Tips for Effective and Safe Handling of Solids

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Get a Handle on Solids’ Flow

Properly accounting for how bulk solids actually will flow in a vessel or overall process can be crucial for successful operations. Learning some simple parameters can often provide a good sense of flowability.

By Don McGlinchey, Glasgow Caledonian University

PROPERLY ACCOUNTING for how bulk solids actually will flow in a vessel or overall process can be crucial for successful operations. So, in this article, we will look at two parameters — the Compressibility Index and the angle of repose (see sidebar) — that can help. While neither provides definitive answers about “flowability,” they do give rough guidance about how a material is likely to behave.

However, before we discuss these parameters, it is important to understand bulk density. It is probably one of the most common and widely used of the bulk characteristics. It is employed to determine wall loading in hopper design, to size volumetric feeders, such as screws and rotary valves, to estimate “flowability,” and in many other ways. It is rather unfortunate then that such a useful characteristic is not a constant for a given material. The bulk density of a material is simply the mass of material divided by the volume that it occupies. The density of the particles themselves can be taken as constant; however, the complication arises because the amount of “space” between the particles depends upon how the material has been handled before the measurement. The volume that a unit mass of product can occupy can differ by 50% between the material being in a compressed and a very loose state. Cement, for example, has a compacted bulk density of 1,400 kg/m$^3$ and an aerated bulk density of 1,000 kg/m$^3$. It is obviously important that the correct bulk density value is selected for any calculation.

BULK DENSITY

The full expression for bulk density is:

$$\rho_b = \frac{\text{Mass}_{\text{solid}} + \text{Mass}_{\text{gas}}}{\text{Volume}_{\text{solid}} + \text{Volume}_{\text{gas}}}$$

(1)

For dry bulk solids, the void spaces would usually contain air or some other gas, the density of which can be taken as negligible compared to the density of the solid particles; so, we can approximate:

$$\rho_b = \frac{\text{Mass}_{\text{solid}}}{\text{Volume}_{\text{solid}}}$$

(2)

We can relate this to another common characteristic, voidage or void fraction, which is the percentage of the total volume not occupied by particles:

$$\varepsilon = \frac{\text{Volume}_{\text{gas}}}{\text{Volume}_{\text{total}}}$$

(3)

Again, assuming air or gas in the void spaces and taking particle density as $\rho_p$, we can write:

$$\rho_b = \rho_p \left(1 - \varepsilon\right)$$

(4)

To illustrate the range of values that voidage can take, consider a static heap of mono-sized spheres. If the spheres are in a regular hexagonal packing (the classic “cannon ball” stack), the voidage would be 26%. In contrast, if they were in regular cubic packing, the voidage would increase to 48%. However, even this does not represent the loosest packing possible for large smooth identical spheres. The cannon ball stack gives each ball six contact points, but simple static mechanics requires only two contact points below the center of gravity of the ball for equilibrium. Therefore, it is possible to have a stable structure with far fewer contact points and a resulting increase in voidage [2]. If the particles are irregular in shape, have a size distribution and in some way cohere to one another, the packing arrangement can be very loose and so the voidage can be very large.
Measurement of bulk density is, in theory, quite simple; it only requires a knowledge of material mass and volume and is generally based on one of two techniques.

The first is to weigh out a quantity of material using a simple balance and put this into a calibrated cylinder in much the same way as you would a liquid. If the particulate material is poured into the cylinder, the volume taken up would be of the material in a loose or poured state; the associated bulk density is commonly described as “poured bulk density.” If this same cylinder is then tapped or dropped from a small height onto the bench several times, the volume would likely decrease and the new value is called the “tapped bulk density.” Similar techniques can be used to determine aerated bulk density from a fluidizing column or compacted bulk density from a material placed under load.

The second technique is to fix the volume of the bulk material by filling a cup-like vessel to overflowing and then leveling it with a straight edge. The vessel then is weighed on a balance and the bulk density calculated. This approach gets around some of the problems of trying to estimate the actual level of powder in a cylinder with a surface that typically is anything but flat and seeing through a glass that has become coated in powder. Table 1 lists typical bulk density values for a few common materials.

One possible complication with bulk density measurements is the effect of the porosity of the particles themselves; so, to avoid ambiguity, it is worthwhile stating whether the bulk density value is inclusive or exclusive of closed pores. Confusion could arise if the method of determining particle density does not take account of internal voids. (Using a helium pycnometer, which determines particle density by a measure of displaced gas, may be advisable when porosity is a factor. The gas generally can penetrate open pores as long as these are not comparable in size to the gas molecule but obviously cannot penetrate closed pores.) These differences become important if, for example, we are concerned with surface area available for reaction or the total solids’ fraction available for reaction.

FLOWABILITY BASED ON BULK DENSITY

Bulk density measurements have been used to give some qualitative prediction of the flowability or “handlability” of a bulk solid — that is, some estimate of the likely ease or difficulty in dealing with these materials. One such predictor is the often-quoted Hausner ratio:

\[ H = \frac{\rho_{tapped}}{\rho_{aerated}} \]  
(5)

Another close relative is Carr’s Compressibility Index:

\[ CI[\%] = \frac{\rho_{tapped} - \rho_{aerated}}{\rho_{aerated}} \times 100 \]  
(6)

The percentages provide a means to rank materials:
- 5-15% free-flowing to excellent flow — granules
- 12-16% free-flowing to good flow — powders
- 18-21% fair to passable powdered granule flow
- 23-28% easily fluidizable powders — poor flow
- 28-35% cohesive powders — poor flow
- 33-38% cohesive powders — very poor flow
- >40% cohesive powders — very very poor flow

These relatively quick and easy measurements can be effective in giving some indication as to how powders will likely behave but are by no means comprehensive; exercise some caution if relying only on this information.

