

CENELEC (1997) defines explosive substances as solid, liquid, pasty or gelatinous substances and preparations that are liable to react exothermally and under rapid gas generation even without the participation of atmospheric oxygen, and which, under specified test conditions under partial confinement, will detonate or deflagrate rapidly during heating. It is important to note that this definition not only comprises the genuine explosives, but also pyrotechnics and propellants. Equivalent to explosive substances are:

(a) substances or preparations capable of exploding under certain realistic conditions, and intended for use as blasting explosives, but not generally defined as explosives

(b) initiation and igniting devices, and pyrotechnics

(c) other products in which explosive substances or substances or preparations in accordance with a) are totally or partially confined and in which an explosion is initiated

6.2 Combustion Propagation in Explosives/Pyrotechnics/ Propellants

6.2.1 Deflagration and Detonation

As in the case of premixed gases and sprays/mists and dust clouds, explosives, pyrotechnics, and propellants also possess two distinctly different modes of exothermal reaction, viz. deflagration and detonation. In fact, with these materials, the difference between the two modes are generally much more dramatic than with gases, sprays/mists, and dusts.

In the case of deflagration, the velocity of propagation of the reaction zone is much lower than the velocity of sound in the material. In a deflagration wave, the direction of flow of the reaction products is opposite to the direction of propagation of the reaction wave. The burning of a gun powder or of a rocket propellant charge is a deflagration process.

In the case of detonation the velocity of propagation of the reaction zone is supersonic and the flow of the reaction products is in the same direction as that of the reaction.

The mode of exothermic reaction of an explosive material once ignited, i.e. deflagration or detonation, depends on the mode of initiation of the reaction. Deflagration results if the ignition source is not generating a shock wave of the strength required for initiation of detonation. However, if the reactive material is confined, the liberated heat in a deflagration will generate a pressure rise in the still unreacted material, and substantial acceleration of the exothermal chemical reaction can result.

Table 6–2 illustrates the dramatic difference between the features of detonation and unconfined deflagration for one specific explosive material. The ratio of the two mass conversion rates is more than 10^7 and of the two energy production rates 10^9.

6.2.2 Experimental Methods for Measuring Detonation Properties

6.2.2.1 Dautriche Method

The Dautriche method for determination of detonation velocities of explosives, as described by Meyer (1987), is illustrated in Figure 6–1.

Table 6–2 Comparison of Deflagration and Detonation Properties of a Nitro-glycole Explosive of Density 1.5 g/cm^3 [a]

Property	Deflagration	Detonation
Flame front propagation velocity	0.3 mm/s	7.3 km/s
Thickness of reaction zone	10 mm	1 mm
Liberated energy in reaction zone	1.9 kJ/g	6.7 kJ/g
Mass conversion rate	0.45 kg/m^2·s	~ 10^7 kg/m^2·s
Energy production rate	85 MJ/m^2·s	$7.4 \cdot 10^{10}$ MJ/m^2·s

a. The data for deflagration were obtained at fully unconfined conditions.

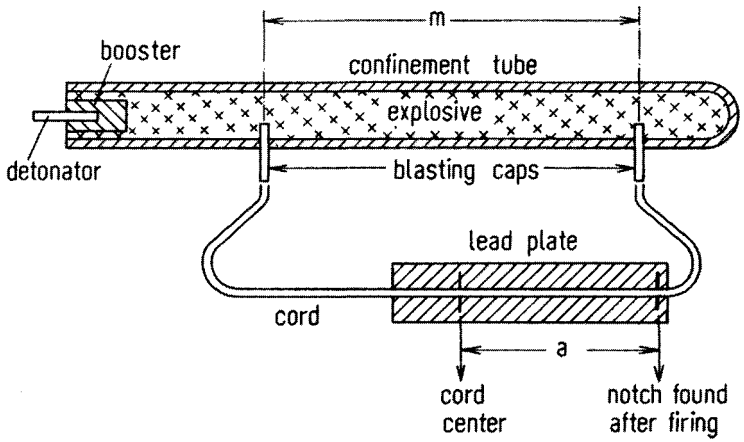


Figure 6–1 Dautriche Method for Measurement of detonation velocities in explosives. From Meyer (1987).

The sample of the explosive to be tested is placed in a confinement tube, which may or may not be surrounded by a steel sleeve. The part of the sample along which the detonation velocity is to be measured, is defined by two blasting caps. The ends of a piece of detonating cord of known detonation velocity is connected to the blasting caps, and the cord is passed across a lead plate as indicated in the figure. The detonation wave in the test sample is initiated by the detonator/booster at the upstream end of the test tube, and a steady detonation wave front starts to travel along the confinement tube. As soon as this wave hits the first blasting cap, a second detonation wave is instantaneously initiated at the end of

the detonating cord connected to that cap. At the same time, the detonation wave in the test sample continues down the confinement tube, and eventually it reaches the second blasting cap. This then initiates a third detonation wave at the other end of the detonating cord. Eventually, the two detonation wave fronts in the cord, traveling at equal speeds but in opposite directions, collide and produce a notch in the lead plate at the point of collission. Clearly, the time required for the two opposite detonation waves to travel from the first blasting cap to the collision point at the lead plate is the same. Hence, for the wave going directly into the detonating cord of total length L, the time is $(\frac{1}{2}L + a)/D_{cord}$, whereas for the other wave it is $m/D_{sample} \cdot (\frac{1}{2}L - a)/D_{cord}$. By equating the two expressions, one gets

$$m/D_{sample} = 2a/D_{cord} \tag{6.1}$$

or

$$D_{sample} = D_{cord} \cdot m/2a \tag{6.2}$$

6.2.2.2 Detonating Cap Sensitivity Test

In this case, the objective of the test is to determine the response of a sample of the explosive to a standardized detonating cap. The results are used to determine the classification of the explosive as a transport hazard. According to Meyer (1987), the test method illustrated in Figure 6–2 has been used in Germany.

The explosive sample is placed into a 200 mm-long cardboard sleeve of 80 mm internal diameter and wall thickness between 1.3 and 1.4 mm. The bottom end of the sleeve is sealed by a thin cardboard disk, which is glued into position. Because the cap sensitivity can be influenced by the density of the sample under test, the density is determined by weighing. The cardboard test sleave with the test sample is placed upright onto a steel plate of 1 mm thickness, resting on a steel ring of 50 mm height, inside diameter 100 mm, and wall thickness 3.5 mm. A standardized test fuse is inserted at the top of the sample and initiated. If a proper detonation wave is initiated in the sample, a sharp circular hole is cut out of the steel plate. If no change in the condition of the steel plate, or only denting with or without fissure is observed, this is classed as a non-detonation.

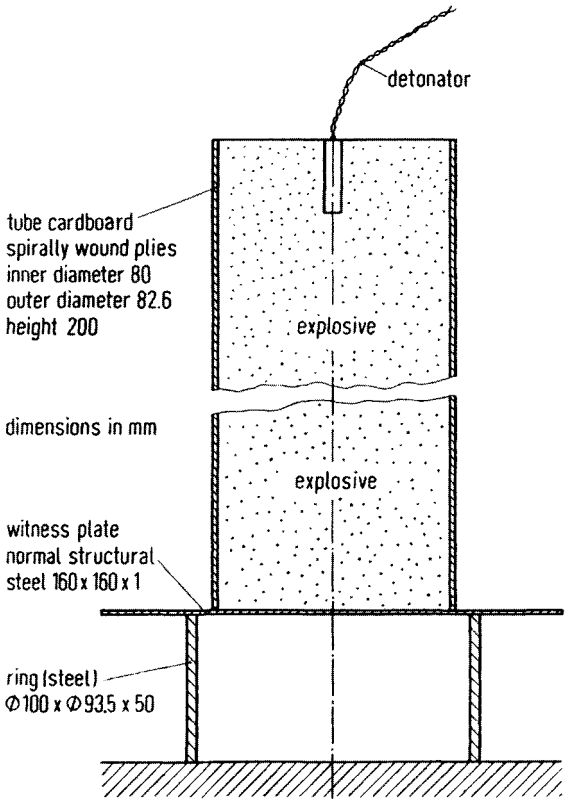


Figure 6–2 The cap sensitivity test for classifying explosives with regard to the explosion hazard that they represent when being transported. From Meyer (1987).

6.2.2.3 Trauzl Lead Block Test

The Trauzl lead block test is a comparative method for the determination of the strength of an explosive. The principle of the test is illustrated in Figure 6–3.

A sample of 10 g of the test material, wrapped in tinfoil, is introduced into a central bore hole (125 mm deep, 25 mm diameter) in a massive cylinder of soft lead, of height and diameter equal to 200 mm. A standardized blasting cap with an electric primer is introduced into the center of the explosive charge, and the remaining free space is filled with quartz sand of standard grain size. After the explosion, the volume of the resulting

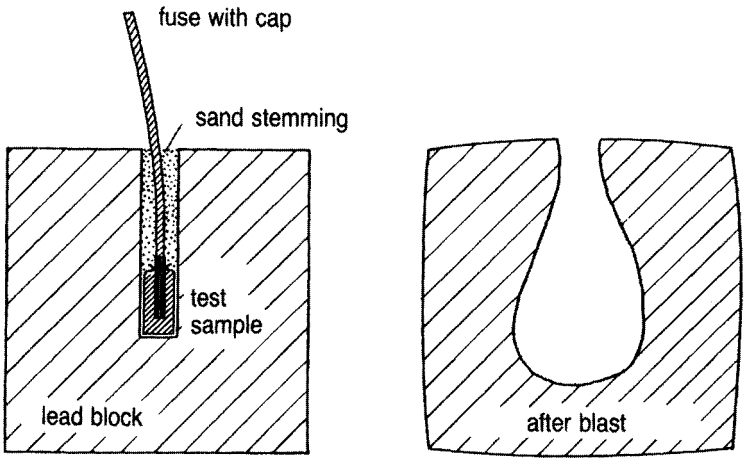


Figure 6–3 The Trauzl lead block test for comparing the strength of explosives. From Meyer (1987).

bulge is determined by filling it with water. The volume of the original cavity is deducted from the result obtained. 10g of picric acid may be used as a standard explosive charge for comparison.

In a modified lead block test, recommended by BAM in Germany, the test sample is prepared as follows: A special instrument wraps the sample in a piece of tinfoil and molds it into a cylinder of 11 cm^3 capacity The resulting bulk density should be only slightly higher than the bulk density by pouring. Liquids are filled into thin-walled cylindrical glass ampoules or, in special cases, directly into the cavity of the lead block. The initiation is effected by means of an electric blasting cap containing lead azide as the initiating charge and PETN as the secondary charge. The empty space above the test sample is filled with dried quartz sand as in the original method.

6.3 Ignition/Initiation of Explosives/Pyrotechnics/Propellants

6.3.1 Decomposition Temperature

The decomposition temperature of explosive materials, propellants, and pyrotechnics is the minimum temperature at which a small sample of the material, placed in an externally heated test tube, bursts into flame, decomposes rapidly or detonates violently.

Meyer (1987) describes the following method for determining the decomposition temperature: 0.5 g (0.01 g in the case of initiating explosives!) is placed in a test tube, which is then immersed in a liquid metal bath (preferably Wood's metal) at 100°C. The temperature of the bath is then raised at a rate of 20°C/min. until deflagration/decomposition takes place.

CENELEC (1997) defines the decomposition temperature as the lowest temperature at which a given quantity of the material tested ignites in a carefully standardized differential thermal analysis (DTA) test. The basic principle of DTA is that the sample under test and a sample of the same size of an inert material are heated in parallel and the temperature difference between the two samples recorded continuously. The sample size is in the range 0.1 and 0.5 g, whereas the heating rate shall not exceed 5°C/min. Figure 6–4 shows an example of the result of a DTA test. In this particular case, the first recorded difference between the temperature of the test sample and the inert reference sample is a modest negative peak due to endothermic melting of the test sample. However, as the temperature rises further, the major exothermal peak of the decomposition reaction appears. The decomposition temperature in this test is defined as the "extrapolated peak initial temperature," i.e., the reference sample temperature at which the steepest tangent to the rising part of the decomposition peak intersects with the base line, as indicated in Figure 6–4.

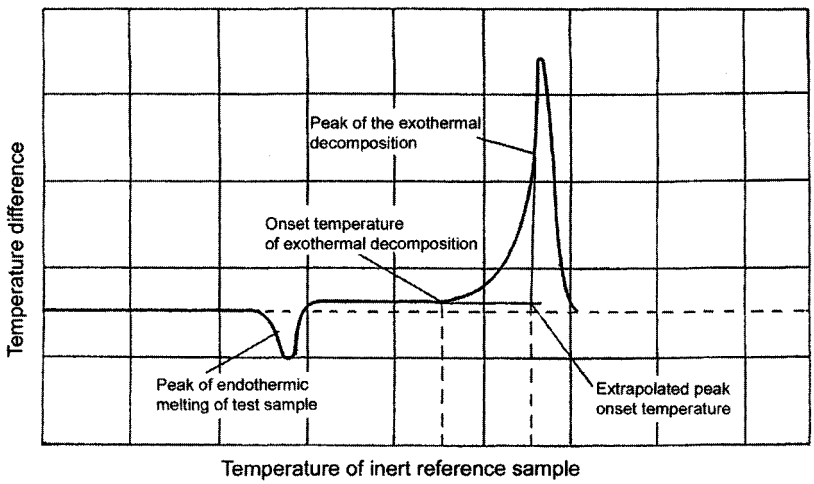


Figure 6–4 Example of result of DTA test for the decomposition temperature of explosives, propellants and pyrotechnics. From CENELEC (1997).

Table 6–3 gives some results for a range of materials obtained by the DTA method.

Table 6–3 Decomposition Temperatures of Explosive Substances Used as Blasting Explosives, Propellants, Ignition Substances, or Pyrotechnics From CENELEC (1997)

Chemically uniform substances				
No.	Chemical name	CAS-No.**	EINECS***-No.	Decomposition temperature °C
1	mannitol hexanitrate	15825-70-4	239-924-6	140 1)
2	mercury difulminate	628-86-4	211-057-8	151 1)
3	glycerol trinitrate	55-63-0	200-240-8	155 2)
4	oxydiethylene dinitrate	693-21-0	211-745-8	170 3)
5	pentaerithrityl tetranitrate	78-11-5	201-084-3	164 1)
6	N-methyl-N,2,4,6-tetranitroaniline	479-45-8	207-063-0	171
7	ethylene dinitrate	628-96-6	211-063-0	172 3)
8	cellulosenitrate (nitrocellulose)	9004-70-0		176 1)
9	perhydro-1,3,5-trinitro-1,3,5-triazine	121-82-4	204-500-1	208 1)
10	picric acid	88-89-1	201-865-9	214 1)
11	styphnic acid	5704-04-1	227-193-6	224 1)
12	dipicrylamine	131-73-7	205-037-8	244 3)
13	lead 2,4,6-trinitro-m-phenylene dioxide	15245-44-0	239-290-0	256 1)
14	lead diazide	13424-46-9	236-542-1	268 1)
15	2,4,6-trinitrotoluene	118-96-7	204-289-6	271 1)
16	dipricrylamine	131-73-7	205-037-8	301 3)
17	1,3,5-trichlorotrinitrobenzene	2631-68-7	220-115-1	305 3)
18	1,3,5-trinitrobenzene	99-35-4	202-752-7	327 3)
19	2-chloro-1,3,5-trinitrobenzene	88-88-0	201-864-3	330 3)
20	2,4,6-trinitrobenzoic acid	129-66-8	204-958-2	337 3)*)
21	ammonium perchlorate	7790-98-9	232-235-1	341 3)
22	5-ISMN			183-193
23	ISDN			175-200

*) At 190 °C exothermal prereaction (decarboxylization)
**) Chemical abstract service number
***) European inventory of existing commercial chemical substances

1) DTA open; 0,1 g sample mass
2) DTA open; 0,5 g sample mass
3) DTA pressure-tight 0,5 g sample mass

6.3.2 Drop Hammer Sensitivity Test

The sensitivity to impact of solid, liquid, or gelatinous explosives is tested by the drop hammer method. The principle of this method is illustrated in Figure 6–5.

The samples of the explosives are subjected to the action of falling weights of different masses. The parameter to be determined is the height

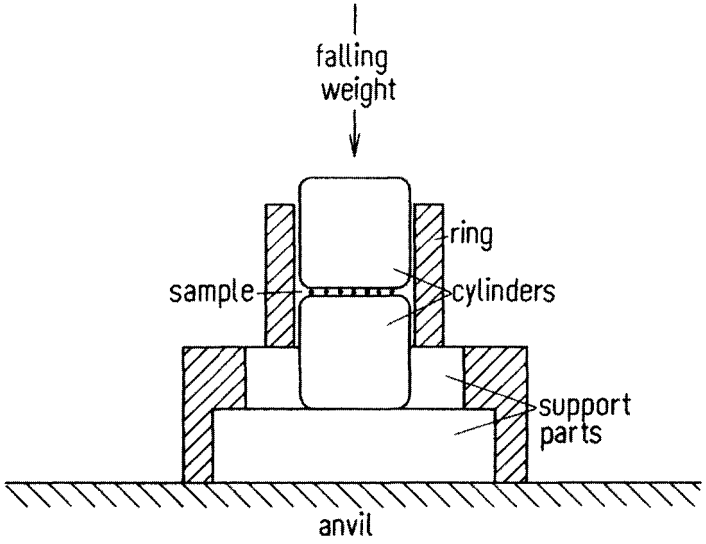


Figure 6–5 Illustration of the drop hammer impact sensitivity test. From Meyer (1987).

of fall, for one specific weight mass, at which a sufficient amount of impact energy is transmitted to the sample for it to decompose or to explode. According to Meyer (1987), the U.S. standard procedures are:

6.3.2.1 Solids

In the U.S. Bureau of Mines test, a sample of 0.020 g of the explosive is tested, using a fixed dropping mass, usually 2 kg. The result of the test is the minimum height at which at least one of 10 repeated trials results in explosion. In the U.S. BM apparatus, the explosives sample is held between two flat, parallel hardened steel surfaces, as illustrated in Figure 6–5.

6.3.2.2 Liquids

In the case of the U.S. BM apparatus (illustrated in Figure 6–5), the same basic procedure as used with solids is applied with the following modifications:

- the mass of explosives tested is 0.007 g
- a disc of desiccated filter paper in which the liquid sample is absorbed is placed on the anvil, and the plunger is lowered onto the sample

According to Meyer (1987) The drop hammer method has been modified by the German Bundesanstalt for Materialprüfung (BAM), so as to obtain better reproducibility of data.

6.3.3 Electric Spark Sensitivity Test

6.3.3.1 Original U.S. Bureau of Mines Test

In this method described by Dorsett et al. (1960), the standard dust layer thickness tested was 1.6 mm. The layer was resting on a 25 mm-diameter steel plate that also served as the negative electrode. The positive needle-point electrode, connected to a capacitor bank charged to 400 V, was lowered by hand towards the surface of the dust layer until a spark discharge occurred. After an ignition occurred, the steel plate was cleaned, a new dust layer formed and the process repeated at progressively lower capacitance values until the lowest that gave at least one ignition in twenty trials was identified. The minimum ignition energy was defined as $\frac{1}{2}CU^2$, where C is the lowest capacitance giving ignition and U the charging potential of 400 volts.

6.3.3.2 Nordtest Fire 016

This method, described by Nordtest (1982), is primarily intended to be used for pyrotechnics and explosives in pulverized form. It may, however, also be applied to normal combustible materials in pulverized form, which, when distributed as a thin layer resting on a flat metal surface, are able to propagate self-sustained combustion.

The test apparatus is illustrated in Figure 6–6.

