THE PARTICLE TECHNOLOGY FORUM (PTF) NEWSLETTER

An American Institute of Chemical Engineers (AIChE) Forum

Message From The Chair

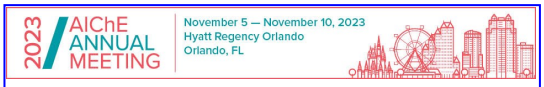
Greetings and Celebration of 30 Years of PTF Community Achievements

As we move into the summer season this year with the promise of normality, I hope that you and your family are enjoying the summer season.

It is now time to register for the AIChE Annual Meeting in Orlando, FL! We have the PTF Awards Dinner on Wednesday evening at Rodizio Grill - Pointe Orlando, social interactions during our poster session and planning meetings, and invited talks by some of our awards winners. A hearty congratulations to all of our award winners! You can read about them in the Fall Issue and hear about their research at the conference in person. I will have much more to say about this year's winners at our PTF Awards Dinner.

These are exciting times for particle technology members playing a pivotal role in both old and new energy sectors. To attract the undergraduate and graduate students, we initiated special PTF Webinar series addressing some of the new energy fields. My gratitude goes to our first speaker, Mr. Michael Molnar for kicking off this series with a first talk on Solar grade polysilicon. It was well received with an attendance exceeding 75 people. I am quite pleased with the success. I am looking forward to the next exciting webinar. We are soliciting ideas and recommendations for the speakers for future webinars in the emerging technologies, so if you have recommendations, please reach out to me.

A Peek At The Contents



AIChE Annual Meeting

November 5 - 10, 2023

Orlando, Florida (USA)

Registration Is Now Open

Registration

Technical Program

Accommodations

Based on input from the PTF community, we have also launched new initiatives to increase diversity, equity, and inclusion within our organization.

My gratitude to this newsletter's editor, Dr. Shrikant Dhodapkar, for creating informative newsletter, Dr. Ben Freireich for planning all PTF sessions, and all members of the executive committee who keep programming and events running smoothly. This years, we have three special sessions with invited speakers. Thanks to Prof. Maria Tomassone, Vice-Chair, who has been taking care of the PTF awards.

We are entering the period where we nominate and elect new leadership for the Particle Technology Forum Executive Committee (EC). We need to elect two industrial and two academic positions on the EC. If you have interest or questions, please contact me or any of the EC members. Nominations are welcome.

Hope to see you in person at the AIChE Annual Meeting - November 5-10, 2023, Hyatt Regency Orlando, Orlando, FL, USA. Remember to purchase the PTF dinner tickets as part of the registration.

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PSRI Particulate Solid Research, Inc.

EDITORIAL

The Particle Playground

A Sandbox Large Enough For Everyone

Ray Cocco and I are excited to introduce a new column to the PTF Newsletter called "The Particle Playground." This feature will be designed to provide a new way to communicate within the PTF Newsletter by focusing on educating, exploring, revealing, and, most importantly, reimagining various topics related to particle technology (particle synthesis and characterization, particle fluid hydrodynamics, particle processing and handling, nanoparticles, and energetic particles & so on…). Our vision is that such a platform should spark ideation and solutions in addressing our global challenges.

In the coming decades, our profession has to build upon fundamental technical knowledge gained during the past century to address newer global challenges, including sustainable energy and feedstock, circularity, and decarbonization with newer concepts & tools such as process intensification, artificial intelligence, and data science and analytics. The pace of innovations must be accelerated with fewer resources and with greater agility to meet the challenges ahead. This can be realized only if the technical community of thinkers and practitioners can meet at the common playground to share their knowledge and experiences as a team. **We invite all members of the PTF to actively participate by proposing topics and write concise, thought-provoking articles to educate, explore, question, reveal, communicate, and share learnings that others can build upon.** We hope that it will become a continuing feature of the PTF Newsletter. All of this starts now with our first Particle Playground article on **particle size distributions**. A deceptively simple topic with many subtle underlying questions.

Let us begin…

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DOW®

Perspective: Modeling and Simulation of Gas-Particle Flows

Professor Sankaran Sundaresan

Winner of 2022 PTF Lifetime Achievement Award

Norman John Sollenberger Professor in Engineering,
Department of Chemical & Biological Engineering,
Princeton University, NJ

Research Interests:
Multiscale flow structures in multiphase flows, role of interparticle forces on gas-particle flows, dry powder inhalation

Over the past few decades, there has been remarkable growth in the use of advanced experimental characterization and large-scale simulations to understand the physics leading to structures in gas-particle flows. Can we calibrate all the important physical properties of the gas-solid system of interest through a few well-defined and widely accepted measurements in small laboratory devices and then rely on simulations to probe the flow behavior of commercial-scale devices? Unfortunately, not yet; however, there is hope that we will get there soon. This article discusses fruitful approaches to getting there.

Most gas-fluidized bed reactors used in industries operate in the turbulent fluidization regime or the circulating fluidized bed mode. In these regimes, mesoscale structures that take the form of bubble-like voids in dense regions with clusters and streamers of particles in the dilute regions, readily formed. Both of which giving rise to persistent fluctuations that span a wide range of length and time scales [1]. The presence of boundaries further complicates the flow behavior, causing macroscale inhomogeneities that are present even in a time-average sense. Higher concentration of particles in the vicinity of bounding walls in riser flows, which could lead to the downflow of particles (and, sometimes, the gas) in the wall region and upflow in the core region, is common in both riser flows and turbulent fluidization. Such macroscale inhomogeneities affect gas and particle residence time distributions in the device and interphase contacting. As a result, how these flow characteristics change upon scale-up and process modification and how they affect reactor performance have been topics of many studies.

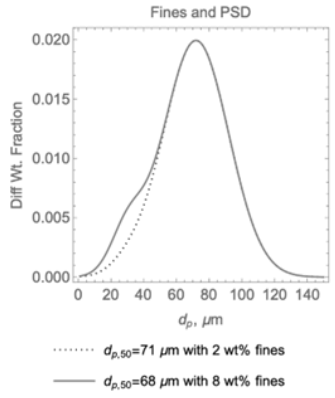
Over the past few decades, there has been remarkable growth in the use of advanced experimental characterization and large-scale simulations to understand the physics leading to structures in gas-particle flows. Many PTF members have contributed immensely to this effort; the list includes many Elsevier Lifetime Achievement Award recipients (Liang-Shih Fan, (the late) John Grace, (the late) Dimitri Gidaspow, Ted Knowlton, Joachim Werther, Jesse Zhu, Jennifer Sinclair Curtis, and Madhav Syamlal) and Thomas Baron Award winners (Roy Jackson, Aibing Yu, Hamid Arastoopour, Rodney Fox, and Christine Hrenya).

It is hoped that the confidence in our ability to model and simulate gas-particles flows will grow to the point where one can calibrate all the important physical properties of the gas-solid system of interest through a few well-defined and widely accepted measurements in small laboratory devices and then rely on simulations to probe the flow behavior commercial-scale devices – at least, to screen various design options. Are we there yet? Unfortunately, not yet; however, there is hope that we will get there soon. In this article, I share my thoughts on the adequacy of the models and simulations and point to fruitful approaches to

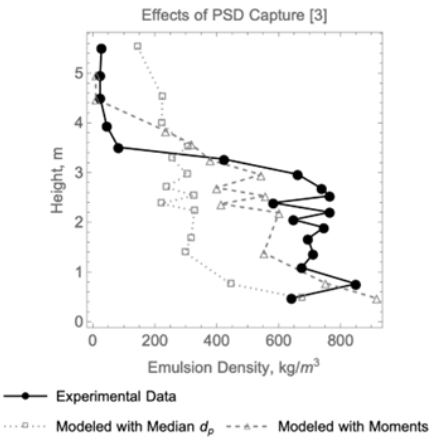
getting there. For simulations to become reliable, the underlying models should be reasonably complete; and for the simulations to be used widely, they must be fast and affordable.

Critiquing the adequacy of multiphase flow models

The results gathered by simulating multiphase flow models are only as good as the physics included in the model. Most models capture some physics accurately, approximate some and ignore others. For example, the transport models conserve mass and energy, and respect Newton's law of motion. They use approximate constitutive models for some of the quantities appearing in the balance equations while ignoring others. Ultimately, the uncertainty associated with the simulation results is determined by (a) the adequacy of the constitutive models, and (b) the compromises made to render the simulations affordable.