<table>
<thead>
<tr>
<th>Material</th>
<th>Tapped bulk density, kg/m³</th>
<th>Poured bulk density, kg/m³</th>
<th>Particle density, kg/m³</th>
</tr>
</thead>
<tbody>
<tr>
<td>Iron powder</td>
<td>3,410</td>
<td>3,360</td>
<td>7,200</td>
</tr>
<tr>
<td>Aluminum powder</td>
<td>1,220</td>
<td>1,095</td>
<td>2,650</td>
</tr>
<tr>
<td>Cement</td>
<td>1,400</td>
<td>1,100</td>
<td>2,700</td>
</tr>
<tr>
<td>Nylon pellets</td>
<td>680</td>
<td>680</td>
<td>1,140</td>
</tr>
</tbody>
</table>

TYPICAL BULK DENSITIES

Table 1. Poured and tapped bulk densities may differ significantly for some materials.
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ANGLE OF REPOSE

Another parameter that is used to determine flowability is the angle of repose, which is defined as the angle of the free surface of a heap of particulate material to the horizontal plane. Unfortunately, we are faced with the same problem that we were with bulk density — the angle of repose is not a constant for a given material and depends upon the method of heap formation. There again are two measurements commonly quoted: the poured angle of repose and the drained angle of repose. The poured angle of repose is the angle measured from a heap formed by pouring material on to a flat horizontal surface (Figure 1). The drained angle of repose is the angle measured on the internal conical face that has been formed when material is drained from a orifice on the flat horizontal bottom of a container (Figure 2). A third angle of repose that you may come across is the dynamic angle of repose, which is the angle to the horizontal of the free surface formed in a relatively slowly rotating drum (Figure 3).

Be aware of several things when using angles of repose: First, the angle formed will depend upon the details of the formation process. For example, the fall height for the poured angle or the orifice size for the drained angle will influence the angle. Therefore, the angle measured is not independent of the measuring apparatus. Second, the same material tested using the three techniques will give a different angle for each (Table 2). The measurements only can be reliably made when using powders that are free-flowing to slightly cohesive and are fairly homogenous. Materials that are a mixture of components or that have a wide size distribution will give angles that are difficult to determine and suffer low repeatability. There also are some uncertainties based on the fundamental physics of the problem, relating to stress history and avalanche behavior [2, 3].
Despite these difficulties, the angle of repose in whatever form can be a useful tool to rank materials. As a rough guide, the relationship between the angle of repose and flowability often follows the structure below:

<table>
<thead>
<tr>
<th>Angle of repose, degrees</th>
<th>Flowability</th>
</tr>
</thead>
<tbody>
<tr>
<td>25-30</td>
<td>very free-flowing</td>
</tr>
<tr>
<td>30-38</td>
<td>free-flowing</td>
</tr>
<tr>
<td>38-45</td>
<td>fair flowing</td>
</tr>
<tr>
<td>45-55</td>
<td>cohesive</td>
</tr>
<tr>
<td>&gt;55</td>
<td>very cohesive</td>
</tr>
</tbody>
</table>

This classification allows us to make some judgement on the likely flow behavior of a material but has very limited value for equipment selection and design. In particular, it is a mistake to use the angle of repose in an estimate of the wall angle required for the converging section of a hopper. However, the angle of repose can serve in some cases to estimate the surcharge (the material at the top of a hopper which forms a “heap”) in a storage vessel or the ground area requirements when forming a stockpile.

**ROUGH GUIDANCE**
The Compressibility Index and angle of repose both give some indication of flowability under different flow conditions, although the applied stresses in both cases can be considered to be relatively low. There is no obvious benefit in combining both test results into a single index value; both may be usefully applied separately to benchmark or rank materials based on known plant performance. For example, if you have experience that a material with a particular flowability value passes through a chute or indeed an entire process without difficulty, then you may expect that a different material with the same flowability value also will not cause problems. (Most times, you will be correct.) However, a material with a worse flowability value needs be treated with more caution. Plants suffering from poor performance require more detailed testing to establish the cause(s) of the flow difficulties or product hangups.

The material in this article has been extracted and adapted from the author’s recent book “Characterisation of bulk solids” [1].

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**REFERENCES**
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Avoid trouble with slurries
Handling slurries (a mix of solids and liquids), should be based on experience and experiments, not theory. Much of the knowledge obtained from pneumatic conveyors and fluidization systems can be used in understanding slurries.

By Thomas R. Blackwood, Healthsite Associates

IN ALMOST all solids processing operations there comes a time when the engineer must deal with a slurry (a mixture of solids and liquids). This can be in a reactor, crystallizer, centrifuge feed tank, or a pipeline. Often slurries can be treated as just another dense liquid, especially when residence time isn’t a concern or horsepower issues have been addressed. What sets a slurry apart from normal liquids is a longer list of physical attributes, including: particle size, distribution, and particle shapes, with the resulting shear, settling, and drag of the particles. Most of the major problems encountered with slurries are due to a poor understanding of these factors, or from poorly obtained slurry experimental data. However, much of the knowledge obtained from pneumatic conveyors and fluidization systems can be used in understanding slurries.

Some of the properties of the liquid become more complicated once it becomes part of a slurry. Viscosity is especially difficult to determine, due to the range of solids concentrations or changes in particle size distribution that can be encountered in normal plant operations. Just as in pneumatic conveyors, the solids can build up in concentration or size along a pipeline or in a tank. Experimental work obtained under laboratory conditions won’t be of much use unless it is obtained over a wide range of conditions and with scaleable tests. Even the type of viscometer used during the test influences measurements.