The dust/powder is poured gently into the disc-shaped cavities formed by the slidable supporting plate/hole plate assembly and excess dust is removed by a scraper. Plane, circular dust/powder samples of thickness 2 mm and diameter 12 mm are thus obtained. The metal bottom of the

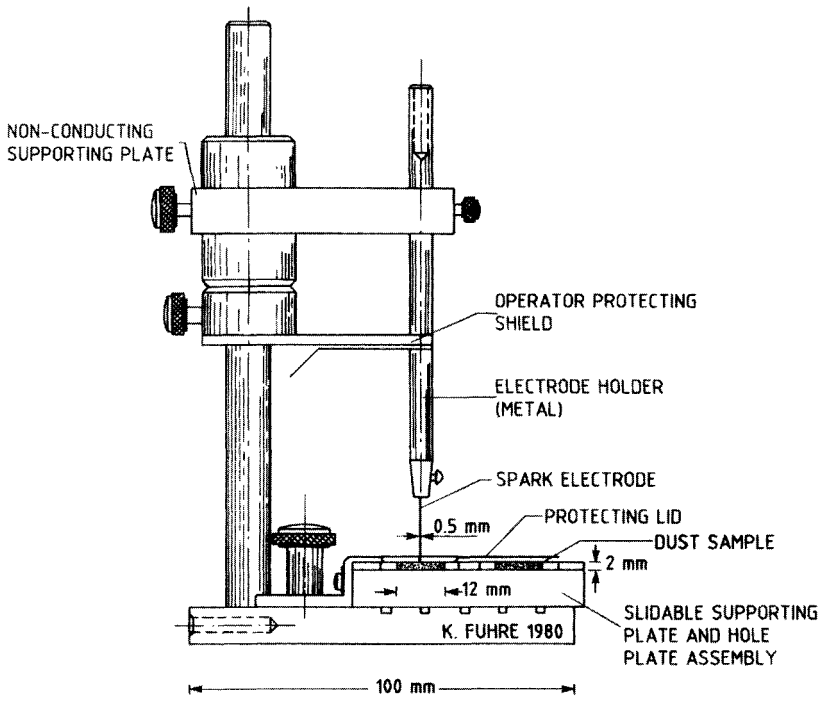


Figure 6–6 Cross-section of Nordtest (1982) apparatus for determination of electric spark sensitivity of layers of explosives, pyrotechnics and propellants.

cavities acts as the earthed one of the two electrodes forming the spark gap. A thin tungsten wire pointing downwards towards the dust/powder layer and with its tip just above the dust/powder surface, acts as the second high voltage electrode.

An electric spark of the desired net energy and discharge time is passed through the sample, and it is observed whether or not ignition occurs. An electric spark generator that permits independent variation of spark energy and spark duration is required. Twenty nominally identical tests are carried out at each combination of net spark energy and discharge time, yielding a frequency of ignition somewhere in the range of 0 to 100%. After each spark discharge, the dust/powder sample is shifted horizontally to allow each spark to pass through fresh dust/powder that has not been exposed to previous sparks. If ignition occurs, the sample tested is discarded and the test continued with a new sample.

The minimum electric spark ignition energy, defined as the net spark energy yielding an ignition frequency of 5%, is determined for various

spark discharge times or durations Δt. The ultimate result of the test is an electric spark sensitivity profile $E_{min}(\Delta t)$, as illustrated by the hatched area in Figure 6–7.

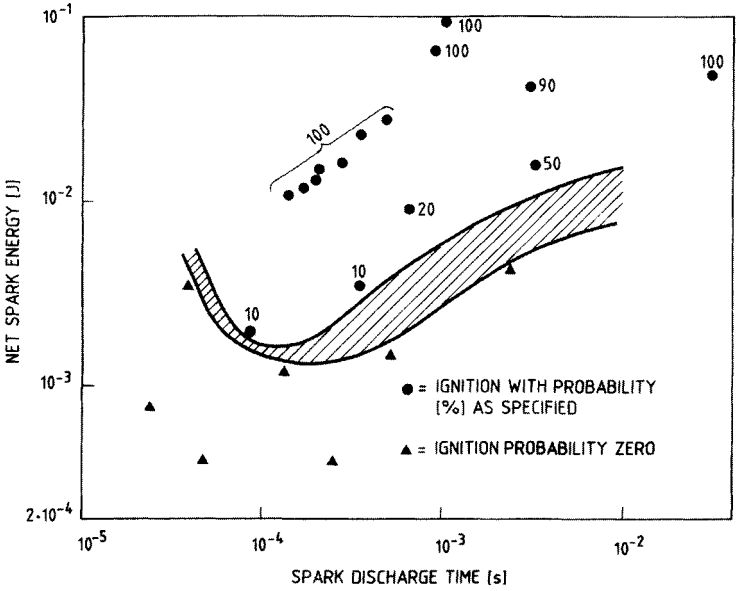


Figure 6–7 Electric spark ignition sensitivity profile of a given pyrotechnics product as determined by the Nordtest (1982) method.

6.4 Case Histories of Accidental Explosives/Pyrotechnics/ Propellants Explosions

6.4.1 Ammonium Nitrate Explosion Catastrophe in Ludwigshafen, Germany (1921)

At 0730 on 21 September 1921 a major explosion occurred at Oppau near Ludwigshafen, in the works of the Badische Anilin und Soda Fabrik (BASF). The account given here is from Marshall (1987). The explosion killed 561 people, including 4 who stayed in Mannheim 7 km away from the explosion center; approximately 1,500 people were injured. In the town of Oppau, 75% of the houses—around 1,000—were destroyed. In the explosion, about 4,500 tons of a mixture of ammonium nitrate and

ammonium sulfate blew up. The material was most probably detonated by explosive charges that were customarily used to break up caked material. According to one source, the material was a co-crystallized mixture of ammonium nitrate with ammonium sulfate, in the molecular proportions 2:1, which would correspond to about 58% ammonium nitrate by weight. Another source claimed that the mixture that blew up was 50% by weight of nitrate and sulfate. The same source claimed that it was proved impossible to detonate any mixture containing less than 60% nitrate, but that the mixture was not homogeneous and may have contained up to 70% nitrate in some places. If, however, as pointed out by Marshall (1987), the mean content was 50% nitrate, some places must have been considerably leaner than this, and it is difficult therefore to see why the whole heap blew up. The decision to use explosives to break up caked material was based on exhaustive tests, and many thousands of actual cake break-up blastings had been conducted. According to BASF, the crater generated by the explosion was 10 m deep, 75 m wide and 115 m long. Within a radius of 250–300 m, all buildings were totally destroyed or so badly damaged that they had to be demolished.

6.4.2 Ammonium Nitrate Explosion Catastrophe in Texas City, U.S.A. (1947)

This account is also based on Marshall (1987). The disaster occurred at 0912 on 16 and 17 April 1947 in the ships *Grandcamp* and *Highflyer*. *Grandcamp* was carrying about 2,300 tons of ammonium nitrate in paper sacks. According to one source, a fire had started among the sacks about an hour before the explosion. It was decided, to avoid damaging the nitrate, not to use water for extinguishing the fire. Instead one tried to put the fire out by battening down the hatches and admitting steam. As pointed out by Marshall, admitting hot steam probably just worsened the situation by heating up the nitrate.

The explosion killed over 200 people who had crowded to the dockside to watch the fire, the ship's crew of 32 people, and 4 people in two aircraft circling overhead. Severe damage occurred and new widespread fires were initiated in a nearby Monsanto complex, as well as in the tank parks of a number of oil companies in the neighborhood. Pieces of wreckage traveled up to one km.

The day after, at 1310 on 17 April, the second ship, *Highflyer*, which had caught fire the previous day as a consequence of the explosion on *Grandcamp*, also blew up. This ship was also carrying ammonium nitrate and, according to some reports, sulphur.

The total death toll due to the two explosions was estimated at 552. In addition, 200 people were missing and 3,000 sustained injuries. The material damage was estimated in excess of U.S. $100 million (1947 prices).

This disaster shows that, in some circumstances, mild deflagrations or anaerobic fires in ammonium nitrate may escalate to a detonation even when barely confined. According to Marshall (1987) the quantity of ammonium nitrate that actually exploded cannot be stated with certainty. He suggested that about 2,000 tons of ammonium nitrate may have detonated in each of the two explosions, and that this would have been equivalent to about 1,000 tons of TNT in each case. The explosions led to pronounced "domino" effects by starting fires in nearby chemical plant oil storages and dry goods warehousing.

From the actions taken, it seems clear that the crew on *Grandcamp* had no training in how to deal with anaerobic fires. Furthermore, the harbor was inadequately equipped to deal with major fires in general. The death toll was undoubtedly considerably aggravated by the phenomenon of "negative evacuation," i.e. people flocked to the scene to witness the fire. Had the authorities acted (or been able to act) to evacuate the area, the death toll could have been far less. According to Marshall (1987) the disaster in Texas City is not the only example of people being killed while being spectators at the scene of a disaster.

6.4.3 Explosives Mix Explosion in Mustang, Nevada, USA (1998)

6.4.3.1 Introduction

On 7 January 1998, at 0754, two explosions in rapid succession destroyed the Sierra Chemical Company's Kean Canyon plant east of Reno near Mustang, Nevada, U.S.A. killing four workers and injuring six others. Because of the loss of life and extensive damage, the United States Chemical Safety and Hazard Investigation Board (CSB) sent a team to investigate the explosion in an attempt to understand the causes of this incident. The present account is a brief summary of the comprehensive CBS (1998) report.

6.4.3.2 The Kean Canyon Plant and its Production

The Sierra Kean Canyon plant manufactured explosive boosters, mixed custom flux for gold smelting operations, and repackaged bulk soda ash for sale to the mining industry. When initiated by a blasting cap or detonation cord, boosters provide the added energy necessary to detonate less sensitive blasting agents or other high explosives. The boosters manufactured at the Kean Canyon plant consisted of a base mix and a second explosive mix, called Pentolite, both of which were poured into cardboard cylinders (see Section 6.2.2.2). The operators working in the plant were responsible for the preparation of the explosive mixtures, the operation of the mixing pots, and the pouring of the mixtures into the cardboard cylinders. Figure 6–8 gives an approximate outline of the part of the plant that was demolished by the accidental explosion.

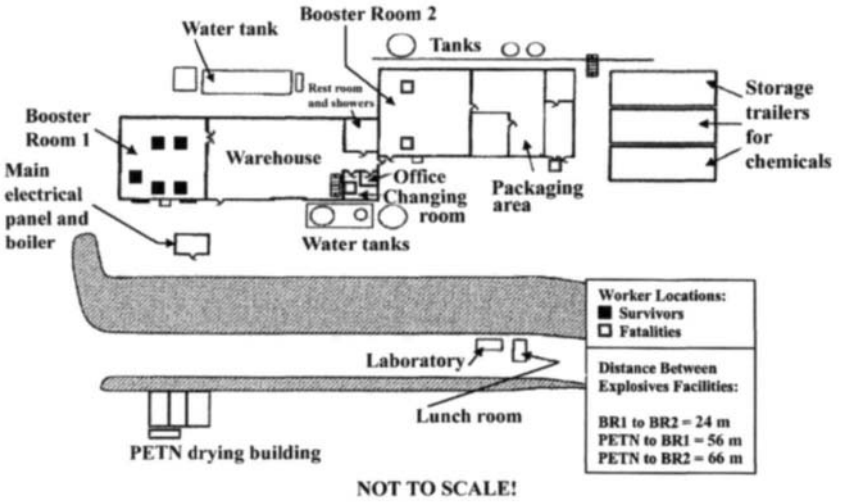


Figure 6–8 Approximate plan of the Sierra Chemical Company Kean Canyon plant east of Reno near Mustang, Nevada, USA. From CSB (1998).

The boosters were filled in the two Booster Rooms indicated in the figure. The primary explosives used in the base mix were TNT (2,4,6 trinitrotoluene), PETN (pentaerythritol tetranitrate), and a mixture of TNT and RDX (hexahydro 1,3,5 trinitro 1,3,5 triazine). Pentolite was a mix of TNT and PETN.

6.4.3.3 The Explosions

CSB investigation team identified four possible explosion scenarios that could account for the explosions that occurred, but based on the seismic data, interviews of workers, and the physical evidence observed during the investigation, they considered one of these as being more likely than the others. They concluded that most probably the first explosion occurred in the plant's Booster Room 2, indicated in Figure 6–8, whereas the second, larger explosion followed 3.5 s later in the PETN building, which is also indicated in Figure 6–8. The two explosions were recorded by the Seismology Laboratory at the University of Nevada, Reno, and they were also able to conclude that the site of the first explosion was somewhat further to the north than that of the second, larger explosion, which in accordance with Booster Room 2 being located further to the north than the PETN building. The interval of 3.5 s between the explosions was estimated by the laboratory to be accurate to ± 0.2 s.

Most probably the first explosion was initiated in a mixing pot in Booster Room 2 (see Figure 6–8), when an operator who had left about 25–50 kg of base mix in his mixing pot the day before, turned on the mixing stirrer motor of the pot the next morning. During the evening and night the mix had stratified and solidified, and the explosion initiation occurred when the bottom of the mixer blade, which was embedded in the solidified explosives mix in the pot, forced the explosive material by impact, shearing, or friction against the pot wall. Alternatively, explosive material was pinched between the mixer blade and the pot wall, causing the initial detonation in the pot. The shock wave from this initial detonation then instantaneously detonated the other several thousand kg of explosives in the Booster Room 2, leading to the first main explosion that completely demolished Booster Room 2, and also the rest of the building. Only three walls were left of Booster Room 1, as indicated in Figure 6–9, whereas Booster Room 2 was completely levelled with the earth, as can be seen in Figure 6–10.

When the first explosion occurred, a worker in Booster Room 1 saw a huge fireball engulf a truck that was parked immediately outside the building. This worker was thrown against one of the walls in Booster Room 1, as the ceiling and another wall of the room collapsed on top of him and the four other workers.

Seconds later the same worker heard a second, more powerful (louder) explosion in the PETN building. CSB concluded that this explosion was

Figure 6–9 Remains of Booster Room 1 of the Sierra Chemical Company Kean Canyon plant east of Reno near Mustang, Nevada, USA, after explosives mix explosion in 1998. From CSB (1998).

most likely initiated by a heavy piece of equipment or burning debris from the first explosion falling through the reinforced concrete roof or the skylight of the building. The site of the PETN building and an adjacent magazine was now transformed to a 13 m × 20 m wide and 2 m deep crater. The explosions were felt as far as 35 km away.

A total of 11 employees were at the site at the time of the explosions. After the explosions, five workers in Booster Room 1 were trapped temporarily under the collapsed building, but were able to crawl out within a few minutes. Three of the five workers were seriously injured, whereas two only received minor injuries. Concerned about possible additional explosions, the workers from Booster Room 1, after calling for other possible survivors, went to the entrance to the facility. There they met two other workers who had been in the gravel pit below the site, approximately 100 m southwest of the PETN building. The other four workers who were believed to have been in or near Booster Room 2 had been killed by the explosions.

The blast effects of the explosions leveled the site and threw structural materials, manufacturing equipment, raw materials from the booster and flux operations, and other fragments up to 1 km away. The legs and cross

Figure 6–10 Remains of Booster Room 2 of the Sierra Chemical Company Kean Canyon plant east of Reno near Mustang, Nevada, USA, after explosives mix explosion in 1998. From CSB (1998).

bracing from an empty tank, which previously stood at the corner of the changing room, were thrown 300 m away from the production building. The doors of one of the large magazines and a portable magazine located west of the production facility were sprung open by the negative pressure pulse; however, large quantities of explosive materials that were stored inside did not detonate. Many un-detonated boosters had been scattered throughout the site by the explosion. Other hazards remaining after the two major explosions as the after-fire progressed included fires, toxic chemicals, and potential detonation of the explosives in Booster Room 1.

6.4.3.4 Events Leading to the Explosion

The CSB investigation concluded that the following sequence of events led to the explosion catastrophe. The day before the accidental explosion, one melt/pour operator working in Booster Room 2 (see Figure 6–8)

needed to leave work early, at 3 p.m. When he left, there were between 25 and 50 kg of base mix left in one of his mixing pots (no. 5). He mentioned that he had left some explosive mix in the pot to another operator in the room, who later confirmed this by visual inspection. This other worker, from having looked into the mixer, indicated that the depth of explosive mix left in pot no. 5 was about 100 mm, which matched his estimate of there being 25–50 of explosives in the mixer. The mixer blades extended about 50 mm down into the mix. Metallurgical analysis of mixer parts retrieved after the explosion provided further evidence supporting the conclusion that explosive material was left in pot no. 5, because the analysis showed that damage to the hub of the mixing blade was consistent with it having been in contact with detonating explosives.

An over-current protection device on the electrical mix motors in Booster Room 2 would stop the motor if the blade was unable to break up the explosives, but not before the maximum start-up torque had been applied to the explosives. Without continuous agitation, the different explosives and binders of the mix tended to stratify due to their different densities. This stratification would increase the initiation sensitivity of portions of the explosive material left in the pot. Turning on an agitator immersed 50 mm into a solidified mass of stratified explosives would present a high risk of explosion from the impact of the mixer blades.

Leaving material in the mixing pots overnight was not according to common practice in the plant, but this practice was not for safety reasons. Several months before the explosion, when material had been left overnight in the Comp B mixing pot in Booster Room 1, management made it clear that this was an unacceptable practice because it delayed the operation of the day shift workers.

The operator who left the material in his pot had been working in Booster Room 2 for eight weeks prior to the incident. His normal practice was to leave both of his mixing pots empty in agreement with common practice. Because he was the only person working his production line, he would normally know whether his pot was empty when he started work the next day. Since the pots in Booster Room 2 heated material much faster than the pots in Booster Room 1 could do, it is possible that he on the day before the explosion thought that leaving material in the pot would not delay production the next day. It was acceptable practice at the Kean Canyon facility to alter normal processes without discussion or management approval.

At the end of each day, operators were instructed to leave a steam line valve to each pot partially open to keep the boiler cycling, to prevent freezing of condensate in the lines. This amount of steam would be insufficient, however, to maintain any quantity of explosive mix in the pots above its melting point if outside temperatures were below the freezing point. During the night before the explosion, the outside temperature dropped to between –4 and –7°C. The operator who had left explosives in his pot offered the remaining material to the operator on the other production line in Booster Room 2. Because the operator who was leaving did not reach a firm agreement on whether the second operator would use the residual explosives, it is possible that no steam valves were left open that afternoon because leaving the valves open would make the remaining base mix too runny to pour. The operator who left early may have mistakenly thought that his remaining base mix would be used that afternoon and, thus, he failed to look in the pot the next morning before turning on the steam and mixer motor. The CSB investigation team concluded that this was the most likely scenario.

6.4.3.5 Key Findings and Root/Contributing Causes

- U.S. standards require that companies using highly hazardous materials have in place an integrated safety management system. CSB's investigation of the Kean Canyon explosion catastrophe revealed that many essential elements of adequate process safety management were missing or deficient. The process hazard analysis (PHA) of the facility and operations was inadequate, and supervisors and workers from the plant had not been involved. The PHA for Booster Room 1 was conducted by company personnel from other locations and did neither consider safe location of buildings nor human factors issues. These deficiencies in the PHA program allowed unsafe conditions and practices to remain unrecognized and uncorrected. No PHA was conducted for Booster Room 2.
- There was no safety inspection or safety auditing program. Safety "walkthrough" inspections were unfocused and did not examine process safety management (PSM) program effectiveness. The result was that management was generally unaware of unsafe practices and conditions.
- The training programs for personnel working in the plant were inadequate. Worker training was conducted primarily in an

ineffective, informal manner that over-relied on the use of on-the-job training. Poor management and worker training led to a lack of knowledge of the hazards involved in manufacturing explosives.

- Written operating procedures were either non-existing or not available to the working personnel. Managers believed that, without using a blasting cap, it was almost impossible to detonate the explosive materials they used or produced. Personnel primarily relied on experience to perform their jobs. Procedures and other safety information were not available in the language spoken by most workers. Operators routinely made changes in the steps they took in manufacturing explosives. This resulted in the use of inconsistent and hazardous work practices. There were no written procedures for Booster Room 2.
- The whole facility was built with insufficient separation distances between different operations, and the construction of buildings was inadequate. Because unrelated chemical operations were located in the same building as Booster Room 2, one additional fatality and additional property damage resulted. Close proximity of structures allowed the explosion to spread to a second building.
- Safety inspections by regulatory organizations were conducted infrequently and inspectors generally did not have expertise in explosives manufacturing safety. This allowed unsafe conditions in the plant to remain uncorrected.
- Employees had not been involved in developing or conducting process safety activities. This resulted in a lack of understanding of process hazards and controls by workers. It also resulted in management not benefiting from the experience and insights of workers. Employees had not been involved in developing or conducting process safety activities. This resulted in a lack of understanding of process hazards and controls by workers. It also resulted in management not benefiting from the experience and insights of workers.
- Reclaimed, demilitarized explosive materials purchased by Sierra sometimes contained foreign objects. The risk of using contaminated explosive materials had not been adequately examined.

See Section 6.5.2 for CSB's recommendations for improvements.

6.4.4 Pyrotechnics Disaster in Enschede, Netherlands (2000)

6.4.4.1 Introduction

The following short summary is based on the accounts by Bruyn et al. (2002) and Weerheijm et al. (2002), which are in turn based on the results of the extensive forensic and technical investigations of this catastrophic accidental fire/explosion undertaken by Netherlands Forensic Science Institute (NFI) and TNO Prins Maurits Laboratory (TNO-PML).