Particle size distribution (PSD) is a common metric in fluidized beds. Fine particles are known to influence the quality of fluidization of Geldart Group A particles. PSD can change with time through processes such as attrition or agglomeration. Do the models capture the whole PSD? If they do not, what can we learn from simulations based on simplified models where only one representative particle size (i.e., Sauter mean or median particle size) is used?



For example, the Euler-Euler two-fluid model (TFM) simulations typically consider a single mean particle size and ignore all inter-particle forces (other than the collisional and frictional contact forces). The commonly advanced argument justifying the use of the simplified model is as follows: *The known mesoscale structures can be captured in simulations that consider only uniformly sized particles. As most macroscale structures are induced by the existence of mesoscale structures, even simulations with a single average particle size should reveal the key flow characteristics*.

One can counter this argument by observing that the addition of fines can change the flow behavior in fluidized beds and circulating fluidized beds appreciably, and the extent of this modification may not be captured by making small changes to the median particle size (i.e. $d_{p,50}$). Given this concern, we must *develop the criteria that must be satisfied for the single-particle-size approximation to be acceptable*.

Of course, a more desirable alternative would be to account for the PSD explicitly. For the Euler-Euler multi-fluid models (MFMs), the particles in the simulations are divided into many "fluids" (e.g., through the direct quadrature method of moments [2]). However, such simulations can become unwieldy when three or more particle sizes are considered. Models such as CFD-DEM [4] and MP-PIC [5] are based on a Lagrangian-Eulerian framework, and the whole PSD can be accounted. Each particle or parcels (groups, clouds) of particles is tracked on an Eulerian fluid flow field.

Yet, is there a difference between including PSD and accounting for all the physics introduced by PSD? To understand this difference, let us first consider the case where inter-particle forces due to van der Waals, electrostatic, and liquid bridge interactions are unimportant. Here, the particles make enduring contacts only in dense assemblies, and they interact with each other only via (mostly) binary collisions in dilute systems. In MFMs, the momentum exchange between particles of different types can occur through a combination of random and directed collisions, both becoming progressively more important with increasing particle volume fractions. (The TFM can be thought of as a limiting case where the rate of momentum exchange between particles of different types is so large that all particles have the same local-average velocity.) Although models for this momentum exchange have been formulated and used in MFM simulations [6, 7], more scrutiny of the quantitative accuracy of such models through particle dynamic simulations would be of interest. Particle-particle collisions are handled indirectly in MP-PIC simulations, and share the same uncertainty associated with the MFMs. In contrast, CFD-DEM simulations resolve all particle-particle contact interactions and can handle PSD; however, as tracking the motion of every primary particle in an industrial process device is impractical and

prohibitively expensive, it is common to simulate the motion of *parcels* consisting of many primary particles of similar attributes (i.e., particle size, density). The rates of momentum exchange between the parcels and the momentum diffusion are only an approximation of the actual rates. Thus, every modeling approach involves approximations of momentum exchange through particle contact that would benefit from further assessment.

Approximations enter the treatment of gas-particle drag force as well. How does the gas affect the particles and vice versa. First, consider the case of uniformly sized particles. The drag force models that we use in TFM simulations apply to homogeneous assemblies (where the particles are distributed randomly), and the force on a particle predicted by these models is an average over many particles. In many simulations seeking to obtain grid-independent solutions (in a statistical sense), these drag models are applied even with fine grids of the order of only a few particle diameters. It seems reasonable to expect that the actual drag force in small systems could differ from the average drag force, requiring, for example, a stochastic correction that scales inversely with grid size. We do not know for sure if such modifications are important.

Next, consider systems with PSD. The drag force models for such systems are usually *ad hoc* adaptations of the models for assemblies of uniformly sized particles (e.g., use the same Wen-Yu [8] drag law for every particle type using its diameter in the Wen-Yu expression). Improved drag models are being developed for polydisperse mixtures [9]; more studies on the sensitivity of simulation predictions to changes in the drag force models are warranted. The adequacy of these models when the full PSD is employed in simulations is even less clear. Thus, going forward, quantifying the uncertainty in the model predictions stemming from uncertainties in the gas-particle and particle-particle interaction models should become a required metric. (Papers dealing with uncertainty quantification (UQ) are beginning to appear in the literature (e.g., see [10]). It would be good to develop guidelines for affordable, minimum required UQ analysis that probes model sensitivity.)

Let us next examine systems with inter-particle forces. In fluidized beds of Geldart Group A particles, the beneficial effect of fines [11] stems from inter-particle cohesion [12]. Hence, models that account for PSD without capturing the consequences of inter-particle forces may not be much better than models that assume a single average particle size.

Inter-particle forces affect the microstructure of particle assemblies. It decreases the particle volume fraction at minimum fluidization conditions [13], and it alters the rheological behavior of the granular phase [14]. While assemblies of particles without inter-particle forces can only sustain compressive stresses, cohesive assemblies can support tension, and they manifest yield stress behavior. At any specified particle volume fraction, the microstructure of an assembly would vary with the strength of the inter-particle forces. As the drag coefficient for fluid flow through particle assemblies depends on the microstructure, should it not depend on the strength of the inter-particle forces?

TFM simulations are commonly performed using the same drag laws for Geldart groups A and B particles. Many drag laws, such as the widely used Wen-Yu model, were formulated for systems in which each particle moves freely [15]. In the presence of inter-particle cohesion, a fine particle may adhere to a larger particle or other fine particles. For the drag, these clusters of particles do not behave as several particles but as one larger “particle.” Hence, one should question the basis for using the Wen-Yu drag law for systems where the particles could move as clusters. The size of the cluster can be expected to depend on the local hydrodynamic conditions and the interparticle interaction force. Furthermore, TFM simulations of fluidized beds of Geldart group A particles often use the kinetic theory of granular materials developed for non-cohesive systems, while turning on an *ad hoc* frictional stress model for particle volume fractions greater than that at minimum fluidization conditions.

When simulations based on such constitutive models for the drag force and particle phase stress yield predictions close to experimental data, one wonders if it is serendipitous, and when the predictions are bad, it is natural to exclaim, “Of course! What would you expect with a deficient model?” Constitutive models for cohesive systems when there is PSD are even more limited. There is a need to develop constitutive models for the drag force and particle phase stress that account for inter-particle forces.

One approach to formulating these constitutive models would exploit detailed simulations. Particle-resolved direct numerical simulations (PR-DNS) of fluid-particle flows [16] can be used to develop constitutive models for fluid-particle and particle-particle drag forces (to be used in MFMs, MP-PIC, and CFD-DEM simulations) that include parameters related to inter-particle forces (such as the Bond number). Realistically, a good first step would be to consider mono- and bi-disperse assemblies of particles before considering the full PSD. Simulation studies focusing on monodisperse

assemblies would be testbeds for code development and testing; one would need at least a bi-disperse system to capture the effects of varying the fines content. Such simulations have not yet been done extensively as they are expensive; furthermore, the types of small-scale flow simulations that would yield results that apply to cohesive assembly flows and span the entire particle volume fraction range are not obvious (at least to this author). Such efforts are ripe for innovative solutions!

A second approach would forego the effort to develop general-purpose constitutive models. Instead, one would begin with measurement of the PSD of the particles of interest and a standard set of experimental tests and simulations to calibrate the inter-particle force model parameters, perform a small number of PR-DNS simulations using these parameters and a user-friendly open-source code (to be developed and made available broadly), and exploit machine learning (ML) algorithms to develop drag force models that apply to the particles of interest. Similarly, PR-DNS or CFD-DEM simulations of fluid-particle flows could be used to formulate explicit [17] or ML models for particle phase stress, allowing for PSD and inter-particle forces. These models can then be included in MFM and MP-PIC simulations.

In some instances, it may not be possible (or is prohibitively expensive) to directly measure the parameters associated with the inter-particle forces. In addition, surface asperities may make it very difficult to directly identify average parameter values [18]. One can use proxy experiments to calibrate the force model in such cases. For example, fluidization-defluidization experiments have long served as important data to calibrate drag models. Other rheological and flow tests have also been used to calibrate and validate DEM model parameters using the data from the proxy experiments (for example, see [19, 20]). It would be beneficial to standardize the set of experimental tests to calibrate and validate the inter-particle force models, which can then be used in the above-mentioned microscale simulations to develop "on-demand" constitutive models.