As an example, obtaining viscosity data requires consideration for particles bridging the gaps in testing equipment. Using viscosity instruments that test a wide range of disk or cone geometries and sizes can eliminate this problem. Particle size and particle attrition can be the primary concern for handling slurries. This may be a lesser problem than in pneumatic conveyors that use gas because of the lower velocities, but it can alter the product or lessen equipment life due to the higher shear rate of the liquid. Attrition products, due to their high surface energy, also may cause excessive build-up on pipe walls, requiring frequent cleaning of the pipes and vessels. Another complication in handling slurries is the reactivity between the solids and the liquid.

While many of the experiences and design rules of solids suspended in air can be applied to slurries, reactivity can lead to scale formation, secondary products of reaction, and agglomeration. This problem is noticeable in a crystallizer. New particles form from a reaction or because of nucleation. Unfortunately, there may be nuclei formed from attrition of larger particles. Newly formed particles tend to be extremely surface-reactive and the source of operational problems, such as scale. Excessive nuclei hinder growth or can cause products of a different morphology to be formed. The principles of good slurry pipeline design, suspension, and handling have been well defined since the early 1950s.

However there have been significant refinements to detailed hydraulic design. In most cases the strategy is to make a settling slurry into a non-settling slurry so that it can be treated as a normal, homogeneous fluid. Settling slurries are more complex, and rely on determination of the settling rate, at least for pipeline design. Several design methods are described in References 1-5 and in some proprietary publications such as Ref. 6. Some authors suggest that there are slurries that can’t be effectively pumped or suspended because they can’t be classified as settling or non-settling slurries [7]. For these, you must rely entirely on previous experience. This article focuses on the basic principles of suspension for solids in vessels and pipes, with an emphasis on practical rather than theoretical considerations.

VESSEL CONSIDERATIONS
Maintaining a uniform suspension in a crystallizer or a slurry tank is impossible, no matter what the theory will tell you. Although colloids and very fine particles can be suspended with appropriate agitation, only a portion of the large particles will be
suspended. Agitators should be selected to provide enough suspension to keep most of the solids off the bottom of the vessel. This is very important when the solids are cohesive. In addition to agitator selection, you must consider tank design, baffle selection, and slurry takeoff points. Many solids can tolerate incomplete suspension with solids accumulating on the bottom of the vessel, but this isn’t good design practice. This is especially true with bottom drain tanks where the outlet can become plugged. Although the solids can be kept off the bottom of the vessel by maintaining a minimum agitator speed, they can accumulate in the bottom drain and nozzles of the vessel. Any design should assume that and provide for back-flushing of the nozzle or dilution of the slurry by having a side port where liquid from the top of the tank can be injected into the nozzle. Specialized valves can minimize the buildup in the nozzle, but aren’t foolproof.

The most common method for estimating this minimum suspension condition is Zwietering’s formula for the minimum agitator speed, \( N \) [8]:

\[
N = \frac{S \, d_p \, \mu_1 \, (g \, \Delta \rho) \, B \, 0.13}{\rho^1 \, 0.55 \, D^0.85} \tag{1}
\]

where:
- \( S \) = Dimensionless parameter from agitator type, and agitator/tank diameter ratio;
- \( d_p \) = Particle diameter, m;
- \( \mu_1 \) = Liquid viscosity, Pa·s;
- \( \Delta \rho \) = Difference in solids and liquid density, kg/m\(^3\);
- \( g \) = Gravitational constant, m/s\(^2\);
- \( B \) = Weight of solids to weight of liquid, %;
- \( \rho_1 \) = Liquid density, kg/m\(^3\); and
- \( D \) = Agitator diameter, m.

A minimum speed doesn’t guarantee that solids won’t be resting on the bottom. Zwietering’s equation assures that settling will be limited to 1-2 seconds of contact. Also, satisfying this condition doesn’t mean that solids are uniformly distributed in the tank. Even low concentrations of solids cannot be considered well mixed. If fact, low concentrations of solids often present the greatest problems in turbulent flow due to the formation of clusters or pockets of solids near the bottom of the vessel.

Tank shape and design details are integral parts of the agitator design. Dished bottoms and baffles help prevent settling in cylindrical tanks. Baffles impart a high vertical component of velocity, help eliminate dead zones, and can allow high agitator speeds before aeration occurs. If baffles are undesirable due to slime build-up or heterogeneous reactions, the agitator should be offset and mounted on an angle to reduce swirl and increase vertical motion (Figure 1: 10 degrees -15 degrees from vertical + offset).

Another option is the use of rectangular tanks since the corners provide baffling, especially for these higher viscosity systems. Agitators and impellers aren’t multi-purpose: choices must be made. While an impeller can handle a variety of services its functionality isn’t limitless. The selection of an agitator requires a balance in power consumption, vendor claims, suspension efficiency, attrition (especially in crystallizers), number needed and type of agitator.

The most common impellers for slurries are marine propellers, 4-bladed/45° pitched turbines, and hydrofoils (usually proprietary designs). Two or more impellers may be required, either on separate shafts or on the same shaft, to provide suspension of solids, as well as mixing or blending of the tank. These impellers may, or may not, be of the same type. Often a choice has to be made between the blending...
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or suspending the slurry, while preserving the desired reaction or products in the vessel. A variety of design procedures have been proposed for specifying impeller diameter and shaft speed for a specific duty. One that is widely available is cited in Reference 9.

An additional consideration for slurry systems is re-starting agitation after a power failure. Agitators are often unable to re-suspend the solids once they have settled. Most pumps are designed for light slurry loads and can’t handle the settled solids, usually due to the high viscosity. Sparging with a liquid is the best solution for agitation. Liquid jets can recover the solids suspension, provided the solids aren’t extremely cohesive. Nozzles should have the capability of either back flow or solids-free liquid injection after power failures. Pumping should be stopped until the suspension is restored unless the pump is designed for such a duty. Positive displacement pumps can handle very high solids concentrations (maybe as high as 95%), but add considerable cost to the project.