6.4.4.2 Overall Development and Consequences of the Disaster

The disastrous explosion occurred on Saturday afternoon, 13 May 2000, on the premises of the company S.E. Fireworks (SEF) in the city of Enschede in the Netherlands. SEF imported and traded with fireworks, and also performed firework displays and shows. Within one hour, the incident developed from a moderate fire and some initiated fireworks in one of the buildings, into a series of three explosions of increasing violence. The first was a relatively small explosion in a container. Within a minute, seven garage boxes (prefabricated concrete storage facilities) exploded. This was followed about one minute later by a further explosion of storage cells in the central storage building, whereby the most violent explosion occurred in storage cell C11. The resulting blast wave was comparable to that from explosion of 4,000–5,000 kg TNT. Most of the evidence for quantifying the final explosion strength was obtained from the window breakage and the observed damage of about 700 houses in the surroundings.

This disaster killed twenty-two people and injured 947. A complete residential area was destroyed, with about 500 houses being completely demolished and about 1,350 partly damaged. The main cause of house damage was massive conflagration of typical old houses with wooden floors, initiated by the throw out of fireworks.

6.4.4.3 The Situation Prior to the Fire/Explosion

The layout of the pyrotechnics storage facility prior to the fire/explosion is illustrated in Figure 6–11.

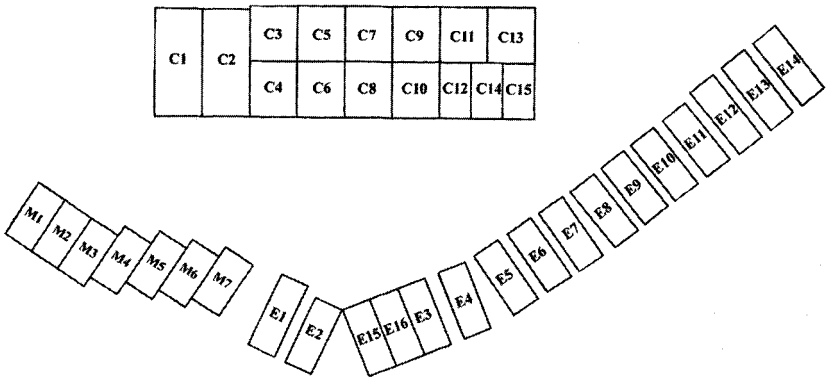


Figure 6–11 Layout of the pyrotechnics storage premises of the company S.E. Fireworks (SEF) in the city of Enschede in the Netherlands in 2000. Based on figure from Weerheijm et al. (2002).

The facility consisted of a central storage building containing 14 storage cells (C2-C15), 7 prefabricated garage boxes (M1-M7), and 14 ISO containers (E1-E14).

The central storage building had wooden doors and was constructed in cast, reinforced concrete with a wall and roof thickness of 20 cm. Cell 2 was used for fireworks preparation and workshops. The majority of the cells in the central building (C3–C11 and C13) were 4 m wide, 4 m deep and 2.8 m high, whereas three smaller cells (C12, C14 and C15) were only 2.5 m wide. The walls and roof of the prefabricated concrete boxes had a thickness of 5 cm. 6 of the prefabricated garage boxes (M1–M6) were 2.6 m wide, 5.2 m deep and 2.3 m high (volume 31 m^3), and fitted with the original thin, corrugated steel sheet doors. M7 was slightly larger and had a wooden door. In order to prevent direct flame contact between cell doors in case of fire, the walls in the central building and the garage boxes were extended externally by 0.5 m. The ISO containers were standard 6 m long (20 ft) transport containers, with wooden floors. They were not equipped with any additional fire protective measures.

The total licensed quantities for storage were 158,500 kg (gross mass) of 1.4S or 1.4G fireworks. In some cells, 3G fireworks could be stored, the maximum permissible stored quantity for the whole facility being 2000 kg.

6.4.4.4 Global Sequence of Events

Due to the fact that many people were attracted to the accident, several private video recordings taken from various angles were made available to the investigators. These were helpful in the reconstruction of the disaster. The reconstructed sequence and times of the main events were:

1445: Effects of fire in C2 first observed (ejection of fireworks)

1500: Fire reported to fire brigade

1508: Preliminary inspection of the site by the fire brigade

1516: Fire in C2 under control

1524: Smoke and fire effects from C4 observed

1533: Smoke from between containers E1 and E2 visible

1534: Massive deflagration and minor explosion of contents in container E2

1540: Massive explosion of garage storage boxes M1–M7

1546: Explosion in C11 in the central building, followed almost simultaneously by explosions in the other cells in the main building, and in a number of the transport containers (E)

Figure 6–12 shows the fire ball from the final explosion from a distance of about 600 m, whereas Figure 6–13 is a top view of the completely demolished explosion area, with search sections taped out for the forensic investigation.

6.4.4.5 Initial Fire in Cell C2

The fire in C2 was first noticed at about 1445 hours when activated firework was ejected and landed outside the SEF premises. A small fire in a garden was reported. When the fire brigade arrived, the doors of cell C2 were open at both sides and the plastic skylights were gone. Some fast pressure build-up must have occurred because the doors on both sides were blown out. No external blast damage was noticed. Possible mechanisms for initiation of the fire/explosion in C2 include:

- ejected fireworks
- sabotage, arson or improvised explosive devices

Figure 6–12 Photograph, taken at a distance of 600 m, of fireball created by the final major explosion on the pyrotechnics storage premises of the company S.E. Fireworks (SEF) in the city of Enschede in the Netherlands, in May 2000. From Bruyn et al. (2002).

Figure 6–13 Top view photograph of the explosion area after the final major explosion on the pyrotechnics storage premises of the company S.E. Fireworks (SEF) in the city of Enschede in the Netherlands, in May 2000. From Bruyn et al. (2002).

- malfunctioning or defects of the electrical and gas installation or other equipment
- fire caused by human activities on the SEF area
- (self) ignition of firework articles, pyrotechnics or other fuels caused by instability or external effects

The effects of the fire in C2 were:

- ejected fireworks, with the possibility of fire initiation outside the main building
- fire jets and heat radiation directed towards the garage boxes (M) and containers (E)
- fire heating of the walls to adjacent C-cells

The fire in C2 was fought from both sides of the central building. However, the firemen on the side facing the garage boxes (M) were killed in the subsequent explosion, and hence no direct witness reports of the fire development on that side of the building were available.

6.4.4.6 Initiation and Possible Consequences Fire/Explosion in E2

The following possible initiation mechanisms for the fire/explosion in E2 were examined:

- external fire
- fireworks on and, or under the container
- burning magnesium (ejected from the workshop) on top of container
- fireworks before door slit

Considering the time frame, the very limited fire resistance of the container, the presence of an old small trailer with wooden floor between E1 and E2, a small fire would be sufficient to initiate a fire and activate the fireworks in E2. The effects following ignition in E2 were:

- smoke from door slit
- strong smoke development followed by intensive firework reactions

- flame jet (in two pulses)
- ejected fireworks and very severe external massive reaction of fireworks

For the possible consequences of the E2 reaction for the further escalation of the disaster the following effects were examined:

- ejection of debris
- failure of container (fragments and blast)
- flame jet (and possible external fireball)
- ejected fireworks

6.4.4.7 Initiation and Possible Consequences of Fires/Explosions in the Garage Boxes (M)

The damage analysis strongly indicates that the most severe explosion of those in then garage boxes occurred in M7. The observed debris velocity of up to 200 m/s was most probably caused by the failure of the 5 cm thick walls of the prefabricated garage box M7 The resulting severe crushing of the fireworks in M6 and the thermal loading caused the exothermal chain reaction of the fireworks in M6, and subsequently in the other garage boxes. The reactions in all of the garage boxes occurred within less than 0.5 seconds. The resulting explosion blast was comparable to that of the explosion of 800 kg of TNT. A fireball of diameter 170 m was generated. The garage boxes were completely destroyed and fragmented into small debris. The combined blast, flame and debris loading was sufficient to initiate the final major explosions in the central building and the remaining E containers.

6.4.4.8 Initiation and Explosion Effects of Explosion in Central Storage Building

Considering the central building, the strength of the explosion was by far sufficient to blow the wooden doors into the cells whereby the fireball engulfed the whole building. The contents of all cells could have been ignited. The local damage, however show clearly that the explosion in storage cell C11 was most severe and dominant. A single explosion in C11 and the sequential sympathetic reactions in the other cells can explain the total damage. Crucial in the explanation is the reaction velocity of the

fireworks in C11. Hypotheses to explain the devastating mass explosion in C11 are:

- storage of firework of the transport class 1.1
- combined storage of 1.3G and 1.1 fireworks. Fireworks of the class 1.3G were stored, but due to door impact the packages were severely damaged and the firework obtained the 1.1 characteristics
- after initiation of the stored 1.3G class fireworks, temperature and confinement conditions accelerated the deflagration process towards a detonation-like reaction flame jet and possible external fireball

6.4.4.9 Observed Explosion Effects

The major explosion effects were craters, fireballs, blasts, and debris. The video recordings and the damage at the explosion area showed that the sequence of three main explosions, i.e. in container E2, in the garage boxes (M1–M7) and the central building (C-cells), had successively increasing strengths. For this reason the final disastrous explosion destroyed evidence and traces of the preceding explosions, which hampered detailed analysis. Nevertheless, the following conclusions could be drawn from remaining evidence.

No crater or evidence for blast damage due to the E2 explosion was found. A very severe fireworks fire and the ejection of firework articles from the container into the surroundings characterized the "explosion" in E2. The observed effects correspond to fireworks of transport classification 1.3G.

As regards the second main explosion in the garage boxes (M1–M7 in Section 6–11), video recordings showed that the fireball of the explosion increased to a diameter of about 85 m in 0.3 seconds. The explosion appeared to be due to an exothermal reaction of the contents of the garage boxes, starting in M7 and terminating in M1. The boxes were completely destroyed, and no remaining debris could be recollected. However, the video recordings showed debris being ejected at velocities of the order of 200 m/s. The exothermal reactions were violent, but a detonation did definitely not occur. The concrete floors in the garage boxes were severely cracked, and the floor slab of M7 was moved more than 1 m horizontally, and a large depression of the soil at the original location was found. However, no crater was formed. The walls of the boxes were clearly sheared

off and the direction of the deformed reinforcement bars formed clear evidence for the propagation direction of the chemical reactions in the cells (from M7 towards M1). In general the blast strength of an explosion can be quantified from the building damage in the surroundings and especially from the window breakage. However, in this case the final explosion destroyed most of this evidence as regards the explosions in the garage containers. At one video recording of the second of these explosions, however, window and roof tile damage is visible. The radius of the observed fireball corresponded to exothermal reaction of about 17,000 kg of propellant. Further valuable information about the strength of the explosion in the garage boxes was obtained from the seismic signals that were recorded of both of the two last major explosions. The ratio of the signals, the local damage and the blast damage to the surrounding area suggested that the explosion in the garage containers produced a blast wave of the same strength as that produced by about 800 kg TNT.

The strength of the explosion in the garage boxes was by far sufficient to blow the wooden doors into the C-cells of the central building. Furthermore, the fireball engulfed the whole central building, and the contents of all the C-cells were probably ignited. The central building was completely destroyed. Violent exothermal reactions occurred in many of the cells in the central building, but the final, major explosion was clearly dominated by the reaction in C11. This emerges from the facts that a large crater of depth 1.3 m was formed just under cell C11. Although the crater extended to the areas below some other C-cells, it was clearly the result of a single major explosion event. The floors of the other cells showed no damage of the type to be expected from independent, severe explosion reactions. Only the edges of the cell floors adjacent to C11 were severely damaged and contributed to the crater.

Debris from the central building caused many casualties and severe damage to surrounding buildings. Debris was found up to a distance of 580 m from the explosion centre. From the angle of impact and the traveling distance of the largest debris, ejection velocities could be estimated. Because the angle of ejection was unknown, only an approximate range of ejection velocities could be determined. For most of the collected debris the estimated velocities were in the range 30–100 m/s. The highest estimated velocity was 150 m/s.

The diameter of the final major fireball was 135 m, corresponding to exothermal reaction of about 86,500 kg of pyrotechnics/propellants. It is evident that this fireball was not caused by the contents of cell C11 only.

Exothermal reactions of the pyrotechnic material stored in other cells and also the storage containers must have contributed. On the video recordings the shape of the fireball shows quite clearly some additional "sub sources" which may be caused by container reactions.

6.4.4.10 Conclusions

- In spite of extensive investigation, no definite evidence for the initial cause of the chain of events had been found by the time of the publication of the papers by Bruyn et al. (2002) and Weerheijm et al. (2002). There was no indication of sabotage, and no traces of high explosives had been detected, only of fireworks-related substances.
- In addition to the initial fire in the central building, three other crucial events promoted the escalation of the accident. These were the severe firework reactions in container E2, followed by the explosions in the garage boxes (M) and finally by the massive explosion in the storage cell C11 of the central building, accompanied by the reactions in the other storage cells and containers.
- Window breakage, structural damage, crater dimensions, debris and the seismic signals enabled quantification of two major explosions. The first occurred in the garage boxes and had a blast strength equivalent to about 800 kg TNT. The blast produced by the final, much more powerful explosion had a strength equivalent to about 4000–5000 kg TNT.
- In spite of the violent character of the explosions, they were all subsonic deflagrations. This is in accordance with the nature of the exothermal reactions being the cause of the explosions. No evidence of detonations was traced.
- The types of stored firework articles, the excessive quantities stored, and partly inadequate storage conditions, probably made it possible for the event to escalate from the initial moderate fire to the final disastrous explosion.
- Much larger quantities of some types of firework were stored than licensed, which contributed to making the storage unsafe. If the stored quantities had been in agreement with the premises of the licenses granted for storage, the fire in the workshop of the central storage building could never have escalated to the disaster that actually occurred.

- The minimal fire resistance of the storage containers, and inadequate safety distances between the various storage units, also contributed to the escalation.
- Weerheijm et al. (2002) concluded that further research would be needed to enable definite conclusions to be drawn on the origin and stepwise escalation of the catastrophic explosions in Enschede, and also for enabling specification of improved requirements for safe storage and transportation of the types of pyrotechnics that gave rise to the catastrophe.

6.5 Measures for Preventing and Mitigating Accidental Explosions of Explosives, Propellants, and Pyrotechnics

6.5.1 General Safety Considerations for Facilities Producing Explosives According To Sejlitz (1987)

6.5.1.1 Basic Rule

The planning of a process and/or a workplace should always be based on the following basic rule:

- a minimum number of persons for handling
- the smallest possible quantity of explosive material in
- the shortest possible time

6.5.1.2 Characteristics of the Explosive Material

Before conducting any work with explosive materials, all their basic safety properties must be known. These data makes it possible to evaluate both the various risks of ignition/initiation associated with handling and processing the material, and expected consequences of such ignition/initiation. The most important basic safety properties are (see Section 6.2 and Section 6.3):

- ignition/initiation sensitivity (e.g. impact, friction, heat, electric/ electrostatic sparks)
- chemical stability (e.g. possibility of exothermal decomposition during storage)
- effect of ignition/initiation (e.g. when the material is unconfined, enclosed, or in the processing vessels/containers
- health risks (e.g. acute and long-term exposure)

6.5.1.3 Explosion Protection

On the basis of the basic safety data of the material(s), one can decide whether the planned work

- can be performed in an open work room, or whether it
- must be performed under shielded conditions

The degree of rigidity of shielding depends on the severity of the effects of ignition/initiation of the material in question, and may take the form of

- simple protection screens inside the work room
- smaller, laboratory-scale bunkers inside the work room
- special full-scale bunkers (often with pressure resistant observation windows)

6.5.1.4 Automation and Remote Process Operation

In the case of dangerous operations it is often preferable to automate the process and run and control it remotely. However, this may also sometimes introduce additional risk elements, such as:

- Reduced operator attention may result in too late detection of operational errors.
- Automated, remotely operated processes may prevent the operator from acquiring the knowledge of the basic processes that he may need in the case or irregularities and errors.

- The possibilities of efficient interaction with the process in case of irregularities and errors are reduced.
- The work may appear less interesting and challenging to the workers.

6.5.1.5 Quantities of Explosive Material

As already mentioned, the quantities of explosive material being handled at any time at each workplace should be as small as practically feasible. The following principles should be taken into account:

- For a given production volume per day, the mass of material handled at any workstation at any time gets smaller if the transfer rate of explosive material through the process (mass per unit time) increases.
- A well-designed material flow system permits frequent transports of explosive material to and from the work rooms, which reduces the amount of explosive material stored in the work rooms at any time. This makes it possible to keep the buffer stores of explosive material in the work rooms at a minimum.

6.5.1.6 Avoidance of Accidental Ignition/Initiation

The equipment in which explosive materials are processed/treated/handled shall be designed to ensure that the materials are not exposed to:

- *Mechanical impact*

  Only approved equipment and tools shall be available at the work place. Experience has shown that if non-approved equipment such as hammers, rubber mallets, crowbars etc. has been available, operators have used this to solve problems that have arisen, and by doing so initiated explosion accidents.
- *Friction*

  Equipment and tools shall be designed/constructed in such a way that dust cannot enter threaded couplings, bearings etc. Furthermore, foreign bodies and impurities (hand tools, bolts and nuts, gravel from transportation boxes etc.) must be prevented from entering

apparatuses/equipment containing explosive material, thereby giving rise to explosion initiation by friction.

- *Exothermal chemical decomposition*

  Accidental explosions have occurred in cases where explosive materials that have first been made chemically unstable by impurities or heating, have been enclosed in e.g. a ball valve or between the walls of a double-wall pipe. Whenever, such possibilities exist, the equipment/apparatus must be controlled and cleaned regularly.

- *Adiabatic compression*

  This can occur during pressing and other treatment of gun powder, if the air in the pressing tool has not been evacuated before pressing, or if the air trapped in the gun powder is being subjected to a pressure rise by other means.

- *Heating/fire*

  All equipment/apparatus where explosive materials are heated electrically shall be equipped with several independent protection systems against over-heating. Hot work (welding etc.) always present a significant fire/explosion risk factor if carried out on premises where the rooms and the equipment/apparatuses have not been adequately cleaned before the hot work is initiated.

- *Electrostatic discharges*

  Static electricity is always a significant risk factor when producing/ treating/handling very sensitive explosive materials. The possibility of ignition/initiation by electrostatic sparks (discharges between two electrically conducting electrodes) can be effectively eliminated by connecting all conducting components/apparatuses/equipment/tools, as well as the operator, to earth. For the operator, using electrically conducting shoes is only effective if the floor is also conducting and connected to earth. If this is not the case, the operator must be equipped with a separate cable to earth. If the material can be ignited/ initiated by brush discharges, special precautions must be taken to avoid such discharges.

6.5.1.7 Good Practices for Apparatus Design

Quite often accidental initiation/ignition of explosive materials occur because they become exposed to excessive mechanical forces in contact

with parts of apparatuses or tools, or to heat. Such accidents can occur even under normal operating conditions, but they are more common under circumstances deviating from normal operation, e.g. start-up, shut-down, cleaning, maintenance/overhaul, repair, and irregularities in the performance of equipment/apparatus (see Section 6.4.3.4).

It is important to consider the possibilities of irregular performance of equipment/apparatus, and the ease by which it can be cleaned, maintained, overhauled and repaired, at the earliest stage of selecting/designing new equipment/apparatus. Careful consideration should be given to the following:

- *Selection of construction materials*

  The construction materials used in equipment/apparatus must not be able to react chemically with the explosive materials produced/treated/handled in the equipment apparatus. They should be able to withstand mechanical stress and wear due to rubbing etc. On the other hand, the use of mechanically strong materials may increase the risk of ignition/initiation by impact, rubbing etc., and in such cases softer and less mechanically strong construction materials will have to be used. However, all types of construction materials can represent an ignition/initiation risk if dimensioned/used in a wrong way. There are no materials that are either fully adequate or fully inadequate in any kind of practical situation.

- *Free space around moving parts of equipment/apparatus*

  Hazardous frictional stresses may be generated if an explosive material gets squeezed between moving parts of equipment/apparatus and another object, e.g a tool. In order to prevent other objects from creating such situations, the free space around moving parts should either be so large that the introduction of another object will not be able to create hazardous stresses, or so small that the another object will not be able to enter the space.