The software tools used to perform the calibration must also become readily available for this path to be realized. The downside of such an approach would be that the underlying interaction force model used to tune the interaction parameters may not even be the actual mode of interaction. For example, cohesion between particles can arise due to Johnson-Kendall-Roberts (JKR) type adhesive force, van der Waals (vdW) force captured via a Hamaker constant, or the force transmitted through liquid bridges. All three types of interaction lead to some similar effects (e.g., all of them manifest yield stress) and some differences. Thus, one could find an apparent Hamaker constant by tuning the model to experimental data gathered for a system where the interaction is largely due to liquid bridges. Even if such a fictitious model succeeds in the validation tests, it does not afford confidence in the macroscale flow behavior predicted by the simulations. Thus, it would be good to understand the origin of the interaction, select appropriate force models and tune its parameters.

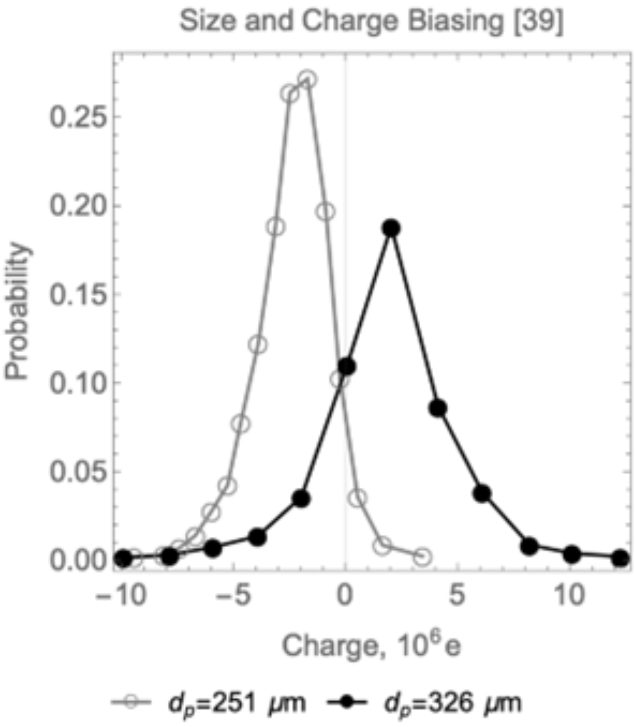
It is worth noting that adding fines can act in opposite ways. In the common case of Geldart Group A particles, the larger particles (> 75 microns) have minimal cohesive interactions among themselves, and the addition of fines (< 45 microns) introduces cohesion [21]. In contrast, in the case of toner (polymer) particles (~ 10 microns), the primary particles are cohesive, and when they are coated with silica nanoparticles, the severity of cohesion decreases [22].

Models for cohesion – namely the JKR model and the model involving the Hamaker constant (see [23] for a discussion of the applicability of these models) – are used widely. Despite complications introduced by surface asperity, effective parameter values (namely, the adhesive energy or the Hamaker constant), which can be thought of as surface averages, can be extracted from proxy experiments. Models for forces due to pendular liquid bridges and the bridge's filling rate and rupture criterion, which involve the liquid-gas interfacial tension, liquid viscosity, and contact angle as parameters, have been developed and used in the literature extensively [24, 25]; these studies typically assume smooth particle surfaces. As asperities on particle surfaces are not resolved in practical simulations, one will likely have to find effective values for one or more parameters (e.g., the apparent contact angle and an effective time constant for bridge filling) through proxy experiments. Demonstrating that such model tuning can be achieved through well-defined proxy experiments and standardizing those experiments would be worthwhile goals of future studies.

Electrostatic charges carried by the particles can lead to lateral segregation of particles in risers [26], affect bubble characteristics in fluidized beds [27], make particles stick to each other and to other surfaces [28], lead to problems such as sheeting in polymerization reactors [29], and create unsafe operating conditions. Unlike the cohesive forces mentioned in the previous paragraph, the electrostatic interaction is long-range in nature, which adds to the computational complexity. Much remains to be learned about the nature of the species involved in tribocharging of

dielectric material [30,31], the effect of PSD on tribocharging (where particles of the same material but different sizes can acquire different charges [32], the role of adsorbed water (which is affected by the prevailing humidity and temperature) on charging [32, 33], the effect of polarization on flow behavior [34] and particle elutriation [35], and the likelihood of micro-discharges in the gas affecting the extent of charging [36, 37].

Although progress has been made in device-scale simulations of the effects of charges on particulate flows (e.g., see [38]), much remains to be understood about the flows of particles with bipolar charges; For example, consider a bi-disperse mixture (such as 90-micron FCC particles with fines). Further, suppose that the large and small particles carry opposite charges, even though the overall mixture may be charge-neutral [39]. The large and small particles can stick together, affecting how they pack (which affects the particle volume fraction at minimum fluidization) and how clustered they remain under flow conditions (which affects both effective gas-particle drag and particle phase stress). A deeper understanding of the importance of (or lack thereof) these effects on device-scale performance is very much needed.



In a bi-disperse mixture where the large and small particles carry opposite charges, the interaction between unlike particles is attractive, while that between like particles is repulsive. In contrast, if the inter-particle force is due to vdW or liquid bridge forces, all interactions are attractive. Thus, it seems reasonable to expect that these systems would manifest telling differences in some flow (or elutriation) behavior; it is unclear at present what it might be and how it can be used to identify readily the nature of the interaction at play in a gas-particle system of interest.

Electrostatic interaction differs from the vdW and liquid bridge forces in another important way. All these interactions can lead to particle clustering, which can alter the effective gas-particle interaction force upon fluidization. (Hydrodynamic clusters form even in particles without any of these interactions but they are weaker in nature.) The dynamics of these clusters play a role in the development of macro-scale inhomogeneities. Absent in the case of vdW and liquid bridge forces, but important in the case of electrostatic charges is the space charge effect arising from the electric field due to all the particles in the flow device. A large electric field can develop a device containing many charged particles and contribute to macro-scale flow structures. Interestingly, dielectric breakdown of the gas can occur, limiting the extent of charging; for example, particles in a laboratory fluidized bed are known to acquire lower charges in argon than in nitrogen, which can be explained by the difference in the dielectric strengths of the gases [36, 37]. It brings to light two observations: (a) The larger the bed, the greater the likelihood that the net charges acquired by particles are dictated by the dielectric strength of the gas (whatever the gas may be). (If this conjecture of this author is correct, micro-discharges are constantly occurring in every large-scale fluidized bed installation!). (b) This dielectric breakdown-limited charge level will depend on both temperature and pressure; for example, as the dielectric strength of a gas increases with pressure, the particles can sustain more charges at higher pressures [38], and so the effect of electrostatic charges on hydrodynamics can be more pronounced at higher pressures than at ambient pressures. Much remains to be understood about the importance of such effects and the means of capturing them in the models.

Affordable simulations

In his article in the Spring 2021 PTF Newsletter, Madhav Syamlal discussed emerging computing technologies and their potential impact on multiphase CFD [40]. These advanced capabilities will dramatically increase the scale of the simulations, such as the number of cells used to discretize the flow domain or the number of particles or parcels employed in the simulations. While these advances will enable deeper inquiries into underlying physical and chemical processes, industrial application of computational (reacting) multiphase flows will likely rely on more affordable and more accessible computers that are a generation or two behind state-of-the-art machines. Furthermore, as industrial

design analysis calls for numerous simulations (with UQ analysis adding to the computational load), coarser simulations that are faster to run (while being reasonably accurate) will continue to be of value for the foreseeable future.

A great deal of work has gone into the development of filtered TFMs that are suitable for coarse simulations of gas-particle flows and the associated constitutive models [41-43]. The bulk of this development has focused on the fluidization of non-interacting particles. Unsurprisingly, a recent study [44] shows that the filtered constitutive models would change when cohesion is included. Based on our experience analyzing non-interacting particles, it seems more productive to develop a readily available toolkit needed to generate on-demand filtered constitutive models for particles with PSD and interparticle forces. Parcel-based simulations of cohesive systems would require such corrections as well.