**TRANSPORT LINES**

Not maintaining a suspension in a transport line can really ruin your day. While a slurry can sometimes be treated as dense and highly viscous liquid, the particle size and distribution create additional problems. Using a larger diameter line to lower pressure drop will cause settling of the solids and slug flow. This leads to higher pressure drop, abrasion, and excessive mechanical stress on the pipeline. In addition, slime may form on the interior of the pipe. And, abrasion can reduce wall roughness over time.

The easiest way to design a pipeline is to maintain non-settling conditions, similar to the design of dilute-phase pneumatic conveyors. As with pneumatic conveyors, operating too close to the saltation velocity can be problematic, and once a minimum velocity has been calculated, a safety factor of at least 25% should be applied. The saltation velocity is determined through an empirical relationship based on the solids loading, particle size, and physical properties of the system. Practical experience and experimental work are preferable to any theoretical model. One method to estimate saltation is shown in Figure 2 [10] and represented by Equation 2.

While the saltation velocity for the largest particle in the suspension is usually used, particle size distribution may be more important. For instance, a slurry with a large percentage of very fine particles may behave as a single larger particle and overshadow the effect of a few large particles. The fine particles may cluster and cause the larger particles to settle at an even higher rate, which would require a higher slurry velocity. One method to overcome this design
problem is to calculate the loading of each particle size, determine the average particle size of the settling mixture, and then use it to determine the effective saltation velocity.

By using the loading that can be carried by each particle size to establish the particle size distribution of the suspension, the average particle size of the suspension gives a more realistic picture of viscometric and drag forces. In this analysis, the designer must account for the piping layout to minimize settling of the solids since this will change the local solids concentration (usually increase it). Best practice dictates longer pipe runs with the minimum of elbows to prevent settling from acceleration losses at the elbows.

Determining the largest particle in the slurry can be tricky, but practical experience suggests that the particle size representing the upper 10% of the total mass in large particles works best. The exception is for particles not well distributed, or with a range of sizes covering over two orders-of-magnitude. In these cases we have to rely completely on past experience. The finer particles will agglomerate in the wake of the larger particles and hydraulically act as even larger particles (i.e., the small particles carry the larger particles).

The two distinct lines in Figure 2 are for uniform and mixed particle size solids. The wider the distribution, the quicker the correlation deviates from a straight line (when \(W/V \rho_f > 0.1\)). If the particle size varies by only 3 to 4 fold, the following equation, which is valid for uniform particle size, can be used until \(W/V \rho_f > 10\):

\[
V^2/g Dp \rho_p^2 = 0.0556 (W/V \rho_f)^{0.33} \tag{2}
\]

where:
- \(V\) = Superficial fluid velocity at the saltation point, ft/s;
- \(Dp\) = Particle diameter, ft;
- \(\rho_p\) = Particle density, lb/ft³;
- \(\rho_f\) = Fluid density, lb/ft³;
- \(W\) = Solids mass flow rate at the saltation point, lb/s/ft²; and
- \(g\) = 32.2 ft/s².

Defining suspension requirements

Some questions to ask regarding particle suspension are:

*Where is this slurry going?* If a filter or centrifuge will be next, it will be important to maintain the slurry integrity. If a dryer follows one of those, this may be less of a concern; the solids could be re-mixed in the dryer. However, operational problems can result in the filter or centrifuge due to settling of the solids. Not maintaining enough motion of the solids can cause problems with the feed to centrifuges or downstream reactors. One very common complaint when feeding batch centrifuges or filters is the variability of the slurry feed particle size. When the feed tank is full, the centrifuge or filter receives mostly large sized particles and the filtration rate is very good. As the tank empties the particle size becomes smaller, the filtration rate is slower, and cake quality becomes poorer. Eventually, the centrifuge or filter may plug. The root cause of the problem is often oversizing of the feed tank (either too large or too tall for the agitator height). One easy fix when this problem is encountered is to recirculate the feed to the centrifuge or filter back to the feed tank so that flow back to the tank is maintained at all times.

*How variable is the particle size distribution?* Many designs are based on the largest particles and fail to account for particle distribution. Fine particles can get into the boundary layer of the flow, increasing the shear rate and overall pressure drop. Particles also may settle in the boundary layer and sluff off, disrupting the laminar layer or turning laminar flow into turbulent flow. While the resulting pressure drop may decrease, the return of laminar flow may cause solids to separate and become a slugging or erratic flow. This may turn a non-settling slurry into a settling slurry, where different design rules apply. Large concentrations of fine particles also may turn a normal non-settling slurry into a settling slurry due to hindered settling effects. The lower drag from local turbulence causes fine particles to cluster and act as a much larger particle.

*What are the consequences of poor suspension?* Most crystallizers will still make an acceptable product even without fully suspended solids. Attempts to mix the suspension further may not be desirable due to attrition or power consumption. It may be more inexpensive to install several re-suspension points on a long pipeline rather than take the chance that a change in the particle size distribution may cause a problem. Can fluid injection along the line or in a vessel return a plugged system to operational? If so, it may be more inexpensive to plan for blockages than
to try to design around them. When faced with a new system or a large scale-up situation, this may be the best course of action.

What experimental data are needed for design? If you don’t have an existing slurry system to draw from, the design will be developed from physical property data and hydraulic models. It will be important to determine the settling characteristics of the slurry and determine to what extent it can be treated as a non-settling slurry. These data are needed in addition to the physical properties and the rheology (Newtonian, shear-thinning, shear-thickening, or Bingham plastic) of the slurry. More effort must go into defining the viscosity, particle size, and variability of the slurry than in scaling up an existing system. However, scale-up of an existing system must insure that settling of the slurry is properly handled, such as accounting for the effect of larger diameter pipe on saltation.