- *Design and construction of equipment/apparatus*

  It is essential that the design and construction of equipment/apparatus facilitates easy cleaning. It is a good working principle that equipment/apparatus should be designed and constructed in such a way that complete cleaning or inspection can be performed daily in a safe and efficient way. This requires that:

- all surfaces are smooth, preferably honed or polished
- there are no narrow gaps, grooves, holes or cracks in which explosive material can escape the cleaning operation
- there are easy-to-handle and well-dimensioned couplings and other fixing elements, and that bolts and nuts are secured in an adequate way
- pipelines are easy to dismantle and inspect
- there are effective arrangements and procedures to prevent spill of explosive material, and accumulation of dust from such materials

6.5.1.8 Written Work Instructions

In addition to a good process design it is important that the operator has the knowledge, skills and motivation to run the process in a safe way. Hence, well-written, easy-to-understand work instructions shall always be readily available. These may include information on the following aspects:

- properties of the explosive material(s)
- the work operations to be undertaken
- risks associated with the work
- guidelines/regulations to ensure safe work procedures
- measures to be taken in case of deviation from normal material and process conditions
- routines for cleaning/housekeeping
- control and maintenance of production equipment
- special work instructions for more comprehensive and difficult installation and maintenance work

6.5.1.9 Ten Mnemonic Rules

(1) Work with the smallest practically possible quantity of explosive material at any time.

(2) Be well informed about methods of work and safety instructions.

(3) Make sure that there are adequate escape routes.

(4) Use only approved tools and other apparatus/equipment.

(5) Keep the work place in order. Materials, tools and apparatus/ equipment, which are not used for the time being shall not be stored in the work rooms.

(6) Give notice of and/or correct any errors or undesired deviations from normality.

(7) Make sure that all explosive materials that are not being processed are well packed and clearly marked.

(8) Make sure that all trash/refuse explosive materials are well packed and clearly marked.

(9) Never conduct production work with explosive material in a store room for explosive materials.

(10) Keep the work place clean and tidy at any time. Spill of explosive material should be removed immediately (by gentle sweeping).

6.5.2 General Safety Considerations for Facilities Producing Explosives According to CSB (1998)

Partly based on the experience gained from investigating the Sierra's Kean Canyon plant explosion catastrophe in 1998 (see Section 6.4.3), CSB (1998) recommended that the following aspects be considered:

- Adequate process safety management (PSM) in explosives manufacturing facilities requires both careful planning and careful implementation.
- Prevention of primary explosions, as well as further propagation of explosions, requires clear understanding of explosives safety principles and safe practices.

- Process Hazard Analyses (PHA) shall include examination of safety-distance requirements, building design, human factors, past incident reports, and lessons learned from explosives manufacturers.
- Written operating procedures shall be specific to the actual process of concern and address all phases of the operation.
- Procedures, chemical hazards, and process safety information shall be communicated in the language(s) understood by personnel involved in manufacturing or handling of explosives.
- Explosives training and certification programs for workers and line managers shall provide and require demonstration of a basic understanding of explosives safety principles and job-specific knowledge.
- Pre start-up safety reviews shall be performed to verify operational readiness whenever changes/modifications have been made.
- Adequate implementation of all elements of mandatory official process safety management standards shall be verified by performing periodic assessments and audits of safety programs.
- The employee-participation program shall effectively include workers and resolve their safety issues.
- Explosives safety programs shall provide an understanding of the hazards and control of detonation initiation sources. These include:
    - foreign objects in raw materials
    - impact by tools or equipment
    - impingement
    - friction
    - sparking
    - other electrostatic discharges

Specific handling requirements for raw materials shall be followed strictly. Substitute raw materials may introduce additional ignition risks requiring special concern.

CSB (1998) further suggested that the following issues be addressed in plant design or modification:

- Unrelated chemical or industrial operations or facilities shall be separated from explosives facilities using quantity distance guidelines.
- Operations in explosives manufacturing plants shall be separated by adequate intra-plant distances to reduce the risk of propagation.
- Process changes, such as the construction or modification of buildings, or changes in explosive ingredients, equipment, or procedures shall be analyzed and PSM elements shall be updated to address these changes.
- Facilities shall be designed to reduce secondary fragmentation that could result in the propagation of explosions.
- Process and safety training guidelines for personnel involved in the manufacture of explosives shall be developed and disseminated. Such guidelines shall include methods for the demonstration and maintenance of proficiency.
- Safety guidelines for screening of reclaimed explosive materials shall be developed and disseminated.
- Adequate frequencies of safety inspections of explosives manufacturing facilities shall be maintained due to their potential for catastrophic accidental explosions.

Chapter 7

Design of Electrical Apparatuses for Hazardous Areas

7.1 Classification of Hazardous Areas

7.1.1 The Overall Objective of Area Classification

The overall objective of area classification is to minimize the probability of accidental ignition of explosive atmospheres in e.g. the process industries. The area classification forms the traditional basis for design of electrical apparatus. The basic philosophy is that more strict requirements have to be enforced to the design of electrical apparatuses to be used in areas where the probability of occurrence of explosive atmospheres is high, than to equipment to be used in areas where this probability is low. The prime objective of area classification is to assign different "zone" classes to the various parts of the industrial plant of concern according to the likelihood that explosive atmospheres may occur.

The area classification philosophy focuses on minimizing the probability of ignition. This is not the same as to minimize the explosion risk. The difference is that the explosion risk concept considers not only the probability of ignition, but also the consequence of an ignition. Explosion risk may then be defined as the product of the probability of ignition and the expected consequence of an ignition. In other words, a minute ignition probability combined with a catastrophic explosion resulting from an ignition, can easily give rise to an unacceptably high explosion risk. A very brief outline of some aspects of comprehensive risk analysis is given in Chapter 8.

7.1.2 Gases and Vapors

7.1.2.1 Definition of Area Classification

The area classification approach has traditionally been used as a basis for specifying requirements for electrical equipment for use in potentially explosive atmospheres. According to The Institute of Petroleum (2002) area classification is the assessed division of a facility into hazardous areas and non-hazardous areas, and the subdivision of the hazardous areas into zones. A hazardous area is defined as a three-dimensional space in which a flammable atmosphere may be expected to be present at such frequencies as to require special precautions for the control of potential ignition sources including fixed electrical equipment. All other areas are non-hazardous in this context, though they may, in part or whole, form part of a wider restricted area within the facility in which all work is carried out under special controls. Examples include petroleum distribution installations and offshore production installations.

The International Electrotechnical Commission (IEC) defines a hazardous area as an area within which an explosive atmosphere is present, or may be expected to be present, in quantities such as to require special precautions for the construction, installation and use of (electrical) apparatus. Area classification is the effort to divide a process area into so-called "zones" characterized by different probability of occurrence of explosive atmospheres. This sub-division is to be documented in maps in all three elevations, showing the extent of the various zones. The zone definitions for gases and vapors are given in Section 7.1.2.7 below.

7.1.2.2 Purpose of Area Classification

The purpose of area classification is to avoid ignition of releases, intentional as well accidental, that may occur in the operation of facilities handling flammable gases, liquids, and vapors. The approach is to reduce to an acceptable minimum level the probability of coincidence of a flammable atmosphere and a source of ignition.

It is not the aim of area classification to guard against the ignition of major catastrophic releases of flammable gases or vapors, e.g. due to rupture of a pressure vessel or a pipeline. The philosophy is that in properly run facilities, such events have a very low probability of occurrence, and

that the likelihood of their occurrence must be kept below acceptable limits by correct design, construction, maintenance, and operation of facilities.

7.1.2.3 Limitations of Area Classification

As mentioned above, the traditional area classification system was developed at a time when more comprehensive risk analysis methods were not available, and does neither address catastrophic failure, nor the consequences of explosions. It is now realized that conventional area classification is not a sufficiently powerful tool to serve as a basis for estimating explosion risks. The absence of acceptance criteria, in terms of maximum acceptable risk (frequency • consequence), is a major problem in applying conventional area classification as the sole criterion for selection of proper equipment. In comprehensive risk analyses, both the probabilities of occurrence of explosive atmospheres and the possible effects of explosion of such atmospheres are quantified

In future, conventional area classification procedures may be supplemented or even replaced by thorough risk assessment procedures, which also take into account the effects of explosions and fires and the resulting quantitative consequences. If the conclusion of this kind of analysis is that the explosion/fire risk is unacceptable, then appropriate measures for reduction of the risk must be taken. This may be accomplished by one or a combination of several of the measures indicated in Figure 2–59. Careful consideration must be given both to technical risks during normal operating conditions and periods of limited technical malfunction during normal operation and also to risks during maintenance and repair work. Possible human errors must also be accounted for. In the context of preventing the occurrence of effective ignition sources, risks associated with the use of tools in maintenance and repair work are of prime concern. In any case, the overall aim must be to ensure the required level of safety by means of cost-effective solutions.

7.1.2.4 Small Scale Operations

Certain locations handling only small quantities of flammable fluids can, in the context of area classification, be classified as "non-hazardous." For example, this may apply to laboratories for testing small petroleum fluid samples. It is not possible to set a cut-off point, as this must be judged according to the circumstances. For instance, when draining gasoline from

a vehicle fuel tank in an enclosed garage or a below-ground inspection pit, due precautions must be taken to prevent ignition. Such areas must be classified as Zone 1, and only Zone 1 (or Zone 0) electrical equipment should be permitted (see section 2.1.2.7). In addition due precautions must be taken to prevent ignition from any other type of ignition source. In making such a judgment, the risk to people should be assessed.

Each vessel containing flammable fluids should be treated individually, by considering the nature of its surroundings and the extent to which people need to be present. As a rough guide, hazardous area classification may not be needed if the maximum amounts of material that could be released are below the quantities given in Table 7–1.

Table 7–1 Capacity Thresholds of Process Equipment Located inside and outside Buildings Below which Area Classification May Not be Required From The Institute of Petroleum (2002)

	Gas (Volume corrected to 1 bar(abs) pressure	Liquefied Flammable Gas	Flammable liquid at a temperature above its flash point
Inside	50 litres	5 litres	25 litres
Outside	1000 litres	100 litres	200 litres

However, in certain circumstances, ignition of quite small quantities of flammable gas/vapor mixed with air can cause danger to persons in the immediate vicinity. Where this is the case, as in a relatively confined location from which rapid escape would be difficult, area classification may be needed down to quite small quantities of fluid.

7.1.2.5 Good Standards of Design and Operation

The area classification technique described here assumes that the facilities to which it is applied are designed, constructed, maintained and operated in accordance with good industry practice so as to reduce releases to a minimum. Equipment and piping should be designed to international standards or national equivalents. The recommendations of the IP Model Codes (see The Institute of Petroleum, 2002) or their equivalents, regarding good operational and maintenance practice, should also be followed in accordance with good industry practice so as to reduce releases

to a minimum. Equipment and piping should be designed to international standards or national equivalents.

7.1.2.6 Management of Area Classification

As a rule area classification should be carried out before the design and layout of equipment are finalized. At this stage, it may be possible to make considerable improvements at little cost. The area classification should always be reviewed and drawings modified, if necessary, on completion of design and before any change is made to existing plants handling flammable fluids.

Area classification should be incorporated into the company's Health, Safety, and Environmental Management System. The person responsible for the coordination of the area classification should be identified and be competent in this field. The work, which requires an interdisciplinary approach, should be carried out by persons who have full knowledge of the process systems and equipment, in consultation with safety, loss prevention, and electrical engineering personnel, as appropriate. Agreements reached on the area classification should be formally recorded and continually reviewed and updated. Records, such as drawings and/or tabulated data sheets, should include details as to the type of protection selected to meet the zone requirements and the apparatus sub-group and temperature class.

In principle, the classification of an area entails consideration of all the actual and potential sources of release of flammable fluid present. In practice, the procedure can be simplified by adopting a standardized area classification diagram. In other cases, a procedure of considering individual point sources will be required. A detailed procedure for this is given by The Institute of Petroleum (2002).

7.1.2.7 Definition of Zones

Areas are subdivided into zones based on the likelihood of occurrence and duration of a flammable atmosphere, as follows:

- Zone 0

  That part of a hazardous area in which a flammable atmosphere is continuously present or present for long periods.

- Zone 1

  That part of a hazardous area in which a flammable atmosphere is likely to occur in normal operation.

- Zone 2

  That part of a hazardous area in which a flammable atmosphere is not likely to occur in normal operation and, if it occurs, will exist only for a short period.

- Non-hazardous areas

  Areas that do not fall into any of the above zones are non-hazardous. A Zone 1 area will often be surrounded by a larger Zone 2 area, but there is no specific requirement for this. However, whenever a Zone 1 area is not part of a larger Zone 2 then the possibility of any large but infrequent release, which would require a larger Zone 2 area, should be considered.

7.1.2.8 Sources and Grades of Release

For the purpose of area classification a point source of release is defined as a point from which a flammable gas, vapor or liquid may be released into the atmosphere. The following three grades of release are defined in terms of their likely frequency and duration.

7.1.2.8.1 Continuous Grade Release

A release that is continuous or nearly so, or that occurs frequently for short periods.

- Primary Grade Release

  A release that is likely to occur periodically or occasionally in normal operation i.e. a release which, in operating procedures, is anticipated to occur.

- Secondary Grade Release

  A release that is unlikely to occur in normal operation and, in any event, will occur only infrequently and for short periods i.e. a release which, in operating procedures, is not anticipated to occur.

The grade of release is dependent solely on the frequency and duration of the release. It is completely independent of the rate and quantity of the

release, the degree of ventilation, or the characteristics of the fluid, although these factors determine the extent of gas/vapor travel and in consequence the dimensional limits of the hazardous zone.

To assist understanding of the boundaries of the definitions of the different grades of release, the following quantities are suggested. A release should be regarded as continuous grade if it is likely to be present for more than 1,000 hours per year and primary grade if it is likely to be present for between 10 and 1,000 hours per year. A release likely to be present for less than 10 hours per year and for short periods only should be regarded as secondary grade. This assessment should take account of any likelihood of leaks remaining undetected. Where releases are likely to be present for less than 10 hours per year but are anticipated in normal operation (e.g. routine sampling points), they should be regarded as primary grade releases.

The allocation of the grade of release should be reviewed in the course of the design stages to determine if practicable and economical design or engineering improvements can be made to reduce the number of continuous and primary grade releases. Assessment of the grade of release is not always obvious and will require experienced engineering and operational judgment. Releases that occur regularly but with short duration should generally be classified as primary grade sources giving rise to a Zone 1 area.

7.1.2.9 Relationship Between Grade of Release and Class of Zone

Under unrestricted 'open air' conditions there is, in most cases, a direct relationship between the grade of release and the type of zone to which it gives rise; i.e.

- continuous grade normally leads to Zone 0
- primary grade normally leads to Zone 1
- secondary grade normally leads to Zone 2

However, this may not always be true. For example, poor ventilation may result in a more stringent zone while, with generous ventilation, the opposite will be true. Also some sources may be considered to have a dual grade of release with a small continuous or primary grade and a larger secondary grade. Examples include vents with dual-purpose process requirement, and pump seals.

7.1.2.10 Area Classification Drawings

Area classification records can comprise detailed drawings with notes and/or can be in the form of tabulations. The area classification drawings should be in sufficient scale to show all the main items of equipment and all the buildings in both plan and elevation. The positions of all openings such as doors, windows and ventilation inlets and outlets, and utility entries if not scaled gas/vapor-tight should be included as the careful positioning of these openings can affect the sizing of related external hazard zones.

Area classification drawings should be marked up to show the boundaries of all hazardous areas and zones present using the shading convention adopted internationally and shown in Figure 7–1. It is acceptable to indicate any requirement for small local zones/areas, e.g. around pumps and control valves, in a note on the drawing.

The final area classification should include a record of all additional supporting details. It is necessary to clearly distinguish regions on the drawing where different gas properties prevail, e.g. hydrogen with a Gas Group IIC on part of a drawing where mainly hydrocarbons are present. This may be illustrated using half-width hatching for the "hydrogen" region.

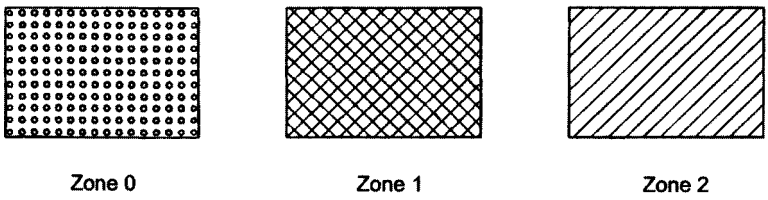


Figure 7–1 Shading convention for area classification drawings of areas where explosive clouds of gases and vapors can occur. From The Institute of Petroleum (2002).

7.1.2.11 Combustible-Fluid Properties and Area Classification

7.1.2.11.1 Class I Fluids

For definitions of fluid classes, see Section 2.1.3.3 in Chapter 2. Class I fluids have very low flash points, and will always produce a vapor in the

flammable range in air even at temperatures far below ambient. Facilities handling Class I fluids must always be area classified.

7.1.2.11.2 Class II and Class III Petroleum and the Distinction between Subdivisions (1) and (2)

Materials with flash points according to in Class II or III (see definitions in 2.1.3.3) will often be stored or handled at temperatures below their flash points, i.e. at temperatures where the vapor pressure is to low to give explosive fuel/air mixtures. In such cases, and where the possibility of releases in the form of a flammable mist or spray can also be excluded, the surrounding area does not have to be classified. However, where a material is held under pressure and there is a possibility of mist or spray formation on release, that material may produce a flammable atmosphere regardless of the storage conditions and flash point. In such circumstances, therefore, those materials classified as Class II(1) or III(1) (i.e. non-hazardous) should be classified as II(2) or III(2) respectively. In such cases it must be confirmed that the liquid temperature cannot be raised by any means in the event of release, e.g. by contact with a hot surface or proximity to an adjacent non-electrical source of ignition. As a typical example, within the UK mainland a maximum ambient temperature of 30 °C can be assumed, while typically offshore in the North Sea a maximum would be 24 °C. Aviation fuels of the kerosene type (flash point > 38 °C) may therefore, under these conditions, be classed as non-hazardous when stored away from processing areas and hot lines and vessels. Because of the greater variability in temperature in typical process areas Class II liquids should, however, be regarded as hazardous, i.e. Class II(2), in such areas. Special assessment may be required in climate regions of high normal ambient temperatures. When petroleum materials are to be stored or handled that are in the Class II(2) or III(2) condition, or are likely to be exposed to conditions of temperature above the flash point, the facilities should be classified as laid down for Class I.

7.1.2.11.3 Unclassified

Petroleum materials having a flash point above 100 °C can be given the subdivision accorded to Class III petroleum and should be regarded as Class III(2) when handled at or above their flash points. A distinction, however, should be made between distillates in this unclassified range

and products such as bitumens. When such materials are stored in heated fixed roof tankage (wherein the ullage space is essentially unventilated), the flash point as sampled or recorded will not be indicative of the presence or absence of a flammable atmosphere that may have accumulated in the ullage space. In common with the ullage space of fixed roof tanks or road or rail tank vehicles containing all classes of petroleum, the ullage space should be classified as Zone 0, with a 1,5 m Zone 1 surrounding vents and other roof openings. It should also be recognised in respect of heated petroleum of high flash point that although flash point and ignition temperature are different characteristics, at very high temperatures the effects can converge, i.e. the high temperatures, for example, of a hot surface can create a flammable condition locally which may be ignited by auto-ignition by that hot surface, as well as by an alternative ignition source.

7.1.2.11.4 Fluid Categories

In some situations and processes flammable fluids may have quite extreme and varied temperatures and pressures. In order to facilitate area classification by the point source method developed by The Institute of Petroleum (2002) in such cases, the concept of fluid categories has been introduced. This categorizes the fluids according to their potential for rapid production of flammable vapor when released to the atmosphere. The fluid categories used are defined in Table 7–2.

Table 7–2 Fluid Categories. From The Institute of Petroleum (2002)

Fluid Category	Description
A	A flammable liquid that, on release, would vapourize rapidly and substantially. This includes: (a) Any liquefied petroleum gas or lighter flammable liquid. (b) Any flammable liquid at a temperature sufficient to produce, on release, more than about 40% vol. vapourization with no heat input other than from the surroundings.
B	A flammable liquid, not of Category A, but at a temperature sufficient for boiling to occur on release.
C	A flammable liquid, not of Categories A or B, but which can, on release, be at a temperature above its flash point, or form a flammable mist or spray.
G(i)	A typical methane-rich natural gas.
G(U)	Refinery hydrogen.