A productive pathway would be to (i) perform the proxy experiments mentioned above and tune the parameters associated with the interparticle forces, (ii) carry out a small number of PR-DNS and CFD-DEM simulations to tune ML-based filtered constitutive models, and then (iii) undertake the device-scale simulations. Individual pieces in this toolkit are available in various research groups and open-source or commercial software platforms. Collaborative efforts among research groups to assemble a report on *best practices approach for these steps,* and a suite of simulation and post-processing codes to be used in various steps would be valuable.

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In Memoriam - Dr. Jerry R. Johanson

LEGACY: FOUNDATION OF BULK SOLIDS FLOW THEORY

Kerry Johanson, Material Flow Solutions

Introduction

Paraphrasing Socrates, "There is nothing new in this world. We can only discover what has been there all the time." Jerry Johanson's experience in discovering the theory of bulk solids flow taught him that innovation or discovery occurs when we accumulate enough knowledge to receive inspiration and then put it all together to innovate or discover what was already there. His part in developing bulk solids flow theory and application is an example of this innovation concept.

In the beginning...

Dr. Andrew W Jenike, a recent 1950s immigrant to Utah from Canada and England where he received his PhD concerning grain silo design, started working at American Smelting and Refining Company. His first assignment, to design a bin for ore, showed him the dearth of knowledge concerning bin and hopper design. The current literature considered only bin wall pressures caused by the ore. These pressures are necessary to determine the wall thickness and structural support. Nothing in the literature suggested the size of an opening or what kind of feeder was necessary to allow the material to flow from the outlet. At that time, Jenike was inspired to devote his career finding a way of describing flow in hoppers and bins. He quit his job and started teaching and researching at the University of Utah. Since the stored materials used by the smelting and refining company all looked like dirt, soil mechanics was obviously related to bulk solids flow. However, soil mechanics was concerned with soil stability while bulk solids flow was concerned with assuring the solids were unstable. Jenike's next inspiration came with the help of R. T. Shield [1], when he developed the effective yield concept. This concept combines the soil mechanics yield locus with a new continuous soil flow concept in which the soil strength (a function of consolidation pressure) varied continually as the soil flowed through the hopper at continually changing consolidation pressures. For calculation convenience, Jenike assumed that this effective yield locus started at the 0.0 position of the shear stress and normal stress coordinates.

At the same time, a Russian scientist named Sokolowski [2] developed the theory of soil mechanics using the general partial differential equations of equilibrium and the yield locus concept. With painstaking hand calculations, Sokolowski developed some soil mechanic stress fields. Without a high-speed computer, the job was endless and so he made some simplifying assumptions and reduced the system of equations to ordinary differential equations solvable without numeric calculations. One of these assumptions, stress increased linearly upward from an apex point (Radial Stress), coincided with Jenike's pressure measurements in a converging hopper[4]. Bulk solids pressures decrease linearly to zero at the hopper apex, precisely the condition that Sokolowski assumed to simplify the equations. Just prior to this time, Jerry Johanson had the good fortune of joining Jenike in his research efforts at the University of Utah where he was an undergraduate student. He has been asked many times, "Did you plan this?" His response, "No, it was just dumb luck." As a newly married third year engineering student, Jerry needed a job and he got it.

Art Nettleship, one of his fellow bus riders to the University during his courting years, and instructor of the Machinery Lab of the Mechanical Engineering Department, approached Jerry after his summer marriage and school started. "You need a job with both of you going to school, and I need a lab assistant. It doesn't pay much, but it'll at least pay your rent," he said with his British accent still in place after several years in Utah. Jerry readily accepted. The Mechanical Lab building was just across the alley from an old coal bin that would soon become the Bulk Solids Flow Laboratory of the University of Utah Engineering Experiment Station. When school ended for the summer, Jerry's job ended. Art introduced Jerry Johanson to Dr. Andrew W. Jenike head of the newly established Bulk Solids Flow Project. Fortunately for Jerry, the project had just received enough funding from the University to hire a student. This was his "dumb luck."

Jerry started working at the Bulk Solids Lab the summer before his senior year. Jenike wasted no time in telling him that the work would require him to get a PhD. So he went home and told his sweetheart, "I am thinking of going to

graduate school for a PhD degree." She had only one question, "How much longer will it take?" With tongue in cheek, Jerry said, "three years." She replied, "All right – three more years but no more, and I want to start our family before then." It turned out that she was already pregnant.

Life in a coal bin

The University had converted their steam generating plant from coal to natural gas leaving a large coal bin space useless and unwanted. This was a perfect place for a Bulk Solids Flow laboratory since they could offload coal using the existing chute into a storage bin, instrumented belt feeder, and the runaround conveyor system Jenike had designed and built. Jerry's hangout, behind the window next to the small door just behind the sign in the photograph, is just large enough for a small desk and a drafting machine. Beyond the office was the shear-testing lab and coal handling system. The office was reasonably clean except when the coal was recirculated through the belt feeder and bin. Without a dust collecting system, black dust coated everything.

Figure 1. Humble beginnings as a grad student

Shear testing

When Jerry joined Jenike in June of 1958, Jenike had already developed a low-pressure shear tester. Jerry's first job was to develop shear testing techniques to produce reliable and consistent results. With soil mechanics standard shear testing techniques, this consistency is controlled by compacting a measured weight of soil into a test cell of known volume, thus controlling the sample's bulk density. This technique reproduces the packing effect of an earth moving tractor and compactor. Unfortunately, soil contained in a bin enters the bin loose and is compacted only by the bulk solids contact pressure created by the weight of the soil as it enters and moves through the bin. Something different was needed and the concept of a developing a critical state of stress during compaction was born.

Jenike had already departed from the standard square shear cell in favor of an easier to manufacture circular test cell that eliminated poorly filled square corners. Unfortunately, the shear cell has limited travel. Jenike proposed twisting the top under load to induce shear and thereby distribute the test specimen and pre-consolidate the specimen before applying the measured shear force. He wondered how they could determine if the twisting was active throughout the specimen. Jerry was inspired to punch a vertical hole in the initially lightly compacted specimen with a straitened paper clip and to fill the tiny hole with talcum powder before twisting. Dissecting the specimen after an initial twist showed that the initially vertical white line of talcum powder was deformed, essentially uniformly, from top to bottom in the direction of the twist. This also showed the maximum effective twisting angle. The resulting additional knowledge established the twisting techniques necessary to bring the sample under full compaction during the deforming conditions existing in the flowing bulk solid in a converging hopper. These twisting techniques are still used today to precondition the sample in the Jenike method of testing.

Calculating stresses in bulk solids flowing in a converging hopper

The university had purchased a Borrows 205 digital computer with flashing lights in octal code, paper tape program and data input, and only machine language to program it. There was no Computer Science department or even a class offered in computer programing. After much self-study, Jerry successfully wrote a machine language program to calculate the stresses in flowing granular materials (bulk solids). Without a digital plotter, every graph was hand plotted. About 30 hours of computing time was needed to do the calculations. That same task could be accomplished in less than a minute today. Because Jerry always had a cautious approach to computer calculations, he always hand calculated at least one known limiting case of the more complicated computerized calculations.

Predicting arches in hoppers

While Jenike was away for the summer visiting his family in Poland, Jerry calculated and plotted the radial stress using computer code. Up until then, Jenike had analyzed the shear test results by the slope of the unconfined yield strength

versus the major principal consolidation pressure line (flow function). Larger flow function slopes indicated a less free-flowing bulk solid. That was the wrong approach. Combining the computed results for pressures in a hopper, the measured strength f_c as a function of compaction pressure σ_1, the calculated stress σ_{1a} at the abutments of a self-supporting arch, and the simple fact that the arch will fail when $\sigma_{1a} > f_c$ (flow no flow criteria) where f_c is determined experimentally at the calculated consolidating pressure, the simplified arch analysis shows:

$$\sigma_{1a} = gB/(1+m)$$

Where g is the bulk density, B is the conical outlet diameter or width of a one-dimensional converging hopper, and m equals one for a conical hopper and zero for a one-dimensional converging hopper.