How will the system (pipeline, agitation, etc.) be controlled? While not a focus of this article, there have been many improvements in on-line instrumentation for slurries (see sidebar); these should be a major part of the design. Many chemical plants lack adequate instruments for solids or slurry flow because these were expensive, unreliable, and the need wasn’t well understood many years ago. Slurry density can be controlled more precisely and blockages eliminated through simple non-intrusive flow meters or velocity sensors. At a minimum, sample ports must be provided to allow for determination of slurry characteristics after the design is installed. The experimental work should have identified and incorporated some key process variables (shear rate, density, and particle size) into the overall control strategy.

SUCCEED WITH SLURRIES
Designing reliable slurry-handling systems starts with an understanding of the slurry properties, especially settling rates, viscosity and density. It is rare that physical properties can be obtained from empirical correlation or existing databases, as may be the case for single-phase liquid mixtures; testing is required before design. Even scale-up from an existing installation may be difficult due to the variability of settling characteristics. However, an existing installation is often a better source for design than a theoretical design using standard viscometric design methods.

With the correct attention to obtaining the physical properties, consideration of adequate instruments, and an understanding of the principles of slurry design, there is little reason that operators should have problems handling slurries.

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REFERENCES
In any process challenge, the difference between the possible and the impossible is knowledge. Put Hapman’s 60 years of industry experience and material handling knowledge to work for you and discover the possibilities.
Explosion Protection System Selection

Handling slurries (a mix of solids and liquids), should be based on experience and experiments, not theory. Much of the knowledge obtained from pneumatic conveyors and fluidization systems can be used in understanding slurries.

By Emre Ergun, MS, MBA, FENWAL Explosion Protection Systems

SELECTING THE right explosion protection system for your combustible dust handling process can be a daunting task considering the availability of many explosion protection technologies. Over the past decade with the introduction of new suppressor and detection technologies with smarter features along with new passive protection products, the number of available explosion protection options has increased. Combined with increased emphasis on combustible dust hazards by OSHA due to recent high profile combustible dust explosions, understanding the available explosion mitigation options has become even more important. Process owners, corporate risk managers and process engineers need to understand the available protection options and applicable regulations in order to address combustible dust hazards within their processes using the most applicable and cost effective mitigation technology.

This paper provides a brief introduction to the explosion protection methods including suppression, isolation and venting. Other protection options such as pressure containment, combustible concentration reduction and inerting are also available, however not within the scope of this paper.

OVERVIEW OF EXPLOSION PROTECTION OPTIONS

Before proceeding with the selection process, let’s briefly look at the various explosion protection options:

Explosion Venting. It is one of the most widely used techniques for mitigating dust explosions. It is based on the opening of a specifically designed weaker membrane installed on a vessel to open and direct the flame, burned/unburned material, overpressure and other combustion byproducts to a safe location during the incipient stages of an explosion. Explosion relief vents are designed to ensure that the pressure rise due to explosion does not exceed the pressure shock resistance of the protected equipment. In the US, explosion relief vents are designed in accordance with the procedures given in NFPA 68.

Flameless Venting. This relatively new technology is finding niche applications in the process industries for protection of indoor process equipment against dust explosions. Flameless vents combine the proven explosion relief venting technology with flame arresting and controlled particle retention. Similar to explosion relief venting, its operation relies on the rupture of a weak membrane during a deflagration event where the overpressure, flame and particulate material discharges through a flame and particulate trap material, thus eliminating the flame and material discharge to surrounding areas. Only hot gas and pressure are discharged to ambient.
According to NFPA 654, vented equipment needs to be isolated in order to prevent flame propagation between connected vessels. Further information on isolation methods is provided in the following sections of this paper.

**Explosion Suppression.** As a well proven technology, explosion suppression systems have been employed in the industry since late 1950s. It is often used where it is not possible to vent the contents of process equipment to a safe place, and where the material being handled is toxic or is a harmful material to people and environment. Deflagration suppression requires that the incipient explosion be detected very soon after ignition, and a sufficient amount of suppressant is discharged into the developing fire ball in the enclosure at a fast enough rate to extinguish all flame before a destructive overpressure develops. Figure 1 below shows the basic steps in deflagration suppression. In the US, NFPA 697 governs the standards on explosion protection systems.

**Explosion Isolation.** Very often process equipment handling particulate solids are connected to each other by ductwork, chutes and conveyors. A deflagration in a vessel can propagate through these connections to upstream or downstream equipment and cause subsequent explosions in the adjacent equipment. Explosion isolation systems are employed to mitigate the flame propagation and pressure piling between connected equipment. According to NFPA 654 Section 7.1.4, where an explosion hazard exists, isolation devices shall be provided to prevent deflagration propagation between pieces of equipment connected by duct work. Some of the commonly used explosion isolation systems are:
Chemical isolation. It is achieved by rapid discharge of a chemical extinguishing agent (e.g. sodium bicarbonate) into interconnecting duct to mitigate flame propagation.

Mechanical Isolation. High-speed gate valves and flow-actuated valves are some of the examples of mechanical isolation products. Active isolation devices such as high speed gate valves close in milliseconds upon detecting an explosion, and mitigate flame and pressure propagation inside ductwork by rapidly deploying a mechanical barrier. Passive isolation devices, such as flap or float type valves are self actuated by the air-flow created by deflagration event, thus require no detection or control components. Passive isolation devices are typically used to isolate nuisance dust handling equipment with relatively low dust loads.