7.1.2.11.5 Simple Relationship Between Petroleum Class and Fluid Category

For simple situations Table 7–3 may be used to convert petroleum classification to fluid category. It will be noted that the degree of vaporization that will occur on release to the atmosphere reduces in going from Category A to Category C, i.e. Category C is the least volatile.

Table 7–3 Relationship between Petroleum Classification and Fluid Category. From The Institute of Petroleum (2002)

Classification of petroleum by the based (except for LPG) on closed cup flash points			Fluid Category		
Class	Description	Handled above flash point	Handled above boiling point	Can be released as mist	Handled below boiling point and cannot be released as mist
Class 0	Liquefied petroleum gases (LPG)	Yes	A	A	A[2]
Class I	Flash point less than 21 °C	Yes	B	C	C
Class II (1)	Flash point 21-55 °C	No	N/A[1]	C	N/A[1]
Class II (2)	Flash point 21-55 °C	Yes	B	C	C
Class III (1)	Flash point 55- 100 °C	No	N/A[1]	C	N/A[1]
Class III (2)	Flash point 55- 100 °C	Yes	B	C	C
Unclassified (1)	Flash point greater than 100 °C	No	N/A[1]	C	N/A[1]
Unclassified (2)	Flash point greater than 100 °C	Yes	B	C	C

Note 1: Not applicable (N/A) because liquids are not handled above their flash point cannot be above their boiling point.
Note 2: Cryogenic fluids need special consideration.

From the definitions it is clear that for the same flammable fluid, at various stages of its processing or handling, there can be different fluid categories depending upon temperature and pressure at the specific points of release. Each situation at a point of release should be separately evaluated.

7.1.2.12 Area Classification by Direct Examples

The Institute of Petroleum (2002) has produced a set of examples of area classification of simple industry operations in the open. Often the specific case requiring classification is very close to one of the examples provided, and this may justify area classification by just copying this specific example, either without any changes or by appropriate adjustment to suit the case to be addressed. The publication advises on modifications that may be made.

Figure 7–2, Figure 7–3, and Figure 7–4 constitute a small selection of the comprehensive collection of examples provided by IP.

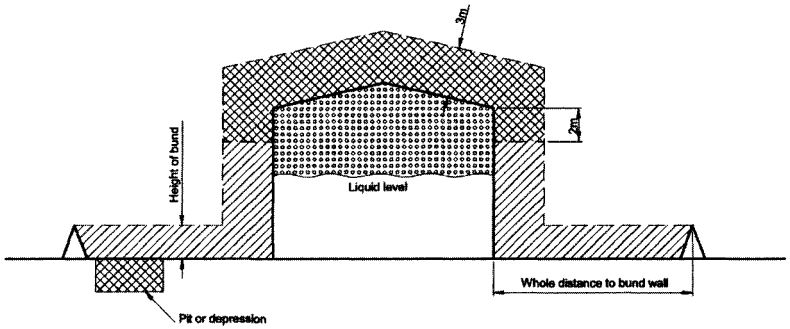


Figure 7–2 Area classification of a bunded tank with cone or dome roof. From The Institute of Petroleum (2002).

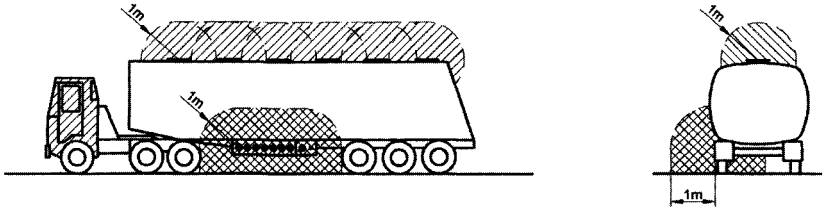


Figure 7–3 Area classification of a fuel tank vehicle during filling. From The Institute of Petroleum (2002).

7.1.2.13 Area Classification by Point Source Method Developed by the Institute of Petroleum (2002)

Generally, process plant will constitute a Zone 2 area inside plant boundaries, within which there may be local Zone 1 and, more rarely, Zone 0 areas.

Some release sources are true point sources, e.g. vents, drains and sample points. Other equipment, such as pump units, constitute assemblies of several individual point sources. Determination of hazard radii using the point source methodology entails consideration of each identifiable potential release point in such assemblies.

The point source method comprises the following steps:

(1) identify point sources

(2) determine grades of release

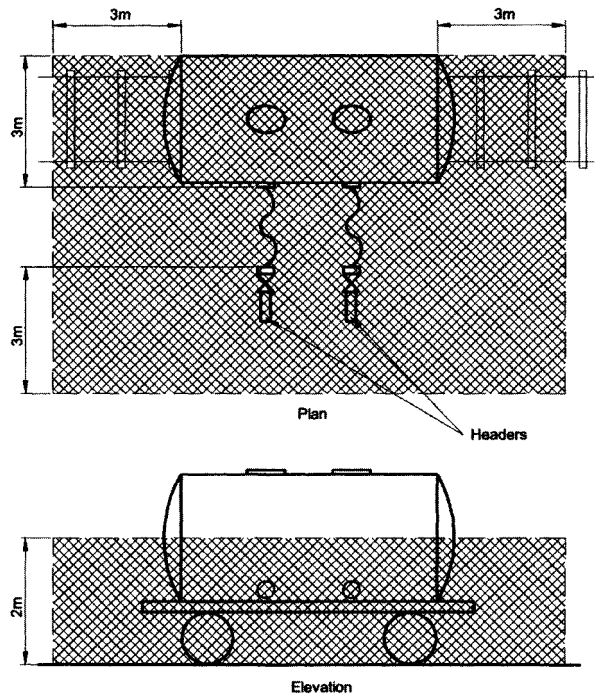


Figure 7–4 Area classification of a rail car discharge of Class I, II and III fuels via hose connection. From The Institute of Petroleum (2002).

(3) determine fluid categories

(4) determine zone categories surrounding each points source

(5) determine hazard radii around each point source

(6) determine global hazardous areas

7.1.2.13.1 Identify Point Sources

The first stage in the area classification of a plant or facility is the identification of the possible point sources of release. Usually these are small, and the associated equipment typically includes valves, flanges, vents, sampling and drainage points, instrument connections, rotating machinery such as pumps and compressors, and any areas where spillage from these sources could collect. The Institute of Petroleum

(2002) provides the following list of equipment configurations that often give rise to leak point sources:

- pumps
- equipment drains and liquid sampling points
- compressors
- vents
- piping systems
- pig receivers and launchers
- liquid pools during spillage
- sumps, interceptors and separators
- pits and depressions
- surface water draining systems

7.1.2.13.2 Determination of Grades of Release

In principle, the area classification procedure entails the consideration of all actual and potential sources of release of flammable gases or vapors. All continuous and primary grade sources of release should be identified and assessed to determine the extent of the resulting Zone 0 and Zone 1 hazardous areas for each source. For secondary grade sources of release, it is often sufficient to consider only the sources located towards the periphery of a plant, which will be the point sources defining the outer boundary of the overall Zone 2 area. The extent of vapor travel, and hence the hazard radius for each point of release to be assessed, will be a function of the fluid characteristics and vapor-forming conditions during release, including mass or mass rate and the rate of vaporization. In assessing the grade of a release, the following points are pertinent:

(a) In the case of operationally controlled releases, e.g. sampling and draining points, vents, filter-cleaning and pig-receiving operations, no single grade of release is generally applicable. This is because the operational frequency chosen will determine whether the equipment release should be graded as primary, secondary or continuous.

(b) For uncontrolled releases, e.g. from pump seals, glands and pipe flanges, the Institute of Petroleum (2002) has provided release hole

sizes assessed and compiled for various situations on the basis of experience. Alternatively the manufacturers of various equipment may have provided typical specific release rates, or actual failure rates have been determined through observation for specific release situations.

7.1.2.13.3 Determination of Fluid Categories

The fluid category for each point as defined in Table 7–2 has to be determined, e.g. by using tables provided by the Institute of Petroleum (2002). It is important to consider that the fluid state under shutdown conditions may give rise to a different vapor producing potential than the state under normal operating conditions. An example would be the fluids in a fractionating column or receiver, which at shutdown might be less well separated from lighter fractions.

7.1.2.13.4 Determination of Zone Category for each Point Source

The zone category around a point of release is a function of the grade of release and the type of ventilation in the area of the release.

7.1.2.13.5 Determination of Hazard Radii for the Zones around Each Point Source

The methodology for determining the hazard radius is dependent on the grade of release (primary, continuous, or secondary). The Institute of Petroleum (2002) has produced a system of block diagrams and tables by which hazard radii may be estimated. However, if the case considered is not covered by the standard fluid categories and tabulated values, dispersion modeling using an adequate numerical code will have to be carried out.

7.1.2.13.6 Determination of Global Hazardous Area Drawing

The global area classification drawing is obtained by constructing the envelopes of the various zone categories on the drawing containing all the individual hazardous zones for the point sources.

7.1.3 Clouds of Liquid Droplets (at Temperatures Below Flash Point of Liquid)

As pointed out by The Institute of Petroleum (2002), flammable atmospheres may also be formed where flammable fluids handled below their flash points are released in the form of a spray or mist. Such materials are normally regarded as non-hazardous from a flammability-of-liquids point of view. However, they should be treated as hazardous when being pumped or stored under pressure, because they may then be capable of producing a spray or mist due to the possibility of a release from a small hole or flange leak. They should be regarded as a Category C fluid generating a hazardous area as appropriate. There is little knowledge on the formation of flammable mists and the appropriate extents of associated hazardous areas. Some of the issues related to the formation and the hazard of flammable mists are discussed by Bowen and Shirvill (1994). Pressure differentials of a few bars are sufficient to atomise commonly encountered liquids. Generation of mists created by the impact of liquid streams on a surface close to the point of release also appears to be possible. However, Bowen and Shirvill do not make any recommendations for the assessment of the extent of flammable clouds of mists and sprays. They do suggest, however, that porous spray guards can be used around flanges and known potential leakage points, causing the material to coalesce back to a liquid below its flash point, rendering it non-hazardous. But this is only practicable in a few specialized applications, and requires rigorous control over maintenance activities to ensure that removed guards are reinstalled after work completion.

Until further research has clarified the situation more fully, The Institute for Petroleum (2002) suggests as follows: Where the fluid temperature is at least 5°C below the fluid's flash point, and the pressure is atmospheric or only a few m liquid head in non-pressurized storage tanks, the fluid can be treated as non-hazardous. However, if being pumped and/or stored under high overpressure, it should be regarded as a Category C fluid (see Table 7–2), generating a hazardous area because of the possibility of mist or spray being generated by leaks from a small hole or a flange.

7.1.4 Dust Clouds

7.1.4.1 Basic Similarities and Differences between Dusts and Gases/Vapors

As discussed in Chapter 2 and Chapter 5 explosive gas/vapor clouds and explosive dust clouds, once existing, exhibit very similar ignition and combustion properties, such as

- flammability/explosivity limits
- laminar burning velocities and quenching distances
- the response of the burning velocity to cloud turbulence
- detonation phenomena
- adiabatic constant-volume explosion pressures of similar magnitudes
- well-defined minimum ignition energies, and
- minimum ignition temperatures for given experimental conditions

Recognition of these similarities may have contributed to the development of a new gas/dust "alignment" concept that has been adopted in the European "Atex" philosophy. However, there are two fundamental ways in which dusts differ substantially from gases, and both should have a major impact on the ways that electrical apparatuses are constructed to prevent them from becoming potential initiators of dust explosions and fires.

The first difference is in the ranges of hazardous fuel concentrations in air. For mixtures of combustible gases/vapors and air, flame propagation is only possible when the fuel/air mixing ratios lie between the lower and the upper flammability limits. Dust flame propagation, however, is not limited only to the flammable dust concentration range of clouds. The state of settled layers/deposits constitutes an additional singular regime of flame propagation. This is because, contrary to combustible gases and liquids, settled powders/dusts in air will always have some air trapped in the voids between the particles, which facilitates sustained, although often very slow, combustion propagation throughout the deposit.

The second basic difference between dusts and gases is in the generation and sustainment of explosive clouds. The paramount question is whether there will be an explosive dust cloud in the first place. The physics of generation and sustainment of dust clouds and premixed gas/vapor clouds are so substantially different that explosive dust clouds are highly unlikely to

be generated in a variety situations where explosive gas/vapor clouds may form quite readily, e.g. by release from process equipment to the open atmosphere. The important consequence is that in practice the explosive dust clouds that give rise to primary dust explosions are always found inside process equipment, such as mills, mixers, ducts, buckets elevators, cyclones, filters, and silos. This, of course will also be reflected in area classification.

NFPA (1997) contains a paragraph that in an excellent way, using practical terms, clarifies the basic difference between gases and dusts with regard to their abilities to migrate through narrow passages. In a slightly modified form this paragraph says as follows:

> Walls are much more important in separating hazardous and non-hazardous zones in the case of combustible dusts than in the case of combustible gases. Only completely non-perforated solid walls make satisfactory barriers in the case of gases, whereas closed doors, light-weight partitions, and even partial partitions could make satisfactory barriers between hazardous and non-hazardous zones in the case of dusts.

7.1.4.2 Definition of Zones

Many countries including U.S.A., Germany, and Norway traditionally used a dust zone classification system based on two hazardous zones only. However, in many countries this system has gradually been replaced by a common three-zone concept. The Atex 118a Directive (1999) defines the three hazardous zone categories that have to be identified in areas containing combustible dusts, in the context of European standardization. These definitions, which have also been adopted by IEC (International Electrotechnical Commission), are:

- Zone 20

  A place in which an explosive atmosphere in the form of a cloud of combustible dust in air is present continuously, or for long periods, or frequently.

- Zone 21

  A place in which an explosive atmosphere in the form of a cloud of combustible dust in air is likely to occur in normal operation occasionally.

- Zone 22

  A place in which an explosive atmosphere in the form of a cloud of combustible dust in air is not likely to occur in normal operation but, if it does occur, will persist for a short period only.

In the context of the philosophy of the Atex 118a Directive (1999), layers/deposits of combustible dusts are not considered as hazardous in themselves. They are only taken into account if they are considered to be possible sources of generation of explosive dust clouds, or if they can give rise to dust fires that can ignite dust clouds.

The symbols used for identifying the various zone categories are shown in Figure 7–5.

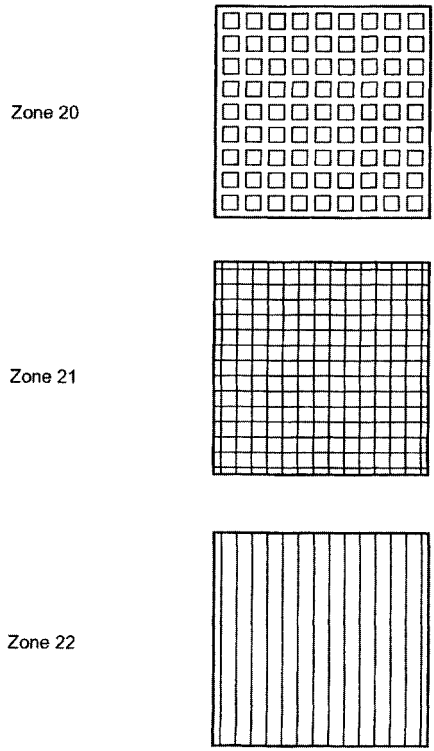
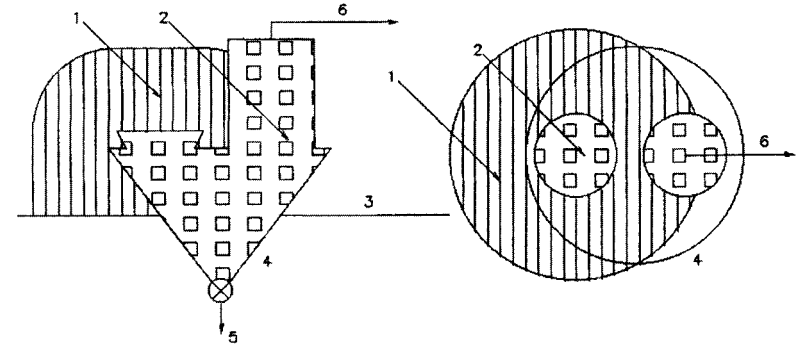


Figure 7–5 Standardized system for marking zones 20, 21 and 22 on hazardous area classification maps of industrial plant. From CENELEC (2002).

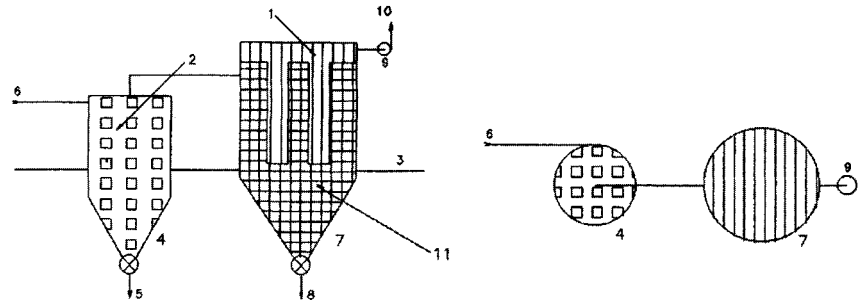
7.1.4.3 Examples of Area Classification

Figure 7–6 and Figure 7–7 illustrate the application of area classification in two practical cases.



Legend:
1 Zone 22 usually I m radius
2 Zone 20
3 Floor
4 Bag discharge hopper
5 To process
6 To extract

Figure 7–6 Marking of zones 20 and 22 on a hazardous area classification map of a station for emptying bags containing combustible powder into a hopper equipped with dust extraction. From CENELEC (2002).



Legend:
1 Zone 22
2 Zone 20
3 Floor
4 Cyclone
5 To product silo
6 Inlet
7 Filter
8 To fines bin
9 Extract fan
10 To outlet
11 Zone 21

Figure 7–7 Marking of zones 20, 21, and 22 on hazardous area classification map of dust collection plant comprising a cyclone and a filter in series. From CENELEC (2002).

The interior of the hopper in Figure 7–6 is classified as Zone 20 because explosive dust cloud will be generated there during the quite frequent emptying of bags into the hopper. Because of the dust extraction system, explosive dust clouds are not expected to extend outside the hopper in normal operation. However, abnormal situations may arise where this may occur, e.g. if a bag bursts during being emptied, or if the dust extraction system fails. Therefore a Zone 22 is assigned to a limited area just outside the hopper opening.

The interior of the cyclone in Figure 7–7 is classified as Zone 20 because of the presence of an explosive dust cloud frequently, or even continuously, for long periods. However, the dust concentration in the gas leaving the cyclone and entering the dusty side of the filter is normally below the minimum explosive concentration, except for short periods when the filter bags are blown or shaken to release accumulated dust. Hence, the dusty side of the filter is classified as Zone 21. Normally only negligible quantities of dust will exist on the clean side of the filter. However, filter bags may burst, or other abnormal situations may arise causing explosive dust clouds to be generated on the clean side of the filter, which is therefore classified as Zone 22.

7.1.5 Explosives, Pyrotechnics, and Propellants

7.1.5.1 Introduction

CENELEC (1997) defines explosive substances as solid, liquid, pasty or gelatinous substances and preparations which are liable to react exothermally and under rapid gas generation even without the participation of atmospheric oxygen, and which, under specified test conditions under partial confinement will detonate or deflagrate rapidly during heating. It is important to note that this definition not only comprise the genuine explosives, but also pyrotechnics and propellants.

7.1.5.2 Definitions of Hazardous Zones According to Cenelec

According to CENELEC (1997), hazardous zones in the context of explosives, pyrotechnics, and propellants are areas where, during manufacture, production, processing or storage of such substances, there may be a finite risk of ignition due to electrical equipment. CENELEC (1997) introduces the following zone definitions:

- Zone E1

  Areas where explosive substances

  - come in contact with electrical equipment as a function of design and/or process, or
  - may to a significant extent appear as dust, vapor, liquid, sublimate, or another state.

- Zone E2

  Areas where explosive substances

  - do not come in contact with electrical equipment as a function of design and/or process, but
  - may occasionally occur in the form of dust, vapor, liquid, sublimate, or another state.

- Zone E3

  Areas where explosive substances

  - do neither come into contact with electrical equipment as a function of design and/or process, nor
  - occur in the form of dust, vapor, liquid, sublimate or another state, neither as a function of design nor of process.