Since the radial stress calculations produced the steady state consolidation pressure σ_1 of a flowing bulk solid in a converging hopper, and $f_c(\sigma_1)$ is determined experimentally, Jerry could calculate the outlet dimension B necessary to prevent arching during steady flow in a converging hopper. When Jenike returned from Poland, Jerry told him of his discovery. Jenike disappeared for two days. When He returned, he presented his flow factor techniques for calculating bulk solids arches in cones and wedges under steady state flow conditions. This applied to conditions of radial stress only and was consistent in practice with Jerry's work. But, Jerry Johanson saw a further reaching approach. His more generalized approach of first calculating the compaction pressure and then measuring the associated unconfined yield became the basis of arch calculations in all situations including arches from initial compaction from falling solids in a bin, compaction from mobile equipment working a pile, time effects on the bulk solids arching, and vertical pipe (rathole) failure.

Observing flow in hoppers

Inspiration hit Jerry Johanson again when his computer stress calculation in a conical channel showed that there was a limited cone angle for which mathematical solutions exists. One of the assumptions of the radial stress theory was that the consolidation stress circle was tangent to the effective yield locus and that the effective yield locus describes the state of stress subject to the condition of continual deformation without volume change (i.e. flow). This suggests that steady flow along the hopper walls will probably not occur for hoppers less steep than this limit. This theoretical limitation could only be substantiated in a physical conical hopper full of a bulk solid. Others had attempted to observe flow in conical hoppers by using a half cone with a glass front. These observations were woefully inconclusive. The flow in a cone was 3-dimensional. So, Jerry developed a series of three-dimensional conical hoppers with reinforcement rings to cut in half after laying them on their side. While in the upright position, the hopper cones were loaded with horizontal marker bands. A limited amount of bulk solid was allowed to flow, the solids void at the hopper top was replaced. After capping the top, the entire assembly was laid horizontally. The true flow pattern was exposed when the top half of the hopper was removed and the markers between the hopper halves revealed.

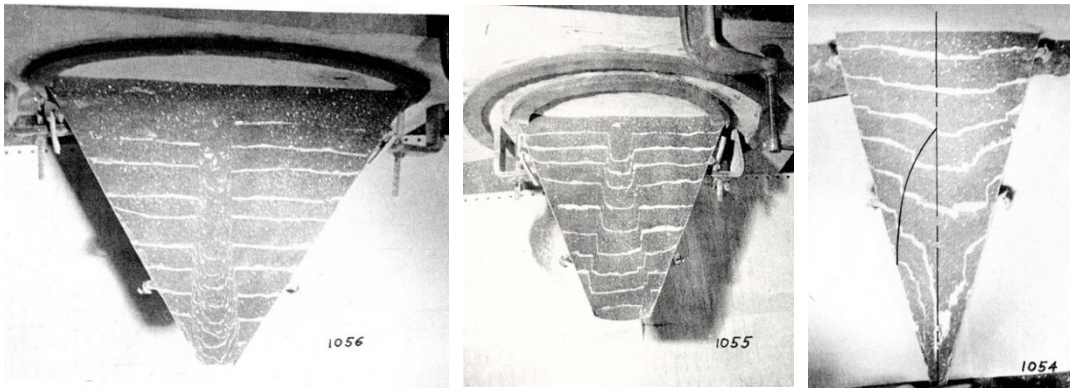

Figure 2. Early experiments to confirm radial stress in hopper flow

This unique technique confirmed the calculated hopper and friction angle limits for conical channel flow. These results became the basis for hopper angle design to achieve flow at conical hopper walls. The concept of mass flow was born. The observed breaks in the original lines in the steep hopper must occur along velocity characteristics of the stress solutions. The lines shown in the photo labeled 1054 are the calculated velocities characteristic of this hopper angle θ' and wall friction angle φ' (Figurer 2). This confirms that the calculated radial stress field corresponds with the actual stress field in the physical model. General calculations in Jerry's thesis showed this radial stress convergence [3] [4].

Jenike developed a mathematical convergence later.

Stable Rathole Calculations

Flow pattern observers in the 1950s always concluded the flow of cohesionless bulk solids was only directly above the hopper outlet and then expanded at the top to the walls. With cohesive solids, much of the material outside of the flow region stayed in the bin and hopper. With very cohesive solids a stable circular pipe formed above the outlet. Jenike noted that Sokolovskii mathematically described the stability of such a structure in earthy media. Using Sokolovskii's derivation, Jenike observed that if one assumes the media has a constant cohesion (unconfined yield stress), the stresses in the pipe are independent of the depth of the pipe, and the pipe is at failure condition, then there is a mathematical relation between the pipe diameter D, the internal friction angle j, the unconfined yield stress f_c and the bulk density g. For this to be valid in bins and hoppers, it is necessary to show all boundary conditions at the top of the bin converged to this unique independent of depth solution. Since Jerry had just finished showing by generalized characteristics calculation the convergence to radial stress, he suggested they do the same for piping starting at the top diameter with a vertical circular horizontal stress free surface with enough top vertical stress to force the solids in the failure ($\sigma_v = f_c$). The calculated characteristics intersecting the extended stress-free wall then determined the wall slope at any point below in the initial vertical slope. This procedure governed the shape of the stress free boundary at failure. Jerry set up the flow chart for Bing C Yen's computer calculation and helped him with the programming. In every case the stable slope diameter converged to that unique value found in Jenike's simplified case. Unfortunately, when Jenike applied this to real life pipes he used the Janssen pressure inside a vertical channel as the consolidation pressure and the strength of the pipe wall instead of the generalized approach using the initial compaction pressure imposed when the bin was first filled.

At this point, Jerry graduated from the University of Utah and began to practice what he had learned as part of his graduate research.

US Steel Research Center

Established in 1956, the US Steel Research Center in Monroeville, PA was noted for its green lawns. When the US Steel research vice president was asked how the lawn got so green, he simply said, "We plant money." (Figure 3)

The US Steel Research Campus in Monroeville PA

Figure 3. US steel plant – home to Jerry Johanson after University of Utah

Jerry knew nothing about making steel, so why did a steel making company want a PhD specializing in solids flow in their Applied Research Laboratory? Because US Steel handled over a hundred million tons of raw materials (coal, iron ore, coke, and limestone) each year and if they saved one penny a ton on the handling cost, they could add one million dollars to their profit margin. Jerry's job was to help them accomplish that.

Million Dollar Rathole

Jerry's first assignment from his section leader Henk Colijn, was to write a movie script and three hour course explaining the bulk solids theory and reduce the highly theoretical work presented in Jenike's "Gravity flow of bulk solids, University of Utah Engineering Experiment Station Bulletin 108" and his own PhD thesis "Stress and Velocity

Fields In Bulk Solids Flow" [3,4] to practical application. The assignment resulted in the movie "Million Dollar Rathole" (US Steel, 1963) and the published paper "New Design Criteria for Hoppers and Bins" [5]. This paper preceded the publication of Jenike's Bulletin 123 .

Jerry had models made, in-plant video shots taken of hoppers and feeders in action, background music, and a professional edit or and actor to narrate the script. He said, " It is nice to have large company resources working with you."

One of the real-life flow problems modeled in their movie was from Clairton Works's multiple outlet coal bunker feeding a lorry-car above a coke battery, showed the bunker capacity reduced to less than 10% of the design capacity. This was attributable to the 2-ft. diameter vertical pipes (ratholes) above each outlet while the rest of the material was stagnant. When vibrators and air lances collapsed the vertical pipes, compacting coal directly above the outlet, the coal arched and stopped flowing.

The movie also showed the 95% capacity increase possible using a series of steep mass flow conical hoppers inserted into the existing pyramid shaped hoppers. When the superintendent of the plant saw the movie, he agreed to make the modification. The first multiple outlet mass flow application worked.

Briquetting iron ore

The open-hearth steel making process requires consistently sized, hard lumps of iron ore to penetrate the molten surface of the steel batch. The only source of such ore for US Steel was Brazil. The Brazilian companies realized this single source and significantly increased their price for the hard-lump ore. The order came: US Steel Research must develop an alternative to the Brazilian ore using hot briquetting. The project was unsuccessfully well underway when Jerry arrived at the Research Lab.