EXPLOSION PROTECTION SYSTEM SELECTION

Considering the large number of explosion protection options available, we developed the following selection guide to illustrate the commonly employed explosion protection options for process owners. Throughout this paper it is assumed that the process owner has already performed a risk assessment, and determined that an explosion protection system is required as one of the means to mitigate combustible dust hazards. Thus our focus in this paper is limited to explosion venting, suppression and isolation.
1. VENTED DUST COLLECTOR WITH FLAP TYPE PASSIVE INLET ISOLATION
This is one of the most cost effective protection options, utilizing a complete passive protection method. In this case, dust collector is located outside, and does not handle any toxic or other harmful materials. Due to valve's design limits, the KST12 of the dust is limited to maximum 300 bar.m/sec, the inlet duct diameter is maximum 22 inches, and the maximum reduced pressure (PRED) is limited to 7 psi. There will be flame and material release due to opening of the venting device during a deflagration event. The periodic maintenance of this system can be done by the system owner. This protection scheme meets NFPA 68, 69 and 654 requirements.

2. FLAMELESS VENTED DUST COLLECTOR WITH FLAP TYPE PASSIVE INLET ISOLATION
In this case dust collector is located indoors, and it does not handle any toxic or other harmful materials. There is sufficient space around the vent area to allow 5 meters of safety perimeter. Due to flameless vent's design limits the maximum allowable KST of the dust is 250 bar.m/sec. And due to flap valve's design limits, the inlet duct diameter is maximum 22 inches, and the maximum reduced pressure (PRED) is limited to 7 psig. The isolation distance is up to 4 meters from the protected vessel. There will be no flame release however some hot-gas will release due to opening of the venting device during deflagration. The periodic maintenance of this system can be done by the system owner. This protection scheme meets NFPA 68, 69 and 654 requirements.

3. VENTED DUST COLLECTOR WITH CHEMICAL INLET ISOLATION
In this case, dust collector is located outdoors, and it does not handle any toxic or other harmful materials. There is no restriction on the size of inlet duct diameter. The maximum reduced pressure (PRED) can be higher than 7 psi. The maximum allowable KST for this system is 500 bar.m/sec. There will be flame and material release due to opening of the venting device during deflagration. The periodic maintenance of this system can be done by manufacturer or by manufacturer trained system owner. This protection scheme meets NFPA 68, 69 and 654 requirements.
4. SUPPRESSED DUST COLLECTOR WITH CHEMICAL INLET ISOLATION
In this case, dust collector can be located indoors or outdoors. There is no restriction on the type of material being handled. Also there is no restriction on the size of inlet duct diameter. The maximum reduced pressure (PRED) of the vessel can be higher than 7 psi. The maximum allowable KST for this system is 500 bar.m/sec. There will be no flame or material release after actuation of the system during a deflagration event. The periodic maintenance of this system can be done by the manufacturer or by manufacturer trained system owner. This protection scheme meets NFPA 69 and 654 requirements.

5. SUPPRESSED DUST COLLECTOR WITH CHEMICAL INLET AND OUTLET ISOLATION
In this case, dust collector can be located indoors or outdoors. However the clean-air exhaust is re-circulated back into the building. There is no restriction on the type of material being handled. There is no restriction on the size of inlet or outlet duct diameter. The maximum reduced pressure (PRED) of the vessel can be higher than 7 psi. The maximum allowable KST for this system is 500 bar.m/sec. There will be no flame or material release after actuation of the system during a deflagration event. The periodic maintenance of this system can be done by the manufacturer or by manufacturer trained system owner. This protection scheme meets NFPA 69 and 654 requirements.
6. SUPPRESSED DUST COLLECTOR WITH CHEMICAL INLET AND OUTLET ISOLATION FOR HYBRID MATERIAL APPLICATIONS

This is a case commonly seen in pharmaceutical or specialty chemical process applications. Dust collector can be located indoors or outdoors. There is no restriction on the type of material being handled. However the material is a mixture of flammable vapor/gas and combustible dust. There is no restriction on the size of inlet or outlet duct diameter. The maximum reduced pressure (PRED) of the vessel can be higher than 7 psi. The maximum allowable KST for this system is 500 bar.m/sec. There will be no flame or material release after actuation of the system during a deflagration event. The periodic maintenance of this system can be done by manufacturer or by manufacturer trained system owner. This protection scheme meets NFPA 69 and 654 requirements.

FINAL THOUGHTS

This paper provides a brief introduction to the commonly employed explosion protection technologies and illustrates the system selection process. Using the process provided in this paper, process engineer or risk manager can make informed decision on their explosion protection requirements. However it is important to note that selection of explosion protection system is application specific, and system design expertise is required to select and size the most appropriate system. The number and size of suppressors, their installation locations and explosion detector settings should only be determined by the application engineer of the system manufacturer using computer aided design methods specific to the system being considered. Thus contact your explosion suppression system supplier for more information on the products and system design criteria.

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10. Siwek, R., Explosion Protection Systems Classification and Design
11. A Tri-fold brochure illustrating the explosion protection options is available at www.fenwalprotection.com
12. KST is the deflagration index of a dust, and is a measure of the relative explosivity of the material
13. PRED is the maximum pressure developed in a protected enclosure during a vented or suppressed deflagration.
Put your critical processes inside the Fenwal Circle of Protection

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Evaluation of Dynamic Tail Pipe Loads during Relief Valve Discharge with RELAP5

By Jens Conzen and Jaehyok Lim, PhD, Fauske & Associates, LLC

PROCESS PIPES in chemical plants often carry a high amount of energy. During the undesirable event of an over-pressurization of the system, caused by an unwanted reaction or transient for example, pressure safety valves (PSVs) will open to reduce the increasing pressure. A fast depressurization transient may occur that triggers a pressure wave that travels at the speed of sound through the fluid. This can lead to force imbalances in particular on long pipe segments. Industry experience has shown that these forces can grow large enough to damage pipe supports and possibly the process pipe itself. In case a pipe break occurs, this force is even further increased by the resulting thrust load.

It is therefore of great importance that these hydrodynamic loads following the opening of a PSV are calculated to assure a safe design of the piping system. Recent industry accidents have shown that in particular the PSV tail pipe is often not sufficiently restrained.