7.1.5.3 Definitions of Hazardous Zones According to a Swedish Standard

The Swedish standard SS 421 08 24(1988) defines the various zones as follow:

- Zone E1

  Areas where

  - explosives are handled, and where
  - considerable quantities of dust, vapor, condensate and/or sublimate from explosive substances can normally occur.

All the interior of containers, vessels, piping etc. for explosive substances is normally classified as Zone E1, irrespective of whether there is a risk of dust generation.

- Zone E2

  Areas where

  - explosives are handled (not stored), and where
  - considerable quantities of dust, vapor, condensate and/or sublimate from explosive substances occur only rarely, or where
  - only non-dusting handling of explosive substances is taking place occasionally, but where specific requirements as to electrical apparatuses are nevertheless considered necessary.

- Zone E3

  Areas where

  - explosive substances are only stored in packed form so that spreading of the explosive substances in the area is effectively prevented.

7.1.5.4 Temperature Classes

SS 421 08 24(1988) specifies the following temperature classes for apparatuses and equipment in contact with explosive substances:

- ET 1: apparatuses and equipment suitable for substances having a minimum ignition temperature of at least 180°C
- ET 2: apparatuses and equipment substances having a minimum ignition temperature of < 180°C

SS 421 08 24(1988) gives minimum ignition temperatures for a range of explosive substances. For TNT the value is 290-310°C, i.e. the temperature class is ET 1.

7.1.5.5 An Example of Area Classification According to the Swedish Standard

The example is the casting plant for TNT illustrated in Figure 7–8.

The process taking place in this example is as follows: At first the drums containing the TNT chips to be cast are pre-heated in the pre-heating room. No significant dust generation is expected, and the entire room is classified as E2.

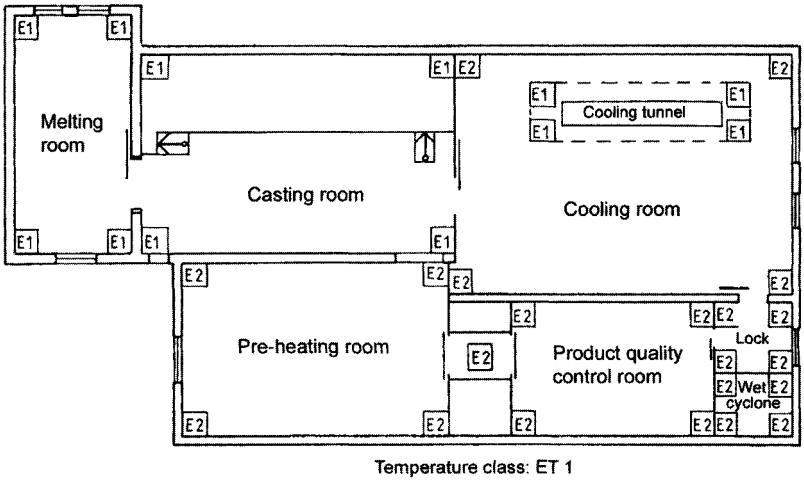


Figure 7–8 Area classification of plant for casting of TNT. From SS 421 08 24 (1988).

The drums are then transferred to the melting room where the TNT chips are poured from the drums into a melting pot. This process creates considerable amounts of TNT dust. In spite of local dust extraction just above the pouring point, the dust cloud may at times extend across the entire room. The molten TNT is poured from the melting pot into intermediate containers. These are transported from the melting room to the casting room by an elevated transportation track running outside the building, from which the containers are hanging freely while being transported.

The entire melting room is classified as E1, whereas the transportation area outside the melting room to the casting room, including a 0.5 m wide zone on each side of, as well as above and below, the suspended containers, is classified as E2.

In the casting room the molten TNT is poured from the transportation container into the molding pot. Further TNT chips are added to the molten material while this is being stirred and shaken. Even during this process there will be some dust generation, most of which is removed by local dust extraction.

When the TNT has reached the desired slurry-like consistency, it is poured into high-explosives shells, mines etc. according to the production

plan. During this process there will be some generation of TNT vapor, which sublimates as soon as it makes contact with cold surfaces. Furthermore, some spill of molten TNT during casting is unavoidable. The floor in the casting room is kept wet by applying water, both in order to collect settling TNT dust and to prevent spilled TNT from the casting operations from becoming firmly stuck to the floor.

The conditions in the casting room are generally such that the entire room is to be classified as E1. The outdoor transportation track from the casting room to the cooling room is classified as E2 in the same way as the track from the melting room to the casting room.

In the cooling room the cast products are cooled down, normally in special cooling cupboards or cooling tunnels, which are usually equipped with local dust extraction. The interior of these tunnels and cupboards is to be classified as E1, whereas the cooling room in general is E2. However, often there is a need to remove excess TNT that has become attached to the outside of the mines, high-explosives shells etc. during the casting process. In such operations TNT dust is likely to be generated, and a local spherical E1 zone with a radius of 2 m around the point of such material removal has to be established.

The room for the wet cyclone that collects the dust from the various local dust extraction stations is classified as E2. Finally the quality of the cast products has to be controlled in a special product quality control room, which is normally classified as E2.

The temperature class for electrical and other equipment requiring consideration is ET 1, which means that the maximum permissible temperature of any hot surface in all classified areas is 180°C.

7.1.5.6 Australian Interim Standard

An Australian interim standard (2004) defines an explosives hazardous area an area in which an explosives vapor atmosphere and/or an explosives dust atmosphere is present, or may be expected to be present in quantities such as to require special precautions for the construction, installation, and use of electrical equipment. For explosives vapor areas three zone categories 0E, 1E and 2E are defined in accordance with the definitions of zones 0, 1, and 2 in section 7.1.2.7. For explosives dust

cloud areas three zone categories 20E, 21E, and 22E are defined in accordance with the definitions in 7.1.4.2.

In addition, a zone category SE is defined as follows: an area in which very small quantities of explosives may be present, where its ignition could not cause the subsequent initiation of other hazardous materials, significant damage to equipment, or injury to personnel.

7.2 Basic Design Concepts for Electrical Apparatus

7.2.1 Gases and Vapors

A series of standardized basic design concepts for electrical apparatuses intended for use in explosive gas atmospheres have been available for a long time. The details are described in a corresponding series of international standards (IEC, CENELC etc.). The following summary is mainly based on BBC (1983). The figures are from Eckhoff (1996).

7.2.1.1 Intrinsic Safety (Ex 'i')

This design concept can be used for apparatuses to be used in all three Zones (0, 1 and 2). An intrinsically safe circuit is a circuit in which no spark or any other thermal effect can be generated, which is capable of causing ignition of a given explosive atmosphere. The basic principles of intrinsically safe design are given in Figure 7–9.

In order to prevent hot-surface ignition of an explosive gas/vapor atmosphere by electrical apparatuses, the apparatuses must be designed so as to ensure that the temperatures of all surfaces that can make contact with the explosive atmosphere are below the minimum ignition temperature of the explosive atmosphere (see Section 2.2.4.4). For practical reasons it has been agreed internationally to standardize on a limited number of temperature classes for electrical appartatuses, and these are given in Table 7–4. Table 2–2 in Chapter 2 gives the minimum ignition temperatures and the corresponding temperature classes for a range of combustible gases and vapors in air.

The intrinsic safety concept originated in UK nearly 100 years ago through the pioneering work by Wheeler and others. British Standard

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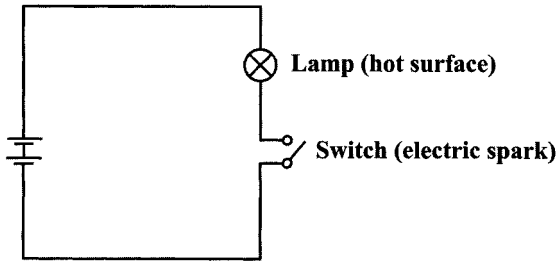

Figure 7–9 Illustration of two types of ignition sources that are to be prevented by intrinsically safe (Ex 'i') systems: Incendiary electric spark/arcs generated by closing or breaking electrical circuits (switch in figure), or incendiary hot surfaces (lamp in figure).

Table 7–4 Temperature Class Symbols Indicating Internationally Standardized Maximum Permissible Sufrace Temperatures of Electrical Apparatus

T1	450°C
T2	300°C
T3	200°C
T4	135°C
T5	100°C
T6	85°C

1259 from 1945 contains detailed specifications for intrinsically safe electrical circuits for mining underground. British Standard 1538 from 1949, dealing with *Intrinsically Safe Transformers Primarily for Bell Signalling Circuits* covered electrical devices for mining underground. These systems had to satisfy the following conditions:

- the secondary voltage of the supply transformer of these circuits was limited to 15 V
- the current was limited (by a resistor) to a maximum of 1.5 A
- the signal was to be given by a bell
- the local cables in the mine were two bare wires laid in parallel

An intrinsically safe apparatus is an electrical apparatus in which all the circuits are intrinsically safe. Intrinsically safe electrical apparatuses are produced to different standards depending on the ignition sensitivity of the explosive atmosphere in which the apparatuses are to be used (gas groups I for coal mines and groups IIA, IIB and IIC for other activities. See Table 2–2 and Section 2.2.7.2 in Chapter 2). The reference gases are methane for group I, and propane, ethylene and hydrogen for groups IIA, IIB and IIC respectively.

In addition intrinsically safe electrical apparatuses are divided into the categories Ex 'ia' and Ex 'ib'. Ex 'ia' comprises electrical apparatus that shall not be capable of causing ignition neither in normal operation, nor with one fault and with any combination of two faults. Category Ex 'ib' comprises electrical apparatus incapable of causing ignition in normal operation and with one single fault. The traditional approach was that only Ex 'ia' apparatus could be used in Zone 0, but more recently allowance has also been granted for some other concepts (see Section 7.2.1.8).

In intrinsically safe circuits a distinction is made between intrinsically safe electrical apparatus and *associated* electrical apparatus. An electrical apparatus is intrinsically safe when all its circuits are intrinsically safe. In an associated electrical apparatus not all circuits are intrinsically safe but contains circuits that can affect the intrinsic safety. Typical examples of intrinsically safe electrical apparatus are passive two- or four-terminal networks, limit switches, and measuring instruments containing a coil. Associated electrical apparatus is, for example, power supply units for intrinsically safe circuits with intrinsically safe output. Associated electrical apparatus must either be mounted outside the hazardous area or be fitted with an additional type of protection.

Safety barriers are isolating elements generally accommodated outside hazardous areas to isolate intrinsically safe circuits from non-intrinsically safe circuits so reliably that any chance of mutual influence is excluded. Safety barriers with galvanic isolation are used, i.e. transformers, d.c. transformers and relays, and also those without galvanic isolation, i.e. safety barriers containing diodes or Zener diodes.

In principle the intrinsic safety of e.g. simple inductive or capacitive circuits can be assessed theoretically by means of the appropriate curves in Figure 2–30 and Figure 2–31 in Section 2.2.6, based on the minimum ignition energy (MIE) of the explosive gas atmosphere in which the apparatus is to be located. However, often direct testing of the electrical

apparatus of concern, using the standard spark test apparatus shown in Figure 2–29 in Section 2.2.6 is required. This apparatus is designed to produce both closing and breaking contacts in the actual explosive gas mixture. The standard test gas mixtures are:

- Gas group I: (8,3 ± 0.3) % methane in air
- Gas group IIA: (5,25 ± 0.25) % propane in air
- Gas group IIB: (7.8 ± 0.5) % ethylene in air
- Gas group IIC: (21 ± 2) % hydrogen in air

7.2.1.2 Flame Proof Enclosures (Ex 'd')

This design concept can be used for apparatuses to be used in Zones 1 and 2. Flame proof design implies that electric parts which can ignite an explosive atmosphere are placed in an enclosure which can withstand the pressure developed during an internal explosion of an explosive mixture and which prevents the transmission of the explosion to any explosive atmosphere surrounding the enclosure. The principle of flame-proofing is illustrated in Figure 7–10.

The following aspects are central in the design of flame proof enclosures:

- For enclosures in which the contained electrical/electronic circuits are essentially activated, the volume to be considered is the free volume. For luminaries, the volume is determined without the lamps fitted.
- The flame proof joint is the place where corresponding surfaces of two parts of an enclosure come together and prevent the transmission of an internal explosion to the explosive atmosphere surrounding the enclosure.
- The width of flame proof joint is the shortest path through the joint, from the inside to the outside of an enclosure.
- The gap of flame proof joint is the distance between the corresponding surfaces of a flame proof joint when the electrical apparatus has been assembled. For cylindrical surfaces, the gap is the difference between the diameters of the bore and the cylindrical component inserted into the bore.

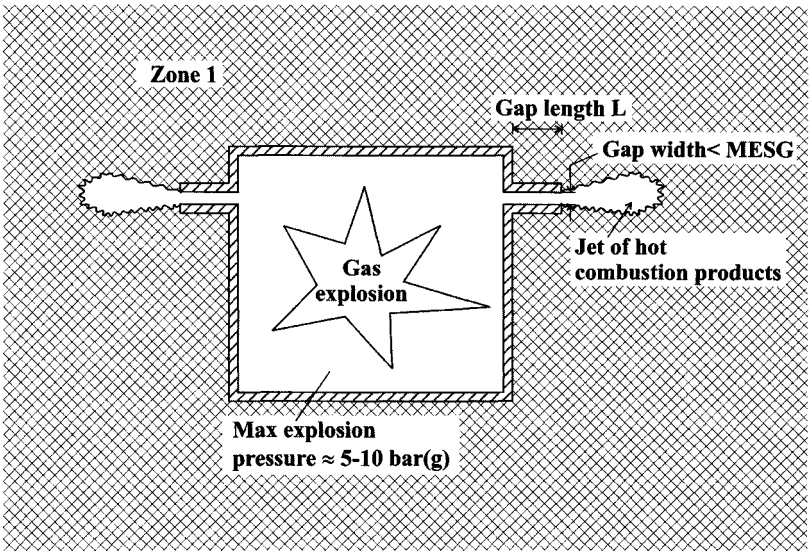

Figure 7–10 Illustration of a flame proof enclosure (Ex 'd'). The enclosure must satisfy three main requirements: Gap widths < MESG at actual conditions; enclosure to withstand the maximum internal overpressure that an internal gas explosion can produce at actual conditions; and temperature of external enclosure surface < min. ign. temp. at actual conditions.

- The maximum experimental safe gap for an explosive mixture is the maximum gap of a joint of 25 mm width which prevents any transmission of an explosion in ten tests made under standard test conditions, using the apparatus illustrated in Figure 2–44 in Section 2.2.7.
- Pressure piling is the increase in the explosion pressure above the normally expected pressure, e.g. as a result of subdivision of the flameproof enclosure. (See Figure 2–60 in Section 2.4.5.2).
- A quick acting door or cover is a door or cover provided with a device which permits opening or closing by a simple operation, such as the movement of a lever or the rotation of a wheel.
- A door or cover fixed by screws is a door or cover the opening or closing of which requires the manipulation of several screws or nuts.

The design of flameproof joints is specified in detail in international standards. The flameproof joints can be flanged, cylindrical, spigot and threaded joints. The design of non threaded gaps shall withstand the mechanical stresses arising. The surfaces of joints shall be machined so that their average roughness does not exceed certain limits. The joint surfaces should be protected against corrosion, by grease, electroplating, chemical treatment etc. Painting is not acceptable. The width of a joint shall be at least the minimum values specified in the standards, and interconnecting compartments shall be precluded as far as possible.

As for intrinsically safe apparatus, the design of flameproof enclosures also depends on the ignition properties of the explosive gas atmosphere in which the enclosure is to be used. If the standard MESG (see Table 2–2 and Section 2.2.7.2 in Chapter 2) of this gas atmosphere is known, the gas group to which the actual gas belongs is also known. International standards then specify the design requirements to be satisfied. Nevertheless, flameproof enclosures shall also be tested for their ability both to withstand the pressure of an explosion inside the enclosure, and to prevent transmission of an internal explosion to an explosive gas cloud on the outside.

The tests of the ability of the enclosure to withstand the pressure of an internal explosion are made in two stages. First the maximum explosion pressure, or reference pressure, is measured by igniting an explosive gas mixture inside the enclosure to be tested. The number of tests to be made and the explosive mixture to be used in the tests depend on which gas group the enclosure is to be exposed to. When the reference pressure has been determined a proper pressure test is carried out, either as a static or a dynamic test. In a static test, a test pressure exceeding the reference pressure by a factor of typically 1.5 shall be maintained inside the enclosure for a specified time. In a dynamic test, the enclosure is also subjected to 1.5 times the reference pressure, and the rate of rise of pressure shall be similar to that obtained during the determination of the reference pressure by the explosion tests. The tests are considered satisfactory if the enclosure has not suffered any damage or permanent deformation liable to weaken any of its parts.

For the test for non transmission of an internal explosion the enclosure is placed in a test chamber and the same explosive mixture is introduced both inside the enclosure and in the test chamber in which the enclosure is placed. The initial pressure before igniting the inside mixture is normally atmospheric both inside and outside the enclosure.

7.2.1.3 Increased Safety (Ex 'e')

This design concept can be used for apparatuses to be used in Zones 1 and 2. Increased safety is a type of protection by which measures are taken so as to prevent the possibility of excessive temperatures and of the occurrence of arcs and sparks inside the enclosure and on the external parts of it, in normal service. The principles of the concept of increased safety are illustrated in Figure 7–11.

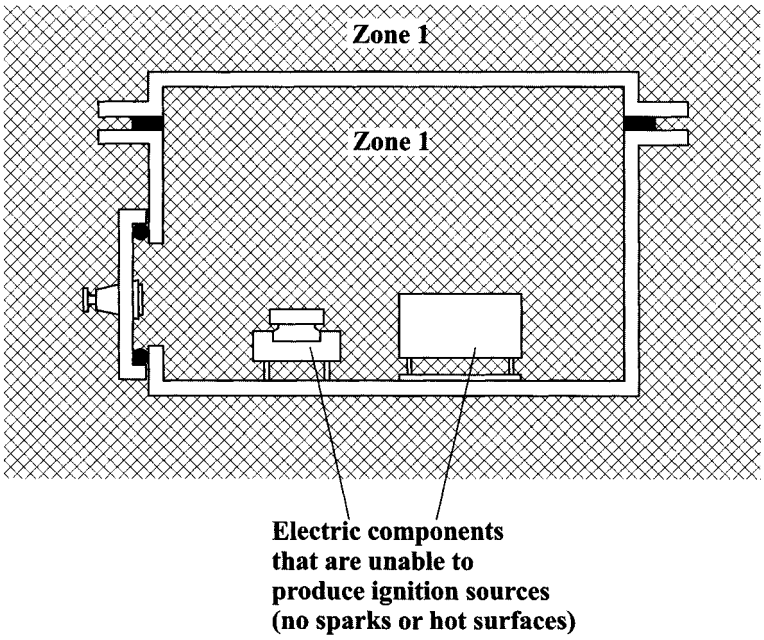

Figure 7–11 Illustration of increased safety enclosure (Ex 'e'). The enclosure must satisfy the following main requirements: No ignition sources permitted inside enclosure; satisfactory sealing (IP-protection) against ingress of water and dust; enclosure body made of earthed anti static material; and temperature of external enclosure surface < min. ign. temp. at actual conditions.

The following aspects are central in the design of increased-safety enclosures:

- The limiting temperature is the highest permissible temperature of an electrical apparatus or a part of an electrical apparatus. This depends

on the minimum ignition temperature of the substances with which the hot surfaces may make contact.

- The minimum clearances between conducting parts at different potential shall satisfy specific requirements stated in international standards.
- The creeping distances between conducting parts at different potential shall meet specific requirements stated in international standards. The creeping distance is the shortest distance between two conducting parts along the surface of the insulating parts.

The insulating parts are often provided with ribs or notches, to increase the creeping distances. Notches in the surface of insulating parts may be taken into consideration for calculating the creeping distance only if the notches are at least 3 mm deep and 3 mm broad; ribs only if they are at least 3 mm high and their width is appropriate to the mechanical strength of the material, the minimum being 1 mm.

For the increased safety type of protection special attention is paid to the design of terminals for external connections and the internal connections. As these parts can be ignition sources, it is required that each connection shall be made such that the contact does not deteriorate due to heating in service, alteration of the insulating material or vibration, i.e. its contact resistance remains constant even during a protracted period in service. Terminals for external connections shall be generously dimensioned to permit the effective connection of conductors of a cross-section at least corresponding to the rated current of the electrical apparatus. The terminals shall be

- fixed in their mountings without possibility of self-loosening
- constructed in such a way that the conductors cannot slip out from their intended location
- designed to assure proper contact without deterioration of the conductors
- designed/machined without sharp edges
- designed so that they cannot be twisted or permanently deformed during normal tightening
- designed to assure that the contact that they are providing are not appreciably impaired by temperature changes in normal service

The size of the terminal compartment should enable perfect connections to be made using common tools. In the interior of electrical apparatus, connections shall not be subject to undue mechanical stress. Solid insulating materials shall have mechanical characteristics which are suitable at temperatures at least 20 K above the temperature attained in continuous rated service.