Jerry suggested that the principles of bulk solids flow applied just as well to roll type briquetting presses as to hoppers and bins. In a few weeks he put together his "Roll Pressing Theory". This showed that there were 18 variables affecting the briquette quality. Up to that point, the US Steel research team experimentally determined the factors that make a good briquette. Finding the interaction of 18 variables experimentally was an impossible task, especially when you don't even know most of those variable. With the new roll press theory [6], a few hot ore strength and compressibility tests successfully predicted the press size and operating conditions that would produce the required quality briquette.

To simplify the roll press calculations, Jerry reduced the soil mechanics hyperbolic type partial differential equations to ordinary differential equations and approximated the characteristic lines by circular lines satisfying the boundary conditions, symmetry at the center line, and friction requirements at the wall surfaces. This unique discovery produced a pseudo two-dimensional problem with a variable horizontal cross-section and a one-dimensional variable compacting stress. This allowed for the calculation of the roll press nip angle as well as the angle from the press horizontal centerline where the solid stops flowing at the walls and is compressed solely by the changing distance between the rolls. The nip angle occurs when the bulk solid density change rate associated with the stress gradient as the soil slips at the roll surface equals the density change rate from the changing distance between the rolls. When Jerry presented "A Rolling Theory for Granular Solids" at the ASME Applied Mechanics Conference at UCLA, Prager (the authority on plasticity theory) questioned the method as unproven. Jerry replied that he had experimental verification with full scale iron ore briquetting applications on an actual roll press located at the old Universal Atlas Cement plant owned by US Steel at Universal, Pennsylvania. The basic rolling theory, reduced to a series of practical use graphs, presented at the International Briquetting Association in 1965, has become the basis of roll press selection and operation. Later, Jerry included entrained air and predicted limiting roll speeds based on measurable solids flow properties. Both discoveries, the circular characteristic and the pseudo two-dimensional assumptions, became valuable in his later work as he added gas flow, liquid flow, chemical reactions, and viscosity to the solids flow theory.

Inserts in bins

The next circular characteristic use was the "Insert placement in bins" [7]. A scientific criteria for insert placement was also experimentally verified with the marker placement technique Jerry used for his thesis work. The paper "Use of inserts to control flow patterns in bins" earned Jerry the "Hennery Hess Award" for the best paper by an associate member of ASME.

"Loads on bin inserts" showed that introducing convergence in a vertical channel produces an arching effect between the insert and the hopper walls producing a vertical insert load much higher than the weight of solids directly above the insert. This experimental observation later helped explain why a protruding brick in a blast furnace vertical wall caused a high pressure worn spot in a brick opposite the protruding brick on the otherwise vertical wall. This is a good example of further information causing further discovery.

Predicting flow rates from hoppers and bins

"Predicting flow rates from hoppers and bins" [8] came about when Jerry recognized that the concept of solids continuity required solids to accelerate in the converging channel, thereby causing an acceleration force opposing the flow. He also recognized that because the vertical stress becomes zero at the plane of separation as the bulk solids free fall from the hopper outlet, the stress on the side of the hopper can still be non-zero because the bulk solids have cohesion. This observation led to a flow rate equation containing the effect of the bulk solid's cohesion.

Forming Jenike and Johanson Inc

After four years at US Steel Research, publishing nine papers and numerous internal US Steel reports, Jenike asked Jerry to join him at his new home in Winchester Mass. US Steel was pushing Jerry toward management responsibilities and the dreaded politics associated with that job. His research desire dictated that he move his family to the Boston area and join Dr. Jenike in July 1966, at his new home in Massachusetts.

Jenike was deeply involved in fine powders, ground anthracite and bentonite. They needed answers to flooding and flushing, a phenomenon Jerry had never encountered at steel plants. Jerry included flowing entrained air in the theory. This theory expansion included the Janssen concept [9], made a constant ratio of average vertical stress and horizontal stress, and the discovery (when calculating radial stress situations) that this ratio is constant even in converging channels. This, combined with air permeability and compressibility measurements on bulk solids, provided sufficient additional knowledge to calculate the transient settlement of bulk solids in vertical channels [10] and the steady flow of powders in converging channels. The resulting calculations verified experimentally and in full scale applications, gave further confidence to their recommendations concerning fine powders. This knowledge expansion, simple solids and air continuity calculations, provided the pressure and air injection rates needed to stipulate the required fine powder flow rate without causing flushing [11, 12]. Again, the accumulation of knowledge allowed the inspiration necessary to expand the bulk solids flow theory.

Just about a year after Bridgewater published a paper on segregation which focused mostly on sifting segregation, Jerry Johanson published a paper suggesting six important segregation mechanisms and what to do to minimize these effects [13].

Jerry had a dream…

Steep hopper walls solve most solids flow bin hang-up problems. Unfortunately, this decreases the bin capacity and requires much more headroom than exists in many plants. One of Jerry's clients could not afford the decreased bin capacity nor the required headroom. So, Jerry slept on it. In his dream he remembered calculating stress fields involving the region between an interior conical structure and an outside conical hopper. The allowable angle to produce flow at the walls between the inside cone and the outside cone was the same as that required for flow at the walls on the inside cone. Thus, the required total included angle for the outer cone was twice the angle required for the inside cone. A simple conical insert could make an existing hopper flow at the walls and eliminate hang-ups. Jerry awoke immediately and wrote up the concept for his novel cone-in-cone invention. After proving the concept with a model hopper, he soon realized that he could control the flow pattern in the bin by adjusting the flow pattern under the conical insert and thus create an in-bin blender or a perfectly uniform velocity throughout the bin – a condition required for a chemical reactor or in-bin dryer or to produce a velocity profile that could make a differential velocity of three-to-one so as to blend a continuous stream of product over three times the bin volume. Thus, the in-bin blender was born (Figure 4). These discoveries came after knowledge accumulation was sufficient for inspiration to put it all together.

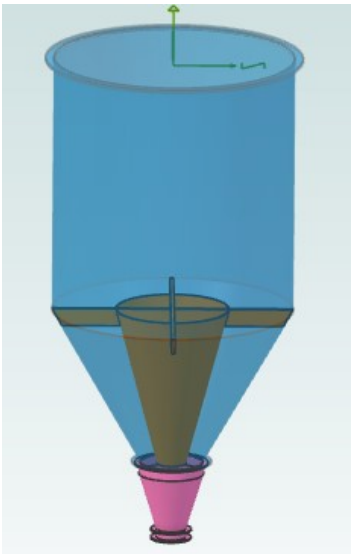

Figure 4. Cone-in-cone

The Diamondback® Design

It was discovered early on that arching was far less severe in a plane flow bin than in a conical bin, but the use of a plane flow bin required the use of a belt feeder or screw to promote mass flow over the entire length of the hopper. Jerry considered that limitation and the desire to make a plane flow hopper with a circular outlet. Over the course of a month, he developed a solution to that perceived limitation. Using an old table saw, he cut a pipe in half and then connected it to a couple of triangular flat plates to create a hopper that reduced from an oval opening to a round opening. Plane flow behavior was observed in this unique geometry. Jerry already knew that a round-to-oval hopper would work as a mass flow device with a reduced arching capacity, But, the combination of round-to-oval and oval-to-round allowed plane flow hoppers to have a circular outlet – and the Diamondback° bin was formed [14] - as shown in Figure 5.

Figure 5. Diamondback™ Design

Amongst his many contributions, Jerry would be remembered for his contributions in ..

- Helping to develop the first test device to measure the cohesive flow properties of granular materials for use in silos
- Helping to develop and implement the concept of radial stress theory
- Helping to develop the concept of predicting arching in hoppers after steady flow
- Extending the prediction of arching to conditions not compatible with radial stress
- Calculating the radial velocity patterns in mass flow devices and discovered that conical hoppers have a natural limit where flow along the walls can exist
- Discovering that plane flow hoppers have no such theoretical limit, but have a practical mass flow limit
- Developing roll press calculations for the case of no gas flow and for the case of gas flow
- Developing limiting mass flow rate equations for fine powders
- Developinglimiting flow mass flow equations for viscous-like materials (e.g. tar/oil sands)
- Identifying multiple segregation mechanisms present in bulk material and developed solutions to solve those segregation problems
- Developing the cone-in-cone as a means of controlling flow pattern in hoppers and reducing headroom requirements to obtain mass flow, thereby, allowing either blending or anti-segregation flow devices in hoppers and bins
- Developing a plane flow geometry that allows plane flow behavior down to a circular hopper outlet

He was a visionary who coupled knowledge with inspiration to solve complex problems in bulk solid handling. Several generations of practitioners, engineers, and students have benefitted from his contributions to the field to date, and will continue to do so in the years to come.