Bounding steady state values can be calculated analytically by using correlations based on choked flow theory. Another way of calculating these loads is to use a software solution. A computer code has the advantage that it can solve for time history data (force versus time) by numerical integration. The time dependent data can be used to assess the structural response of pipe supports and the piping system. If the transient is fast, the structure might not fully absorb the peak load due to its inertia, which would provide the designer with more margin. For a dynamic stress analysis, a separate finite element or piping analysis package is required.

The RELAP software can be used to analyze fluid transients and compute resulting loads. RELAP stands for Reactor Excursion and Leakage Analysis Program and is a light water reactor transient analysis code developed for the U.S. Nuclear Regulatory Commission (NRC) for simulation of a wide variety of hydraulic and thermal transients in both nuclear and nonnuclear systems involving mixtures of steam, water, noncondensable and solute under single phase and two phase conditions.

In a typical analysis with RELAP the piping system is finely nodalized to allow the detailed computation of loads on segments throughout the system. Depending on the node size the time step size must be below the acoustic courant limit of the system to assure that the pressure wave can be captured. The pipe reaction force is then calculated as a post-process activity. The development of the transient force time history is based on the general force equations for a container (Moody, 1990). The open pipe reaction force during unsteady discharge can be written as:

\[
F = -\left\{ (P_v - P_a)A_o + \frac{m_v V_v}{\theta_c} \right\} + \left( -\frac{1}{\theta_c} \right) \frac{d}{dt} \rho AV dx
\]
It is the summation of differential pressure force, momentum flux force, and the wave load. The wave load is the unsteady force caused by the rate of fluid momentum change within a pipe segment. The calculation performs a two fluid treatment with the consideration of vapor and liquid components.

Figure 1 shows an example system that has been analyzed with RELAP and Figure 2 shows the transient force time history on the tail pipe (pipe 157 with 5 nodes). The reaction force can be computed for any pipe segment in the system. The black line shows the total force which is summation of momentum flow (green), wave load (red) and pressure force (blue). The time scale shows that the force peak is of rather short duration before the depressurization transitions into a steady discharge.

The application of RELAP5 could be further extended to carry out scoping analysis of the PSV performance. A high momentum flow and velocity in the tail pipe can potentially lead to high vibrations and noise levels. For that reason the tail pipe diameter should be larger than the process pipe diameter. The software can compute the Mach number at any point in the tail pipe, which can be used as a design parameter. It is further possible to identify where choked flow occurs (at the PSV exit or at the tail pipe exit) or whether or not a second choke point is created based on different tail pipe diameters.

The evaluation of pipe loads is a solid part of our daily business. Beyond that RELAP5 has found a broad range of applications within FAI such as fluid dynamics experiment design or sizing analysis of plant components, for example. Several of our engineers are proficiently trained with the software.

REFERENCES
RISK MANAGEMENT SERVICES

COMBUSTIBLE DUST EXPLOSION AND FIRE HAZARDS
Onsite dust hazard assessments, OSHA Combustible Dust NEP compliance and other services related to characterizing, preventing and mitigating combustible dust explosion and fire hazards

OSHA PROCESS SAFETY MANAGEMENT (PSM)
Services for all 14 elements of OSHA PSM including PHA as well as human factors and facility siting

SAFE PROCESS SCALE-UP
Process safety to support chemical development programs for facilities including kilo lab, pilot plant and commercial scale plants

PHA/LAYERS OF PROTECTION ANALYSIS (LOPA)
PHA for all phases from design through production. Often paired with PHA, LOPA is a simplified form of risk assessment and assists in PSM and RMP compliance. LOPA helps direct limited resources to the most critical safeguards

INCIDENT INVESTIGATION
Expertise in discovering possible conditions leading to a recent explosion, fire, incident or near miss, performing root cause analysis and testing to include dust explosion severity, flammability and chemical reactivity as well on-site investigation services

PROCESS SAFETY DUE DILIGENCE FOR MERGERS AND ACQUISITIONS
Compliance reviews per regulations and guidelines, gap analysis and prioritized recommendations for future effort

RELIEF SYSTEM DESIGN REVIEW
One of the principal scientists involved in the Design Institute of Emergency Relief Systems (DIERS), Dr. Hans K. Fauske continues to contribute to developments in relief system design technology

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Pump Up the Volume with Advanced Bin Level Measurement

By Jenny Nielson Christensen, MBA, Director of Marketing, BinMaster Level Controls

MONITORING INVENTORY levels in bins, tanks and silos has become increasingly challenging for companies that produce goods, the vendors that supply raw materials, and the solution providers that help companies monitor their inventories. Many processors deal with high value materials and no doubt the cost and value of raw materials, as well as inventory carrying costs are increasing. Suppliers to processing companies often share in the monitoring of material inventories and are partners in helping their customers manage just-in-time inventory, optimizing inventory turns and controlling costs. Companies that manufacture systems that help processors and suppliers manage their inventories are designing solutions to meet the evolving requirements of level sensors, creating advanced communications networks, and software that makes data easily accessible in real time.

When it comes to measuring powders and solids in bins, tanks and silos, plant personnel are often faced with complex issues such as:

- Inventory must be more accurate
- Material volume as well as level is needed
- The material surface being measured is uneven
- Powders and solids in the vessel are prone to buildup
- There are high levels of dust present
- Devices must be highly reliable with minimal maintenance

Over the past few years, acoustics-based level measurement is one technology that has emerged as a proven solution to address these issues. The sensing device – more commonly referred to as a 3DLevelScanner – is coupled with a suite of solutions that can accurately measure most types of powders and bulk solids in challenging environments. It delivers both level and volume data using intuitive monitoring software that can be customized to meet the needs of multiple departments within an organization or shared with vendors who are partners in managing the inventory of materials.

