Synthesis of Bulk Solids Processing Systems

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A systematic procedure for the synthesis of bulk solids processing systems guides the user in an evolutionary manner to generate flowsheet alternatives to transform a solids feed with given attributes into products with desired attributes. The first step is to select a functional structure based on the primary objective of the processing system. This is followed by the selection of equipment for the functional structure. In the third step, the functional structure is expanded by adding other necessary unit operations to form a flowsheet. This is accompanied by the assignment of a suitable equipment unit to each unit operation. Storage and transportation units are then added to the flowsheet in the fourth step. Rules and heuristics to help make decisions at each step are presented, as well as simulations used to evaluate flowsheet alternatives. Design variables are identified and short-cut models based on discretized population balance equations are used to quantify equipment performance. The procedure is illustrated with commercial examples including the production of potash, pesticide granules, aspirin and an azo dye.

Introduction

The chemical processing industries in the developed countries have been gradually shifting from commodity chemicals towards high-value-added chemicals. Examples include pharmaceuticals, agricultural chemicals, and specialty chemicals, which are mostly sold in solid form. Better methods for the design and synthesis of processes involving solids have been identified to be a critical need in making such a shift (Nelson et al., 1995; Jacob, 1998).

A solids processing system can be divided into three major sections. The first section involves chemical and phase changes in which solid particles are generated, consumed, or transformed; reaction, crystallization, and dissolution belong to this section. The product coming out of the reactor or crystallizer is often a slurry of solid crystals. Separation processes such as flotation, filtration, dewatering, and drying constitute the second section. Most of the liquid content of the original slurry is removed at this stage. With moisture content below about 5%, the solids leave the second section in bulk form. The third section of a solids processing system deals with crushing, agglomeration, blending, and size classification of these bulk solids. In this step, product specifications such as size, shape, and composition are met.

A hierarchical design procedure for the first two sections of the process train has been developed (Rajagopal et al., 1992). An extension of this work for the separation system around a crystallizer, including filtration, washing, dewatering, and recrystallization, has also been considered (Chang and Ng, 1998). However, a systematic procedure for synthesizing bulk solids processes is not yet available. This is despite the numerous fundamental investigations on solids behaviors and equipment performance in the area of powder technology.

This article presents a systematic procedure that guides the user to synthesize flowsheet alternatives for bulk solids processing systems. The aim is to transform the solids feed with given attributes into products with desired attributes. Rules and heuristics are provided for making decisions at each step of the procedure. Equipment and cost models provide quantitative measures for screening process alternatives.

Procedure

The step-by-step procedure, summarized in Table 1, is discussed alongside an example on the production of potash. Simulation results are also presented to highlight the salient features and trade-offs in the process. This is followed by additional examples which highlight the procedure from other angles.

Step 1: Selection of Functional Structure

Figure 1 shows a generic flowsheet of bulk solids processing systems which serves two separate but related objectives:

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Table 1. Step-by-Step Procedure for Bulk Solids Processing System Synthesis

Step 1	Selection of functional structure
Step 2	Selection of equipment for functional structure
Step 3	Generation of flowsheet configurations
Step 4	Selection of storage and transportation equipment
Step 5	Evaluation of alternatives

mixing-demixing and size change. *Mixing-demixing* refers to blending and classification, while *size change* includes size reduction (crushing) and size enlargement (agglomeration). Upstream of these operations, the input may need a pretreatment step such as drying, addition of liquid, or solid-solid separation. The design for such operations is not considered in this article.

Table 2 presents the rules for selecting between mixing-demixing, size reduction, and size enlargement functional structures. Basically, this is determined by the given feed attributes and product specifications. If mixing-demixing is the primary objective, the structure is usually only a single unit operation. The only problem is equipment selection and design. On the contrary, if size change is the primary objective, the structure is more involved and includes mixing-demixing operations, as indicated by the up and down arrows in Figure 1. The size reduction structure is chosen in cases where all feed streams have particles with a size larger than the desired product. If any of the feed streams have particles with a size smaller than the product, we must choose the size enlargement structure which can include a size reduction unit, as will be discussed in Step 3. Also, if the desired product is in the form of granules or tablets having a fixed composition of different materials, a size enlargement structure is needed because the ingredient particles must have much smaller sizes than the final granules or tablets.