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Particle Playground : Understanding Particle Size Distributions

Ray Cocco, Particles in Motion, LLC

Arguably, the particle size distribution (PSD) is typically the first and last data you look at on your particle technology project. It may also be your best metric during your particle technology project. An analysis is easy, robust, and, if done right, accurate. However, we often reduce the PSD to a median or Sauter mean particle size. There is a lot of data in that PSD curve and nuances regarding how it was measured. Here, we will discuss what you can see with a PSD curve, how the interpolation of that curve depends on how it was measured, and the common pitfalls when collecting and analyzing PSD data.

Types of PSDs

PSDs are typically represented as a cumulative or distributive curve in terms of weight percent or fraction, as shown in Figure 1. The cumulative curve is just the integration of the distributive curve and the differential of the cumulative curve yields the distributive curve. Since the cumulative is the integration of the distributive curve, cumulative curves often conceal data due to the integration process. Thus, comparisons of PSDs should be made with distributive curves [1].

Distributive or differential curves are often represented with a Gaussian or Log-Normal distribution, although such curves are idealistic compared to real data. A convoluted array of Gaussian or Log-Normal distributions can be used for more complicated PSDs, as shown in Figure 2. Cumulative curves can be modeled with a Rosin-Rammler distribution which is a Log-Log distribution [2]. This distribution was popular before easily assessable computers, but it is rarely used today.

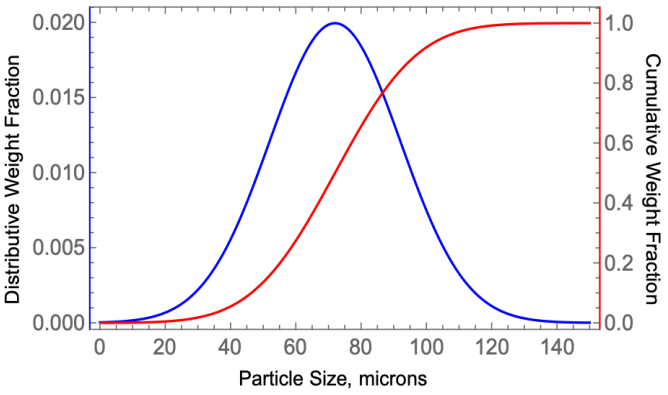


Figure 1: Cumulative versus distributive distribution for the same data.

What is perhaps more important with the type of PSD is the basis of the sample collection, that being by weight, volume, or number. With a broad distribution, weight, and volume fractions can underweight the smaller particle sizes. A classic example is the fines content in a fluidized bed of Geldart Group A particles. A small increase in the level of the fines or fines level can result in significant differences in the fluidization quality and bed density [3]. Fines, defined as particles smaller than 44 microns, is an arbitrary metric stemming from the sieve tray analysis. In general, particles could not be easily sieved below the No. 325 tray, which corresponds to a spacing of 44 microns. Thus, all the material below this tray was considered fines.

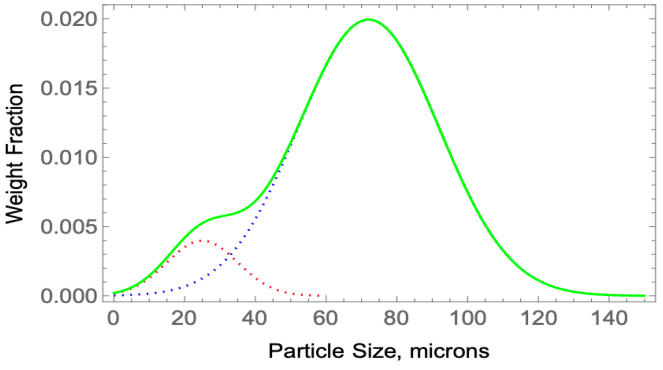


Figure 2: Convoluted curve from two Gaussian curves.

Knowing the fines level or the change in fines level is important in many applications, such as cyclones, pneumatic conveying, and fluidization. The fluidization application presents a good example of this importance. A change in fines levels in a fluidized bed from 2 to 6 wt% fines results in smaller bubbles, smoother fluidization, and a lower bed density (i.e., higher bed expansion).

Figures 3a and 3b shows the differences between these two bed materials. The differences between the two distributive weight fraction curves (blue) are subtle. It seems unlikely that such a small change can have a big impact. However, a bed with 2% by weight or 2 wt% fines corresponds to a bed of 54% by number of particles or 54 num% fines. When the fines level increased to 6 wt%, the distributive number fraction curve showed a marked increase in the smaller particles with a bed of 78 num% fines.

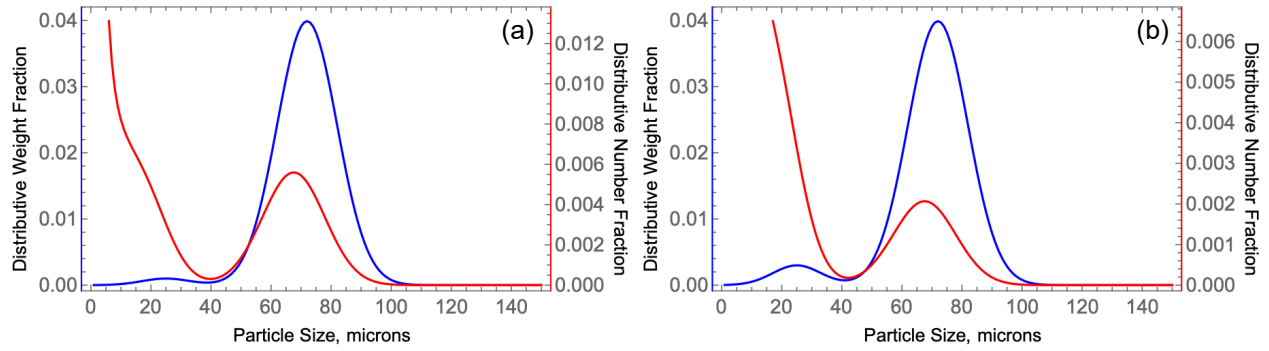


Figure 3: PSDs in terms of weight and number fractions for a Geldart Group A material with (a) 2 wt% and (b) 6 wt% fines (d_p<44 microns).

What the distributive weight percent masks is first, the number of small particles in this bed is significant. For this example, a bed of 2 wt% particles seems small, but in reality, 54 num% of those particles in the bed are fines. Second, increasing the fines level in the bed from 2 to 6 wt% also seems like a small increase but on a number basis, that level increased to 78 num%. When comparing the PSDs on a number basis, it is much easier to understand why the level of fines in a fluidized bed of Geldart Group A powders controls the bed hydrodynamics.

Fortunately, the conversion from a weight or volume fraction to a number fraction can be determined from the expression

$$n_i = \frac{\frac{x_i}{\frac{4}{3}\pi d_{p,i}^3 \rho_{p,i}}}{\sum_i^N \frac{x_i}{\frac{4}{3}\pi d_{p,i}^3 \rho_{p,i}}} \rightarrow constant \rho_p \rightarrow \frac{\frac{x_i}{\frac{4}{3}\pi d_{p,i}^3 \rho_p}}{\sum_i^N \frac{x_i}{\frac{4}{3}\pi d_{p,i}^3 \rho_p}}$$

where n_i is the number fraction of particle size i, x_i is the weight fraction of particle size i, and $d_{p,i}$ is the particle size. For volume fraction to number fraction, the particle density is not needed although this cancel out of the expression anyway if the particle density is the same for all particle sizes.

Measuring Particle Size Distributions

How the PSD was obtained also needs to be considered. The most common methods for measuring a PSD are sieve analysis, LASER (laser) diffraction, electrical zone (i.e., Wallace Coulter), settling velocities (i.e., terminal velocity), and image analysis. Sieve analysis uses an array of trays of decreasing mesh size. Each sieve tray has this mesh with specifically sized square openings. Material is added to the top, and the assembly of trays is vibrated. Material is separated by the smallest dimension that can pass through the trays as it falls through the tray openings. As noted earlier, sieve trays are limited by the PSD resolution tied to the number of trays (e.g., usually less than six) and discerning particles smaller than 44 microns (i.e., 325 mesh size).