MULTIPLE-POINT MEASUREMENT FOR BETTER INVENTORY ACCURACY

The non-contact, acoustics-based technology used by the 3DLevelScanner sends pulses in a 70° beam...
angle, taking multiple measurements from the material surface. Most devices in use today take only a single measurement and dependent upon where the measurement is taken, the diameter of the vessel and the degree of irregularity of the material surface, a single measurement point may not accurately represent the true level of material in the vessel. Unlike single-point devices, the 3DLevelScanner continuously maps the surface of the material to account for changes in level, overall volume and surface topography. The software reports the lowest point detected, the highest point detected and the average level based upon a weighted average of all of the measurements detected in the bin. The data is used to provide the user detailed information about the level of material in the bin and calculate an estimation of the volume of material in the bin.

REPORTS BOTH LEVEL AND VOLUME
Advanced algorithms in the processing firmware are used to convert the detailed level measurement data to a volume estimate – such as tons, pounds, cubic or metric feet – for the vessel. In many cases, especially in applications with powders that are prone to irregular material surfaces or sidewall buildup, there will be points in the vessel that can be significantly lower or higher than the majority of the contents. By detecting irregularities in the material surface, varied topography and excessive buildup can be accounted for in volume calculations. If a simple average formula was used to determine the average height of the material, it could be inaccurate. By using an algorithm that bases the average height from all of the points and the weights associated with them to determine the average volume and height/distance, the 3DLevelScanner can provide an estimation of bin volume that is more accurate than a volume estimate offered by any single point measurement device.

3D VISUALIZATION OF UNEVEN MATERIAL SURFACES
The 3DLevelScanner also offers the ability to generate a 3D representation of the material contents. By taking multiple measurements within the bin and then mapping the topography in the bin, the computerized profile created by the 3DLevelScanner can show high and low spots in the vessel as well as material built up on the sides of the vessel. With single point devices, a measurement may show the bin is almost empty, even when a significant amount of material remains in the bin. This visualization feature also helps alert maintenance management to the need for bin cleaning at the optimal time.

The image on the left shows the irregular material surface during the empty cycle; the image on the right is the 3D visual representation created by the software.
PERFORMS IN HIGH LEVELS OF DUST
The 3DLevelScanner uses a very low frequency acoustical signal to penetrate dust and take measurements that are determined by how long the signal takes to reach the surface of solid or powder material and return to the device. These very low frequency acoustical signals are able to penetrate suspended dust, unlike other technologies whose signals become unreliable or inaccurate when attempting to take measurements in dusty environments. The acoustical signals – which make a chirping noise – combined with a non-stick material in the horns of the sensor, prevent material from adhering to the internal workings of the device ensuring long-term reliable performance. The 3DLevelScanner is self-cleaning, offering very low maintenance in even the dustiest environments.

HANDLES VOLATILE OR HARSH ENVIRONMENTS
A combination of dust, humidity and heat may be present in vessels storing chemicals. A 3DLevelScanner with an extended operating temperature range of up to 250°F (120°C) can accommodate these higher temperatures without sacrificing measurement accuracy. Materials that have been heated in the production process may be conveyed into the silo when they are still relatively hot. Unlike some non-contact devices that are prone to becoming unreliable under challenging conditions, the 3DLevelScanner designed for harsh environments will maintain a high level of measurement accuracy. It will perform in a wide variety of materials including powders and fine granular substances such as alumina or silica.

The outside of the sensor is coated with dust. Inside the transducers are clean and fully operational.
ACCURACY IN WIDE DIAMETER OR LARGE BINS
Suppliers and processors of materials such as alumina, bentonite, lime, silica, or sodium may store these and other bulk materials in vessels that are very wide or large. As the 3DLevelScanner measures and maps the material surface based upon a 70° beam angle, a single 3DLevelScanner provides the most accurate data in a vessel up to 45’ in diameter. If an operation has wider vessels or the vessel is not at least as tall as it is wide, the 70° beam angle may not detect the entire material surface.

To achieve a high level of accuracy for large diameter bins or bins that are not very tall, multiple 3DLevelScanners can be mounted on the same vessel at locations that allow the sensor to more completely cover the material surface. Often two 3DLevelScanners – one mounted closer to the perimeter and one closer to center, but away from the fill stream – will provide the desired level of accuracy. However, a single scanner can be used in a very large bin if the ultimate goal for the system is not precise volume accuracy, but rather continuous, reliable level indication.

REAL-TIME MONITORING OF MULTIPLE VESSELS
Many processors have multiple types of bins, tanks and silos containing a variety of materials on site. Different people and departments across the facility may have need for bin level or volume data. Plant managers, production, maintenance, purchasing, logistics and accounting personnel utilize inventory data to perform their jobs. Highly accurate, timely data helps them streamline production, purchase and deliver optimally, and report fiscally. Advancements in software allow multiple users from multiple departments to quickly and simply view the data for all the vessels at the site or just selected vessels on a single screen.

Customized views allow users to quickly focus on those bins that need their attention. Data is available 24/7 and in real-time to help users optimize operations and make better decisions based on timely, accurate data. For example, production personnel may be concerned with bins whose levels are low and may need replenishment, or are reaching capacity and need to stop filling so as not to overfill and waste or damage material. By simply clicking on a single bin while in the multi-bin view, a user can view the detailed information for that bin and view the 3D profile. People from different departments can view the data simultaneously. Hosting the data on a single server assures it is accurate and all from the same database.
3DLevelScanner

Non-Contact, Dust-Penetrating Level and Volume Measurement

- Multiple-point measurement
- Detects and maps uneven surfaces
- Creates visual representation of contents
- Works reliably in dusty environments
- Self-cleaning, minimal maintenance

Unsurpassed Inventory Accuracy

MVL Multiple-Scanner System

Accuracy in large bins of powders and solids

3DMultiVision Software

Monitor all your bins in a single window from a PC