Normally enclosures containing live bare parts shall as a minimum provide a degree of protection of IP 54, whereas enclosures containing insulated parts only shall as a minimum provide a degree of protection of IP 44.

7.2.1.4 Pressurized Apparatus (Exp)

This design concept can be used for apparatuses to be used in Zones 1 and 2. Pressurized apparatus Exp is a type of protection by which the entry of a surrounding atmosphere into the enclosure of the electrical apparatus is prevented by maintaining, inside the enclosure, a protective gas at a higher pressure than that of the surrounding atmosphere. The principle is illustrated in Figure 7–12.

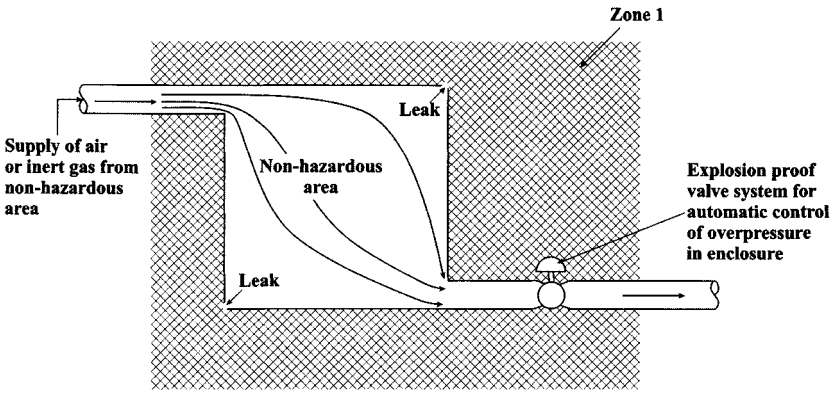

Figure 7–12 Illustration of pressurized enclosure (Ex 'p').

The overpressure is maintained either with or without a continuous flow of the protective gas. It is then assumed that no flammable gas or vapor is introduced into the enclosure and that it contains no internal source of

flammable gas or vapor. The protective gas can be can be air, inert gas or another suitable gas.

Purging is the process of passing a quantity of protective gas through the enclosure and ducts, before the application of voltage to the electrical apparatus inside the enclosure, so that any explosive atmosphere possibly present in the enclosure is expelled and any such explosive atmosphere in the pressurized enclosure is reduced to a concentration significantly below the lower explosive limit. The quantity of protective gas required for purging shall be at least 5 times the volume of the free space in the enclosure and its associated ducts. The overpressure is maintained within the pressurized enclosure by continuous circulation of the protective gas through the enclosure after purging. This type of pressurized enclosure, besides providing protection against explosion, also aids dissipation of any heat from electrical apparatus inside the purged enclosure.

Pressurization with leakage compensation provides this type of protection as long as the supply of protective gas is sufficient to compensate for any inevitable leakages from the pressurized enclosure and its ducts.

The measures specified for this type of protection include specific requirements for the construction of the enclosure and its associated components, including the inlet and exhaust ducts and for the auxiliary controlled apparatus necessary to ensure that the overpressure is established and maintained safely. The protection devices and the ducting for the protective gas shall prevent sparks or incandescent particles from being ejected from the enclosure. The enclosure, ducts and their connecting parts shall be able to withstand an overpressure equal to 1.5 times the maximum overpressure specified in normal service. The materials used for the enclosure and the ducts shall not be affected either by the specified protective gas or by the flammable gases or vapors in which they are to be used. Doors and covers that can be opened without using a tool shall be interlocked so that the electrical supply is disconnected automatically when they are opened and cannot be restored until they are closed.

An automatic device shall be provided by the operator at his own responsibility to operate automatically when the overpressure fails below the minimum prescribed value. A minimum overpressure of 0.5 mbar (50 Pa) shall be maintained relative to the external pressure at any point within the enclosure.

7.2.1.5 Oil-Filled Enclosures (Ex 'o')

This design concept can be used for apparatuses to be used in Zones 1 and 2. Oil immersion is a type of protection in which the electrical apparatus or parts of it are immersed in a specified quality mineral oil in such a way that an explosive atmosphere which may be above the oil or outside the enclosure cannot be ignited by the electrical components. The principle is illustrated in Figure 7–13.

Current standards assume that the electrical apparatus immersed in the oil is fixed in its operating position in accordance with specific installation instructions. All parts capable of producing arcs or sparks in normal service shall be immersed in the oil at a minimum depth. Devices shall be provided so that the oil level can be easily checked when the electrical apparatus is in service. The highest and lowest oil levels permissible in normal service should be clearly marked. To ensure that electrical apparatuses satisfy the requirements for this type of protection certain tests shall be carried out with the highest and the lowest oil level corresponding to a fault in the level gauge, and with an explosive mixture of air and hydrogen above the oil.

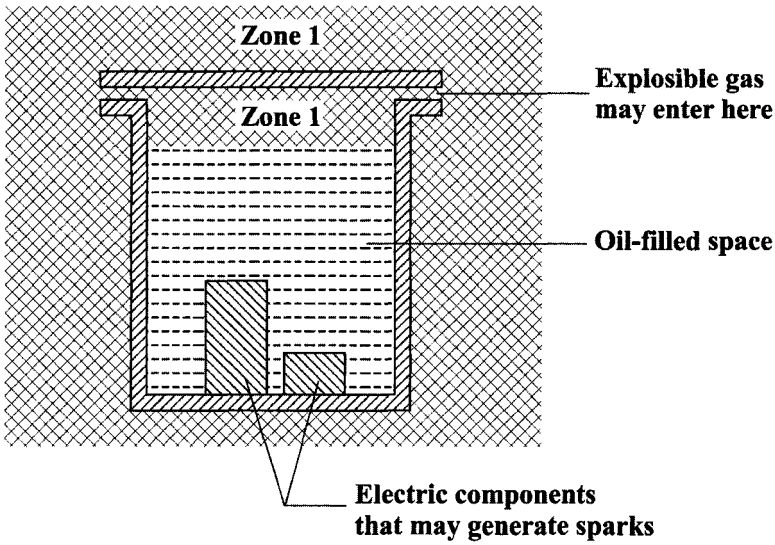

Figure 7–13 Illustration of oil filled enclosure (Ex 'o').

7.2.1.6 Sand-Filled Enclosures (Ex 'q')

This design concept, illustrated in Figure 7–14, can be used for apparatuses to be used in Zones 1 and 2. The following is based on BBC (1983). Sand filling is a type of protection in which the enclosure of electrical apparatus is filled with a non-combustible solid material in a finely granulated state so that, in the intended conditions of service, any electric spark occurring within the enclosure of an electrical apparatus will not ignite a surrounding explosive atmosphere.

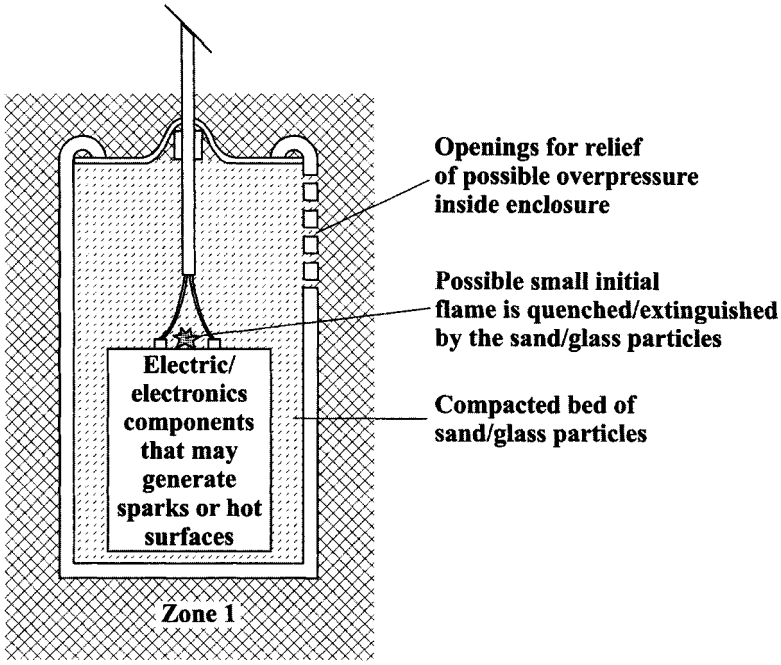

Figure 7–14 Illustration of sand filled enclosure (Ex 'q').

Short circuit time or arcing time is the time during which the arcing current flows through the electrical apparatus, counted from its initiation to its final extinction.

The screen is a perforated metal sheet fixed inside the enclosure, within the mass of the filling material, in such a way as to cover all the live parts of the electrical apparatus inside the enclosure. The minimum safe height is the shortest vertical distance between the free surface of the filling material, after it is suitably shaken down, and the nearest live part, which

prevents transmission of ignition by an electric arc of such current and duration as has been specified for the construction of the electrical apparatus. When the electrical apparatus is provided with a screen, the minimum safe height is the sum of the protection height and the height of the reserve layer. The protection height is the distance between the screen and the nearest live part of the electrical apparatus inside the enclosure. The height of the reserve layer is the thickness of the filling material above the screen, designed to fill up any accidentally formed voids in the safety layer.

The enclosure of powder filled electrical apparatus shall generally be of metal. Some other materials may be permitted, provided their mechanical and thermal properties are satisfactory. The enclosure in its normal service condition, with all openings closed as in normal use, shall comply with the degree of protection IP 54 (see Section 7.2.3.2.1), and adequate mechanical strength shall be confirmed in a hydraulic test.

The quartz sand used as filling material or a filling material of similar quality shall not contain more than 0.1% by weight of water at the time of filling and shall be suitably shaken down in order to prevent the formation of voids. No part made of any organic material shall be used above the live parts, or between them and the walls of the enclosure.

Bare live parts shall be suitably spaced from each other and from the walls of the enclosure. For electrical apparatus with nominal voltage less than or equal to 500 V and factory-enclosed so as to prevent dismantling without destruction, a distance of 4 mm between live parts, and a distance of 5 mm between live parts and the enclosure are sufficient.

Irrespective of its volume, the enclosure shall be subjected to a hydraulic type test with a pressure of 0.5 bar (g) without the occurrence of permanent deformation exceeding 0.5 mm in any of its dimensions. The pressure shall be applied for at least 1 minute.

7.2.1.7 Encapsulation (Ex 'm')

This design concept can be used for apparatuses to be used in Zones 1 and 2. The basic principle is illustrated in Figure 7–15.

Encapsulation (Ex 'm') is a type of protection in which the parts that could ignite an explosive gas atmosphere are enclosed by a polymer compound seal in such a way that the explosive atmosphere cannot come in contact with any sparking or heating which may occur within the encapsulation. The polymer compound can be thermo-setting, thermo-plastic

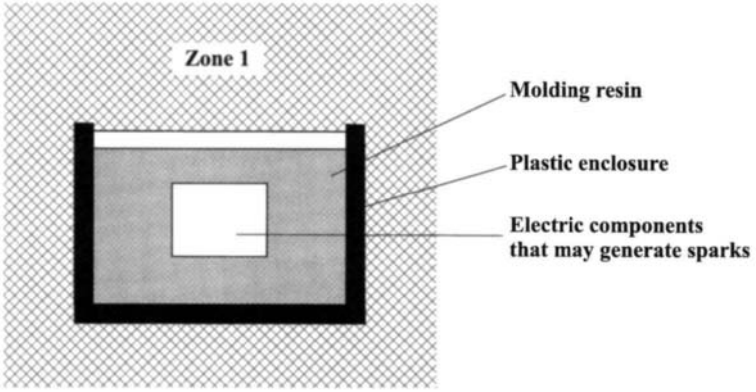

Figure 7–15 Illustration of encapsulated/molded enclosure (Ex 'm').

and elastomer materials with and without fillers and/or other additives. The temperature range of the polymer is the temperature range within which its characteristics satisfy the standard requirements both in service and during storage. The continuous service temperature of the polymer is the maximum temperature to which it can be continuously exposed. The choice of polymer for a particular application depends on the duty that the polymer has to perform in the apparatus to be encapsulated.

The suitability of the polymer and the encapsulation in which it is used are verified by tests of e.g.

- dielectric strength of the polymer
- water absorption of the polymer
- thermal endurance of the polymer and other parts of the component
- maximum surface temperature of the component

7.2.1.8 Design Concepts for Zone 0 Other than Ex 'ia'

During the last decade it has been agreed internationally (International Electrotechnical Commission) that there is a need for additional design concepts than just Ex 'ia' (see Section 7.2.1.1) for electrical apparatuses to be used in Zone 0. The new Zone 0 concepts are typically combinations of at least two independent applications of design concepts for Zone 1 apparatuses.

7.2.1.9 Special Design for Use in Zone 2 Only (Ex 'n')

In the course of the last twenty years special concepts for design of special apparatuses for use in Zone 2 only have been developed. These concepts are essentially just more liberal versions of the design concepts adopted for Zone 1 (and 0) apparatuses. Up to now all the Zone 2 design requirements have been collected in a single independent Ex 'n' standard. In the future specific minimum Zone 2 requirements may be integrated in the various separate standards describing the design concepts outlined in Section 7.2.1.1 to Section 7.2.1.7 above.

7.2.2 Clouds of Liquid Droplets

No specific standards or guidelines for designing electrical apparatuses enclosures to prevent ingress of explosive clouds of liquid droplets seems to exist. However, in the case of sprays of liquids of low boiling points, where one can assume immediate complete evaporation as soon as the drops have been generated, the protection methods for gases described in Section 7.2.1 apply.

7.2.3 Dust Clouds

7.2.3.1 Introduction

The paragraph from NFPA (1997) quoted in Section 7.1.4.1 also indeed applies to design of electrical apparatuses. In this context, the paragraph may be slightly re-phrased by replacing the initial word "walls" by "simple enclosures." Hence:

> Simple enclosures are much more important in separating hazardous and non-hazardous zones in the case of combustible dusts than in the case of combustible gases. Only completely non-perforated solid walls make satisfactory barriers in the case of gases, whereas closed doors, light-weight partitions, and even partial partitions could make satisfactory barriers between hazardous and non-hazardous zones in the case of dusts.

7.2.3.2 Preventing Ignition of Combustible Dusts by Keeping Potential Ignition Sources inside Enclosures that Prevent Hazardous Ingress of Dust

7.2.3.2.1 The "IP" Code for Prevention of Dust Ingress

The use of suitable enclosures to keep dust away from delicate electrical and mechanical components has long traditions. Irrespective of specific hazardous effects, the presence of dusts is generally incompatible with delicate equipment and components, just from the point of view of cleanliness and tidiness. The various reasons for applying this concept include:

- Combustible dust can form an explosive cloud inside the enclosure and cause a dust explosion there.
- Combustible dust can form dust layers inside the closure and cause dust fire there.
- Electrically conductive dust can cause short-circuiting inside the enclosure.
- Abrasive and/or corrosive dusts can damage delicate mechanical components inside the enclosure.

In the context of preventing ignition of combustible dusts, only the first and second reasons are relevant. Furthermore, when considering that formation of explosive dust clouds inside enclosures of a reasonable standard, by ingress of dust from the outside, is highly unlikely, the possibility of dust fire is in fact the only hazard that has some relevance in the present context.

IEC (2001) has produced a standard, named the "IP" code, which defines various "degrees of protection" against ingress of solid objects, including ingress of dust particles, and ingress of water. The initials "IP" just means "International Protection." According to IEC (2001), the "degrees of protection" offered by a given enclosure are to be specified by two digits, the first referring to ingress of solid objects, the second to ingress of water. For solid objects 6 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 protection from gentle dripping (digit 1) to protection from continuous complete immersion (digit 8). The code also specifies specific test methods by which enclosures can be tested for compliance with the requirements of the various degrees of protection.

It is important to note that the IEC "IP" code, IEC (2001), does not cover protection against ingress of explosive gases. Satisfactory protection against ingress of gas is very difficult to achieve using simple enclosure technology, e.g. flanges with gaskets, because gas molecules will migrate though even very tiny openings, in particular if there is a pressure drop across the opening. For this reason it was necessary to introduce all the additional technologies to either prevent ingress of explosive gases into enclosures (oil-filled, pressurized and molded enclosures) described in Section 7.2.1, or to prevent ignition of explosive gas that has entered the enclosure (enclosures filled with glass beads or sand, elimination of potential ignition sources inside enclosure), or to prevent transmission of an explosion inside the enclosure to a possible external explosive atmosphere (flame proof enclosures), also described in Section 7.2.1.

However, in the case of dusts none of these additional measures are required to prevent formation of explosive dust clouds inside enclosures. Common enclosure technology, e.g. flanges with gaskets, provides the protection required. The IEC IP code specifies two levels of prevention of ingress of dust into enclosures, viz. "dust protected" (digit 5) and "dust-tight" (digit 6), which are defined as follows:

- Dust protected

  A limited quantity of dust is allowed to penetrate into the enclosure under certain conditions (IP 5X).

- Dust tight

  No dust is allowed to penetrate into the enclosure (IP 6X).

IEC (2001) also specifies the tests to which enclosures of categories IP 5X and IP 6X have to be subjected. The enclosure to be tested is placed inside a test chamber where a very dense cloud of fine talcum powder is maintained continuously during the test period either by a powder circulation pump, as illustrated in Figure 7–16, or by some other means. Depending on the practical circumstances in industry in which the enclosure to be tested is to operate, tests can be conducted either with a slight negative pressure inside the enclosure to be tested, as also illustrated in Figure 7–16, or with no pressure difference across the enclosure wall.

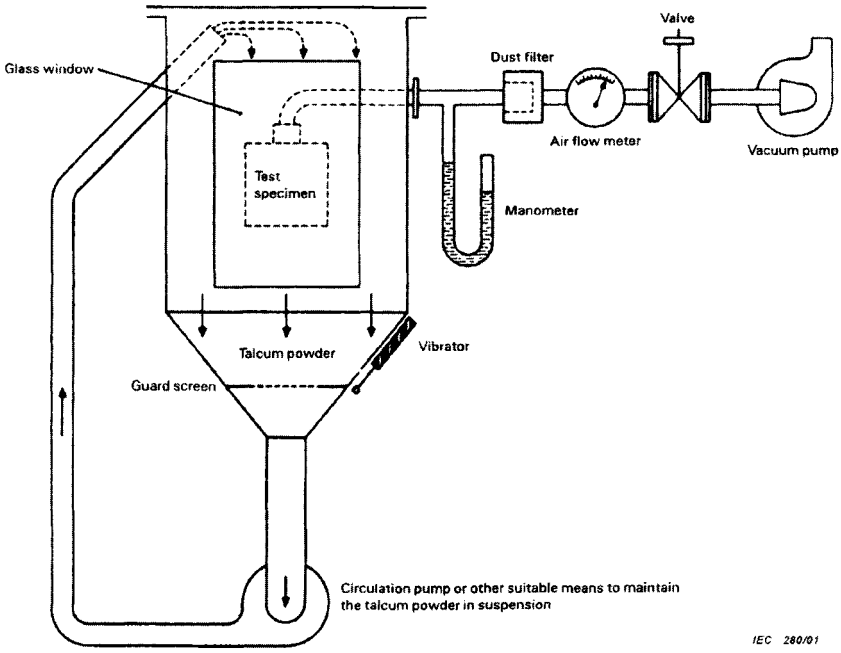

Figure 7–16 Illustration of method used for testing enclosures for their ability to prevent ingress of dust. From IEC (2001).

Figure 7–17 shows an actual test cabinet according to IEC (2001) during dust ingress testing of a three-phased geared electric motor. The circulation of talcum powder had been stopped before the photograph was taken.

CENELEC (1998) specifies the IP requirements for combustible dusts as follows:

- Zone 20 and Zone 21: IP 6X
- Zone 22: IP 6X for electrically conductive dusts IP 5X for electrically non-conductive dusts

CENELEC (1998) also specifies marking codes to be used to identify the degree of protection offered by a given enclosure.