Step 1 for potash process

Potash (KCl) is a widely used fertilizer for crops. The Food and Agriculture Organization has predicted the world demand for this fertilizer to be above 60 million tonne per year by the turn of the century (Hignett, 1987). It can be produced using fractional crystallization from sylvinite ore, which is a mixture of KCl and NaCl (Rajagopal et al., 1988). The KCl crystals obtained from the crystallizer are usually of size less than 1 mm, while it is desired to produce granules of size about 2–4 mm for ease of spreading. The input information

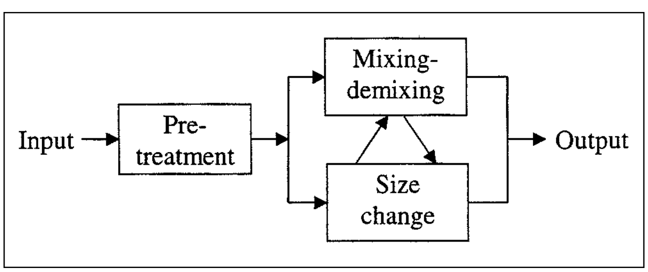


Figure 1. General functional structure of bulk solids processing systems.

Table 2. Rules for Selection of Functional Structures

Rule 1	Choose mixing-demixing structure if the objective is blending or classification of the input(s).
Rule 2	Choose a size reduction structure if the product(s) has to be of a smaller size than *all* available feed(s).
Rule 3	Choose a size enlargement structure if the product(s) has to be of a larger size than *any* of the available feed(s).
Rule 4	Choose a size enlargement structure if the product has to be in the form of granules or tablets having a fixed composition.

is compiled in Table 3. The crystal size is smaller than the required size. Clearly, a size enlargement structure is needed for this process (Rule 3).

Step 2: Selection of Equipment for Functional Structure

After choosing a functional structure, an equipment unit should be assigned to perform the function. Extensive description and comparison of various solids processing equipment can be found in Walas (1988), Fayed and Otten (1997), and Perry et al. (1997), among others. Table 4 gives a summary of the most common types along with their typical performance characteristics.

To assist in the selection process, major equipment types are classified into various classes according to the primary mechanism involved (Figure 2). Rules for equipment selected are listed in Table 5. Some are general rules (Rules 5 to 7), which are applicable to any unit operation, while others (Rules 8 to 22) are specific to a particular unit operation. Discussions on some of the rules are provided below. If there is more than one possible equipment type for the given objective, we should always keep all of them as alternatives.

Blending

Blending of solids can be achieved by diffusive, convective, or shear movement of particles (Carson et al., 1996). Good performance in a blending process can be achieved by match-

Table 3. Input Data for Potash Process Example

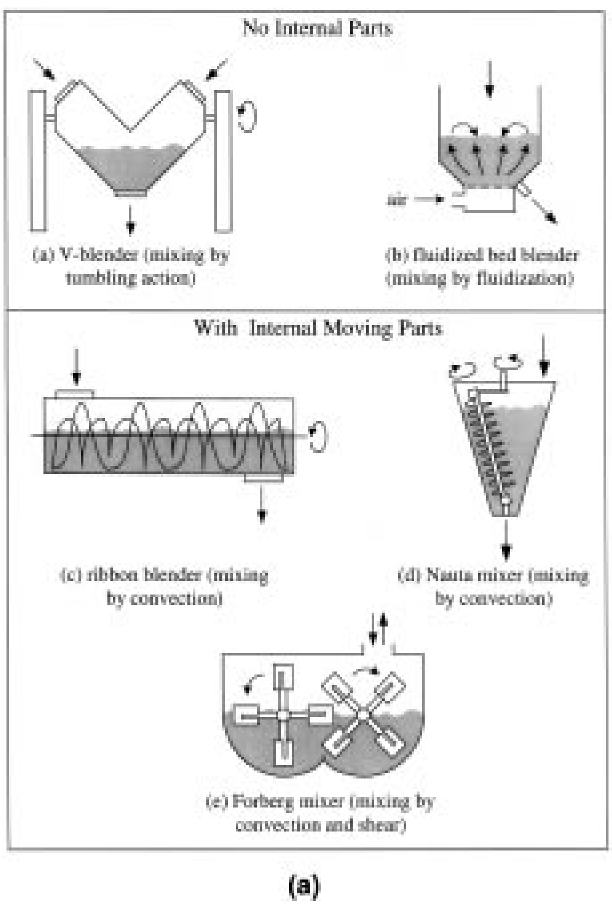
Feed Specifications	
Production rate of potash (based on dry KCl)	2×10^5 tonne/y
Moisture content (kg water/kg dry solid)	0.01
Source of particles	MSMPR crystallizer
Solids bulk density	1,000 kg/m^3
Crystallization System	
Mass of crystal per unit volume of slurry	43.9 kg/m^3
Crystal density	1683.45 kg/m^3
Crystal shape factor*	0.5235
Dominant crystal size	200 μm
Product Specifications	
Fraction smaller than 2 mm	Less than 10%
Fraction larger than 4 mm	Less than 10%
Energy Cost	
Electricity	7.5 ¢/kWh