The laser diffraction method is based on the size of the diffraction pattern being inversely proportional to the particle size. Small particles cast a larger circular diffraction pattern than larger particles. Measurements can be done of liquid-solid or gas-solid samples. Some units can run continuously, allowing large samples or in situ analysis to be measured. The key to an accurate PSD from laser diffraction is getting the optimum dispersion of the particles in the fluid phase. Two low of a particle concentration limits statistical significance, and too high of a particle concentration results in an error due to overlapping diffraction patterns. Most units have a nozzle or vibrator to promote particle dispersion in the fluid. However, caution is needed here as such mechanisms can cause attrition of more friable materials during analysis.

Particle sizes can also be determined by their impact on an electrical field. In 1943, Wallace Coulter showed that if particles flow through an orifice under an electric field, the particles perturbed the electric impedance [14]. The particles are suspended in an electrolyte and pushed through a small opening between electrodes called the sensing zone. Originally designed to count particles such as blood cells, it was later discovered that the perturbation level correlates to the particle size. This method is commonly referred to as the electrical zone sensing method. It can be quite accurate even for particles smaller than 10 microns, but the particles need to be suspended in a liquid electrolyte solution. Particle swelling needs to be considered.

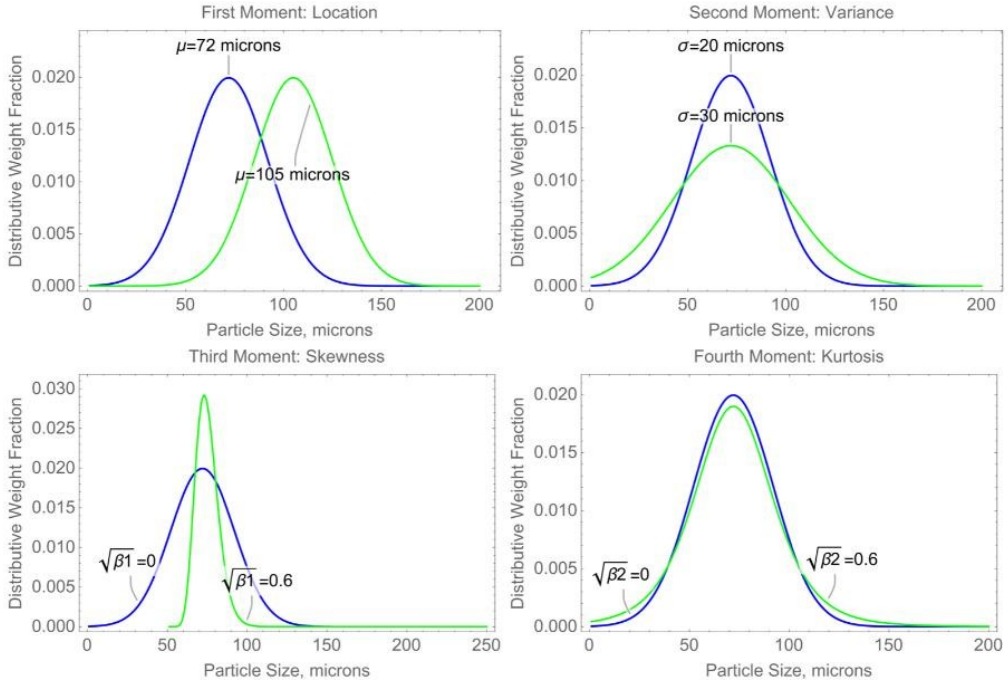
Another method for determining the particle size distribution is using the settling velocity in a dispersed stream. The terminal velocity of a free-falling particle in a fluid is dependent on its size and shape. Once a particle achieves its terminal velocity (i.e., acceleration is zero), the time of flight between two points can be measured using a range of techniques, such as cross-correlation of light scattering waveforms. Unlike the previously noted techniques, shape is a factor here too. Sieve analysis, laser diffraction, and electrical zone sensing all assume the particles are spherical. With the settling velocity, however, particle shape is a factor. Particles with lower sphericity will be recorded with a smaller diameter. This lower diameter is often called the aerodynamics diameter, which can be the more important metric for the design of conveying lines and cyclones.

The newest technique on the market is based on image analysis. A two-dimensional projection of particles is collected. The projected area, projected perimeter, minimum diameter that can encompass the particle, and minimum dimensions crossing the particle can be determined. The technique requires lots of images to achieve statistical significance, and, as with laser diffraction, electric zone sensing, and settling velocity, the particles need to be dispersed. Particles as small as 2 microns have been measured with this technique [14]. This technique can be used in situ in processing conditions as well.

Thus, each technique has its advantages and limits. It would be better to use a PSD based on settling velocity than that from laser diffraction when designing classifiers such as cyclones. If fines level is a key scale-up parameter, it would be best not to rely on PSDs determined from sieve trays.

Moments

In the last two decades, the importance of looking at the whole distribution instead of one "representative" particle size (i.e., Sauter mean, median dp) has come to light [6-9]. However, how do you use an entire PSD through a calculation designed for one representative particle size? This question is especially true when using a two-fluid CFD model for simulating granular fluid flows [10]. One way around this limitation is to reduce the PSD to its statistical moments. Moments can be used to quantify various aspects of a distribution, such as its location (e.g., mean), dispersion (e.g., variance), skewness (e.g., asymmetry), and kurtosis (e.g., base broadening or tailedness). For example, the mean (or



average), variance, skewness, and kurtosis are the first through fourth moments that can be used to characterize a dataset, respectively. Figure 4 illustrates how these moments can describe a PSD.

For the two-fluid CFD models, a method of moments formulation was developed, which significantly reduced the number of equations needed to describe the particle phase [11,12]. A typical particle size distribution consists of 23 to 128 elements, while a method of moments can describe this same PSD using as little as five moments in most cases [13]. That reduction results in a significant improvement in the computational time.

Considerations with PSD Data

As noted above, PSDs are probably a key design parameter for your particle technology project. Understanding how the PSDs are obtained and in what form the data is presented is paramount to the success of that project. Some considerations should be used when evaluating PSD data, including;

- Is one representative particle size sufficient for your design calculations/model, or should you be considering the particle size distribution? If using statistical moments, are you using enough of them? Is the fines level important? What would the addition or removal of smaller or larger particles do to my product or the entire process?
- Has the sampling been done correctly? Is the sampling being done at the right place or places and at the right time? Is the sampling procedure sufficient to collect a representative sample free from segregation and operator error? Can the entire sample be used for the particle size measurements, or is a reduced sample size needed? If so, what assurances are in place to establish that the sample is being riffled correctly? Are there representative standards that can be used to test the analyzer, sampling, riffling, etc.?
- Is the correct analyzer being used for the data that is needed? Is measuring error a concern (i.e., how does the accuracy of the particle size distribution measurement or the representative particle size affect my calculations)?
- Should the individual particle size or the cluster/agglomerate size be considered in the calculations? Do the clusters/agglomerate need to be broken up prior to analysis?
- Do you need to be concerned about interparticle forces such as electrostatics with respect to sampling and analysis?
- Is particle attrition during analysis (i.e., sampling dispersion) an issue? Some analyzers are better with attrition than others. Some analyzers have controls that allow for the optimum level of dispersion while minimizing attrition.
- Will the particles swell or break up if put into a liquid? If so, consider using a dry technique for getting the PSDs. and
- Does the PSD look right? If not, consider the failure mode (i.e., sampling, analyzer operation, etc.)

Summary

PSD data should be collected with the end use in mind. In other words, the application should dictate what analyzer should be used and how the data should be presented. If a correlation works well with one representative particle size, then perhaps the median or Sauter mean particle size is all that is needed. If the size and shape of the distribution are known to affect the hydrodynamics, then the whole PSD may need to be considered. If CPU requires limiting that capability, then perhaps a method of moment technique with those calculations may be needed.

Understanding how the particle size will be sampled and measured and how that data needs to be presented is often paramount to a project's success. Don't assume the PSD is correct, and if it is correct, is it correct for you?

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https://www.partec.info/en

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https://waset.org/powder-technology-and-particle-science-conference

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