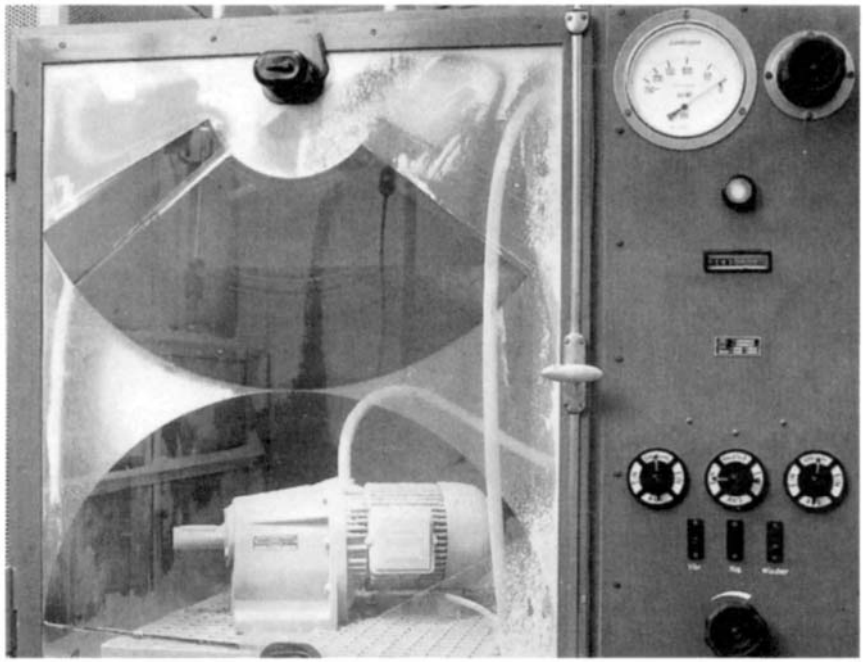

Figure 7–17 Photograph of an actual test cabinet according to IEC (2001) during dust ingress testing of a three-phased geared electric motor. From Greiner (2001).

7.2.3.2.2 Prevent Ion of Ignition of Dust Clouds and Dust Layers by Hot External Enclosure Surfaces

Ignition of dust clouds and dust layers by hot surfaces is discussed in Chapter 5. Although it is known that the minimum ignition temperature of a given dust cloud is not a true inherent constant of the cloud, results from the laboratory scale tests illustrated in Figure 5–25 are regarded as representative of the dust tested. It is customary to require that the maximum temperature of the dust free enclosure surface does not exceed 2/3 of the minimum ignition temperature for dust clouds, in °C, as measured in this standard test.

In the case of ignition of dust layers by hot surfaces, the test method illustrated in Figure 4–4 in Chapter 4 is used. Tests with a given dust, varying the thickness of the dust layer, shows that the minimum ignition temperature decreases systematically with increasing layer thickness. It is customary to require that the maximum temperature of the enclosure surface be at least 75°C lower than the minimum ignition temperature determined

in the test. Figure 7–18 indicates how the maximum permissible enclosure surface temperature decreases systematically with increasing dust layer thickness, for three different dusts having minimum hot surface ignition temperatures of 250°C, 320°C, and 400°C respectively for 5 mm layer thickness. However, if a large part of the hot surface is covered by a comparatively thick dust layer, the surface temperature of the enclosure may increase to a value significantly higher than that attained in the absence of dust. In that case Figure 7–18 does not apply, and special assessment will be required. This may imply both special tests and mathematical model simulations.

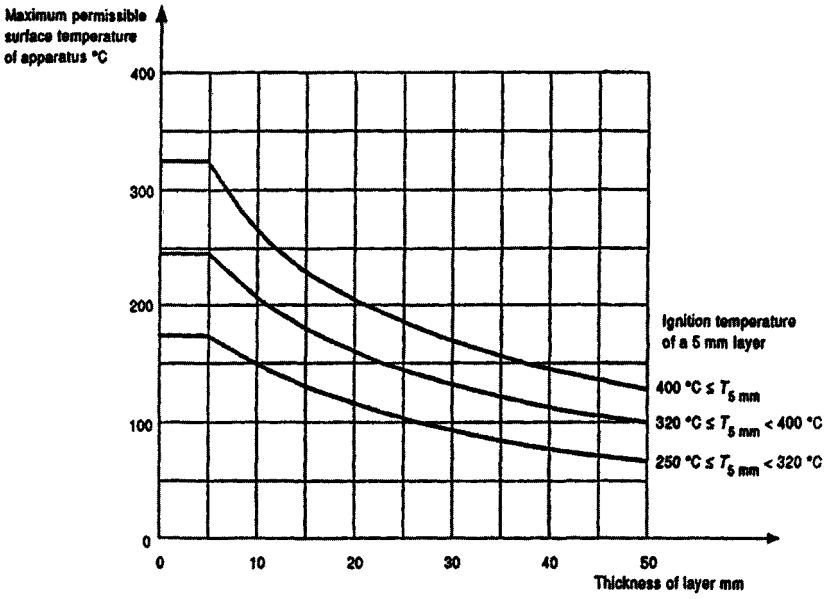

Figure 7–18 Graphs for maximum permissible temperatures of enclosure surfaces as functions of the thickness of a layer of combustible dust on the surface. It is assumed that the dust layer on enclosure surface does not significantly increase the temperature of the surface as compared to that attained without a dust layer. From CENELEC (1998).

7.2.3.2.3 Other Requirements to IP Enclosures

- Enclosures made of plastic materials must be able to withstand certain specified thermal load tests, both in the range of low and high

temperatures. This also applies to any plastic materials used for cementing.

- Enclosures must be able to withstand relevant tests for mechanical strength.
- All metal parts, which by becoming electrostatically charged can give rise to electrostatic discharges that can ignite clouds or layers of the dust in question, must be properly earthed and bonded.
- In order to avoid propagating brush discharges (see Chapter 2 and Chapter 5) enclosures made of plastic materials must satisfy certain requirements to the maximum permissible insulation resistance to earth, the maximum permissible breakdown voltage across the thickness of the plastic wall, or the minimum permissible thickness of external plastic insulation on metal.
- Exposed parts of enclosures must not contain metals that are able to generate impact sparks that can ignite clouds or layers of the actual dust.

7.2.3.3 Other Proposed Methods for Design of Enclosures for Electrical Apparatus to be Used in Areas Containing Combustible Dusts

7.2.3.3.1 Background

For a long time design of equipment for areas containing combustible dusts in order to prevent ignition of the dust has essentially been based on the two basic principles discussed so far. The first is isolation of potential ignition sources by means of enclosures that keep the dust out to the required extent (IP 5X or 6X), the second is prevention of ignition of the actual dust (layer or cloud) by the enclosure surface. This philosophy was up to recently governing both the standardization work in IEC and in Europe. There does not seem to be any valid reason for departing from this simple, sound philosophy as the basis in international standardisation work also in the future. Nevertheless, a new series of dust standards for electrical apparatus has been produced in an attempt to "align" standards for dusts with those for gases.

7.2.3.3.2 Pressurized Enclosures

The basic idea of the Exp standard for gases has been outlined in Section 7.2.1. The controversy presented by this standard when adapted to combustible dusts is discussed by Eckhoff (2003). The dust standard is both superfluous and inherently self-contradictory.

If the interior of electrical apparatus enclosures has to be kept entirely free of dust particles (dusts of abrasive or corrosive materials may damage delicate moving components, or accumulation of layers of electrically conductive dusts may short circuit electric/electronic circuitry), it generally does not represent any substantial difficulties to design dust tight enclosures satisfying the requirements of the enclosure standard IP 6X. In the case of comparatively large apparatus enclosures, e.g. instrument cabinets, with doors and windows fitted with rubber gaskets and locking arrangements, it may be difficult to completely prevent ingress of dust. However, transmission of explosive dust clouds from the outside of the enclosure to its interior, through possible narrow gaps and holes can be entirely excluded. Hence, a pressurization standard for combustible dusts to prevent dust explosions inside enclosures, is superfluous.

The self-contradiction of the standard lies in the following erratic assumption: If particles in a dust cloud embracing an enclosure should enter it though narrow holes and gaps, the particles will stay suspended in the atmosphere inside the enclosure, and eventually form an explosive dust cloud inside the enclosure, as if the dust particles were gas molecules. This assumption is basically wrong, If significant quantities of dust particles do, over some time, enter an enclosure at all, they will not accumulate as cloud, but as a layer. For a more extensive discussion, see Eckhoff (2003).

7.2.3.3.3 Encapsulation by Molding

As part of the effort to align dust standards with gas standards a new standard for encapsulation of electrical equipment for combustible dust atmospheres by molding has been produced. This is a type of protection by which electrical parts that can ignite an explosive atmosphere are molded into a compound material in such a way that the atmosphere cannot make contact with these parts. The compounds can be thermosetting, thermoplastic, epoxy resins, elastomers etc., with or without fillers. It is difficult to see that this comprehensive standard for dusts, which is to a large extent an edited copy of the corresponding gas standard, is very helpful.

As discussed in Section 7.2.1, the basic issue of the molding concept, i.e. to prevent the formation of an explosive atmosphere inside enclosures, is not relevant for dusts. A relevant specific issue with dusts would rather be to make sure that molded components, if embedded in dust deposits, do not give rise to self-heating/self-ignition of the dust deposits.

7.2.3.3.4 Intrinsically Safe Electrical Apparatus

In practice electrical circuits, switches, etc., to be used in areas containing combustible and/or electrically conductive dusts, should always be kept inside enclosures. This will prevent significant quantities of dust from making contact with electrical components in general, including components that may generate electric sparks and/or hot surfaces. Hence, direct adoption of the entire concept of "intrinsically safe design" from the gas/vapor domain into the domain of combustible dusts does not seem to be an optimal approach.

However, there are some highly special applications where there is a genuine need for intrinsically safe apparatus in environments containing combustible powders/dusts. One example is capacitive level indicators for solid bulk materials stored in silos and bins. In this case a live capacitor "plate," in the form of a bare metal rod/rope, carrying a voltage with respect to earth, is directly exposed to the combustible powder/dust inside the silo or bin. An ignition risk could arise from electrical sparks generated by direct contact between the energised bare metal rope and any grounded metal part of the silo.

Figure 7–19 illustrates the application of this type of level indicator. The basic principle of measurement is as follows: A short electric pulse (ns wave package) is emitted from the sensor head at the top of the silo and travels down the vertical metal rod/rope. At the point where the rod/rope becomes immersed in the powder, its impedance changes abruptly, which causes a partial reflection of the electrical pulse from this point and backwards to the sensor head at the top. The distance d from the sensor head at the silo top to the powder surface is then $d = 1/2\ ct$. Here c is the speed of the electromagnetic wave pulse passing along the metal rod/rope, and t is the time from pulse emission at the sensor head till the return of the reflected pulse from the point along the rod/rope where the surrounding medium changes from gas to bulk powder.

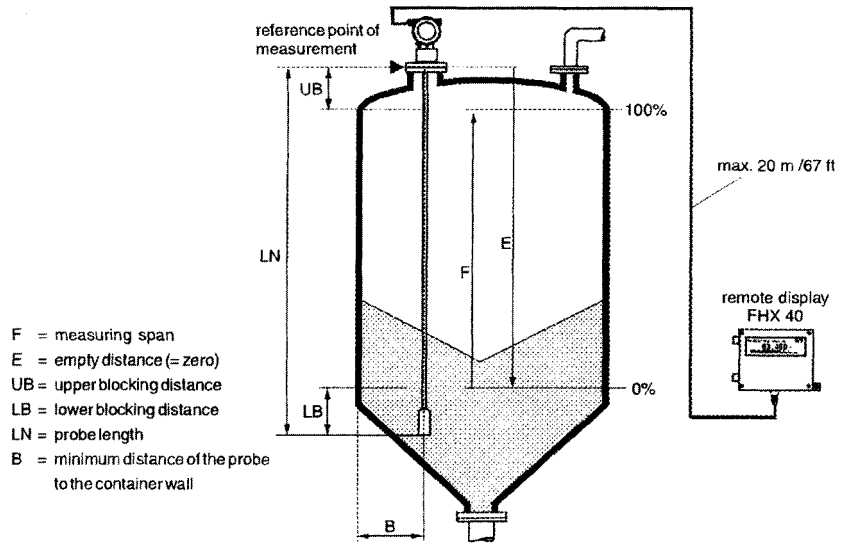

Figure 7–19 Illustration of the use of a capacitive level indicator for powders in silos. Courtesy of G. Klotz-Engmann, Endress + Hauser, Germany.

7.2.4 Explosives/Pyrotechnics

7.2.4.1 CENELEC (1997)

CENELEC (1997) does not specify any specific standards for design of electrical equipment to be used in areas where explosives, propellants or pyrotechnics are produced and/or handled. Instead the standard specifies conditions under which electrical apparatus standards for gases and vapors (Section 7.2.1) or for dusts (Section 7.2.2) apply. The following general guidelines are given:

- In areas endangered by explosive substances only the electrical apparatus which are absolutely necessary for the operation of the electrical installations shall be used. Where this is not practicable, apparatus should be installed in the area with the lowest risk.
- Electrical apparatus shall be selected on the basis of its design or protected by additional structural measures and installed in such a way that in prescribed use the protection against the penetration of

dust and water and against electrical, chemical, thermal or mechanical influences required for operational safety is maintained.

- Cable glands should be either made from a suitable non-corrodible material or be electroplated with e.g. chromium or nickel.
- The design of enclosures for electrical apparatuses should such as to minimize the possibilities of accumulation of dust layers, and facilitate cleaning operations.
- In cases where electrical apparatus in an area E1 is subjected only to accumulation of explosive substances as layers, at the same time as the apparatus is provided with additional devices for preventing accumulation of such substances on and inside the electrical apparatus (e.g. dust covers or additional enclosures), the requirements for area E2 apply.
- Electrical apparatus, with the exception of cables and cords, may be used in general at ambient temperatures up to 40°C. The influence of adjacent heat sources shall be taken into account.
- In cases where ambient temperatures exceed 40°C electrical apparatus that has been specially designed for these conditions shall be used.
- Where explosion-proof electrical apparatus are used, it shall be ensured that the explosion protection as in prescribed use is retained.

7.2.4.2 Australian Interim Standard

The Australian interim standard (2004) specifies the following requirements of electrical equipment to be installed and used in the various zones as defined in Section 7.1.5.6:

- Zone 0E: Exia and other special solutions permitted in zone 0 (see Section 7.2.1)
- Zone 1E: Exia and other special solutions permitted in zone 0, Exib, Exe, Exp, Exo, Exq, Exm (Section 7.2.1)
- Zone 2E: Same as for zone 1E
- Zone 20E: Dust tight enclosures (IP 6X see Section 7.2.3.2.1), Exia, Exm
- Zone 21E: Dust tight enclosures, Exia and Exib, Exp, Exm

- Zone 22E: Dust tight enclosures, Exia and Exib, Exp, Exm
- Zone SE: Normal electrical equipment can be used

The standard also specifies the following maximum permissible surface temperatures of electrical equipment:

- Zone 0E: T5 (see Table 7–4)
- Zone 1E: T5
- Zone 2E: T4
- Zone 20E: T5 or cloud ign. temp. of dust minus 75°C, whichever is the lower
- Zone 21E: T5 or cloud ign. temp. of dust minus 75°C, whichever is the lower
- Zone 22E: T4 or cloud ign. temp. of dust minus 75°C, whichever is the lower
- Zone SE: Normal electrical equipment can be used

Chapter 8

Outline of Methods for Hazard and Risk Analysis

8.1 Introduction

Hazard and risk analysis is a large subject in itself, covered by a substantial quantity of published information. The term *hazard analysis* comprises a number of different systematic methods for identifying the hazards to be associated with a given process or plant. Such analyses can also be used as a basis for optimizing the selection of means for preventing and mitigating dust explosions.

Risk analysis consists of four major steps: identification of a representative set of failure cases, calculation of consequences, estimation of failure probabilities, and assessment of overall impact.

Cox (1986, 1987) gave an informative summary of the various techniques in use for hazard and risk analysis, which is quoted more or less literally, under the following five headings.

8.2 Hazard Surveys or Inventories

These methods are essential preliminaries to many safety studies. The survey consists of making an inventory of all stocks of hazardous material or energy and noting relevant details of storage conditions. When carried out at the conceptual stage of a project, such a survey can contribute to layout optimization and may suggest process changes to reduce stored quantities. It generates information that can be used in a preliminary risk assessment, but

the hazard survey itself is little more than a "screening" exercise designed to identify problem areas.

8.3 Hazard and Operability Studies (HAZOP) and Failure Modes and Effects Analysis (FMEA)

These two techniques have very similar objectives and methods of approach. The purpose is to identify systematically all of the possible ways in which the system investigated could fail and to evaluate these and formulate recommendations for preventive and mitigatory measures.

FMEA is the simpler of the two techniques. The procedure is to take each plant item and component in turn, list all possible failure modes and consider the consequences of each. The results are recorded in a standard format in which recommendations for action can be included. The weakness of FMEA is that there is no specified method for identifying the failure modes and their effects. The engineer is expected to do this from first principles or past experience, and the only discipline imposed on him or her is that of the reporting format itself.

HAZOP overcomes this difficulty by introducing a systematic method for identifying failure modes. This involves scrutiny of a large number of possible deviations from normal operating conditions, which are generated by applying guide words such as more, less, reverse, etc., to each of the parameters describing process conditions in each component, plant item, or line in the plant. However, HAZOP in its original form has disadvantages, and some industrial companies have modified the way in which the results of the study are handled. Instead of "recommendations," the output is "identified problems," leaving more room for a coordinated rational design revision that is not only cheaper, but also probably safer.

8.4 Analysis of Systems Reliability by Fault Tree Analysis

This method is applied to complex systems, whether the complexity is due to the nature of the process itself or to the instrumentation required for running the process. In the basic technique, the "Fault Tree Analysis," the failure modes must first be identified, e.g. by HAZOP. These failure

modes are named "top events." An example of a "top event" could be a dust explosion in a milling plant.

For each "top event." the analyst must then identify all those events or combinations of events that could lead directly to the failure. The precise logical relationship between cause and effect is expressed by AND or OR gates and is usually presented in diagrammatic form. The immediate causes of the top event have their own contributory causes, and these can be presented in a similar way so that a complete fault tree is built up. This process ceases when all of the causative factors at the bottom of the tree are of a simple kind for which frequencies of occurrence or probabilities can be estimated.

The synthesis of fault tree is a job that is best done by an engineer with good experience of the type of system under consideration. It is much easier to teach such a person how to construct a fault tree than to teach a reliability specialist everything about the system. However, the quantitative analysis of a fault tree is a separate activity in which the reliability specialist will play the dominant role.

An illustrative example of a quite comprehensive fault tree for a grain dust explosion in a grain storage facility was given by National Materials Advisory Board (1982).

8.5 Quantitative Risk Analysis by Event Tree Analysis

Quantitative risk analysis (QRA) consists of the following steps.

Failure cases are identified first by establishing the location of the main inventories of hazardous material and then by scrutinizing in detail the process flow and instrumentation diagrams using checklist methods or HAZOP.

Once the failure cases have been identified, the consequences of the failure must be calculated. Event tree analysis is a useful method in this process. An event tree is the reverse of a fault tree, starting with the initial or "bottom events" and exploring all possible "top events" that can result from it. Each outcome has further outcomes and all of these can be related by means of decision gates. At each gate the conditional probabilities must be estimated for each of the alternative branches. On this basis the probabilities of the final hazard, or 'top event', can be calculated.

Criteria have been suggested whereby calculated risks can be judged. Almost all criteria proposed so far are based on the concept of comparability with the existing general risk background. Cost/benefit and "risk perception" arguments have been advanced.

Risk analysis has been criticized by pointing at

- inaccurate mathematical models
- incomplete analysis of actual practical problem
- inaccurate primary failure probability data
- inadequate acceptability criteria
- difficulty of checking final result
- complexity and laboriousness of method

Hawksley (1989) discussed the conditions under which the various elements of quantitative risk analysis are useful in the assessment of risks in practice.

8.6 Safety Audits

Once a plant enters operation, hardware and procedures will start to change from those originally established by the commissioning team. Usually, there are good reasons for this: the plant engineers and operators may find simpler or more economic procedures, and the operational requirements themselves may change. However, it is also quite possible that safety standards fall off with time because experience of satisfactory operation leads to overconfidence and a false sense of security.

For these reasons, safety audits are used in many operating companies. These may vary from a half-day tour by the works manager to a review lasting several weeks carried out by a team of engineers covering different disciplines and independent of the regular operational management of the plant. For the most penetrating audits, the study should not be announced in advance.

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