*Defined as the ratio of the volume of a crystal to the volume of a cube with sides equal to the crystal size.

Table 4. Typical Performance Characteristics of Selected Bulk Solids Processing Equipment

Equipment Type	Typical Max. Capacity	Typical Power Consumption	Dry/Wet	Batch/Continuous
Blending				
Twin shell (V-blender)	5 m^3	5 kW/tonne	D,W	B
Fluidized bed	1,000 m^3	20 kW/tonne*	D	B,C
Ribbon blender	15 m^3	12 kW/tonne	D	B,C
Nauta mixer	60 m^3	3.5 kW/tonne	D	B,C
Forberg mixer	5 m^3	20 kW/tonne	D	B
Size Classification				
Vibrating screen	65 m^2	0.125 kWh/tonne	D	C
Air classifier	5,000 m^3/h	10 kWh/tonne*	D	C
Size Reduction				
Jaw crusher	1,200 tonne/h	0.7–2.5 kWh/tonne	D	C
Roller crusher	1,000 tonne/h	0.3–0.7 kWh/tonne	D	C
Fluid jet mill	6 tonne/h	300–1000 kWh/tonne	D	C
Hammer mill	850 tonne/h	2–3 kWh/tonne	D	C
Pin disc mill	35 tonne/h	1.5–300 kWh/tonne	D	C
Ball mill	300 tonne/h	4–25 kWh/tonne	D,W	B
Size Enlargement				
Roller compactor	75 tonne/h	2–16 kWh/tonne	D	C
Tableting machine	1 tonne/h	7–30 kWh/tonne	D	C
Drum granulator	800 tonne/h	0.8 kWh/tonne	W	C
Pan granulator	800 tonne/h	0.8 kWh/tonne	W	C
Fluidized-bed granulator	50 tonne/h	10 kWh/tonne*	W	C

*Based on power consumption for blower.



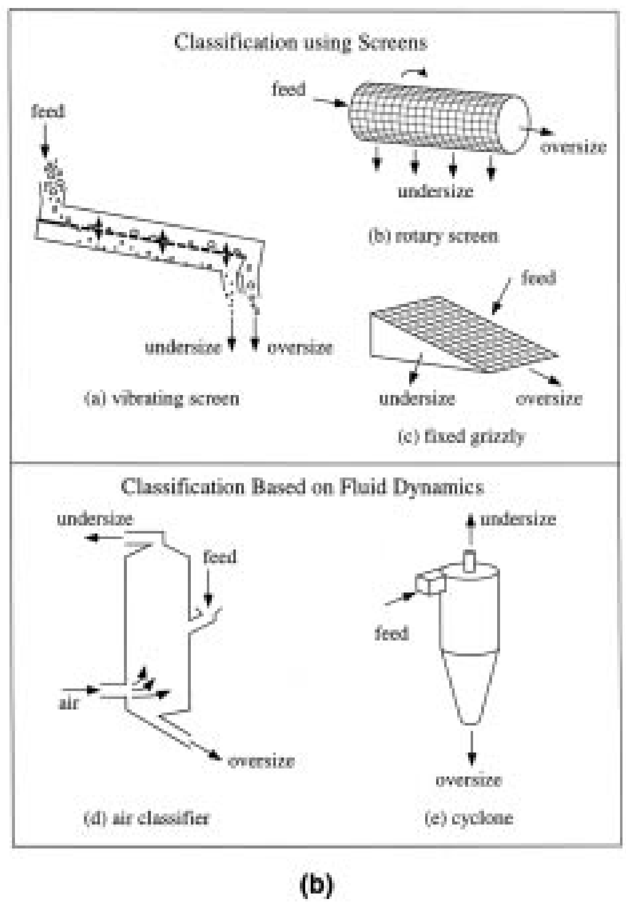


Figure 2. Selected types of (a) blenders and (b) size classification equipment *(continued)*.

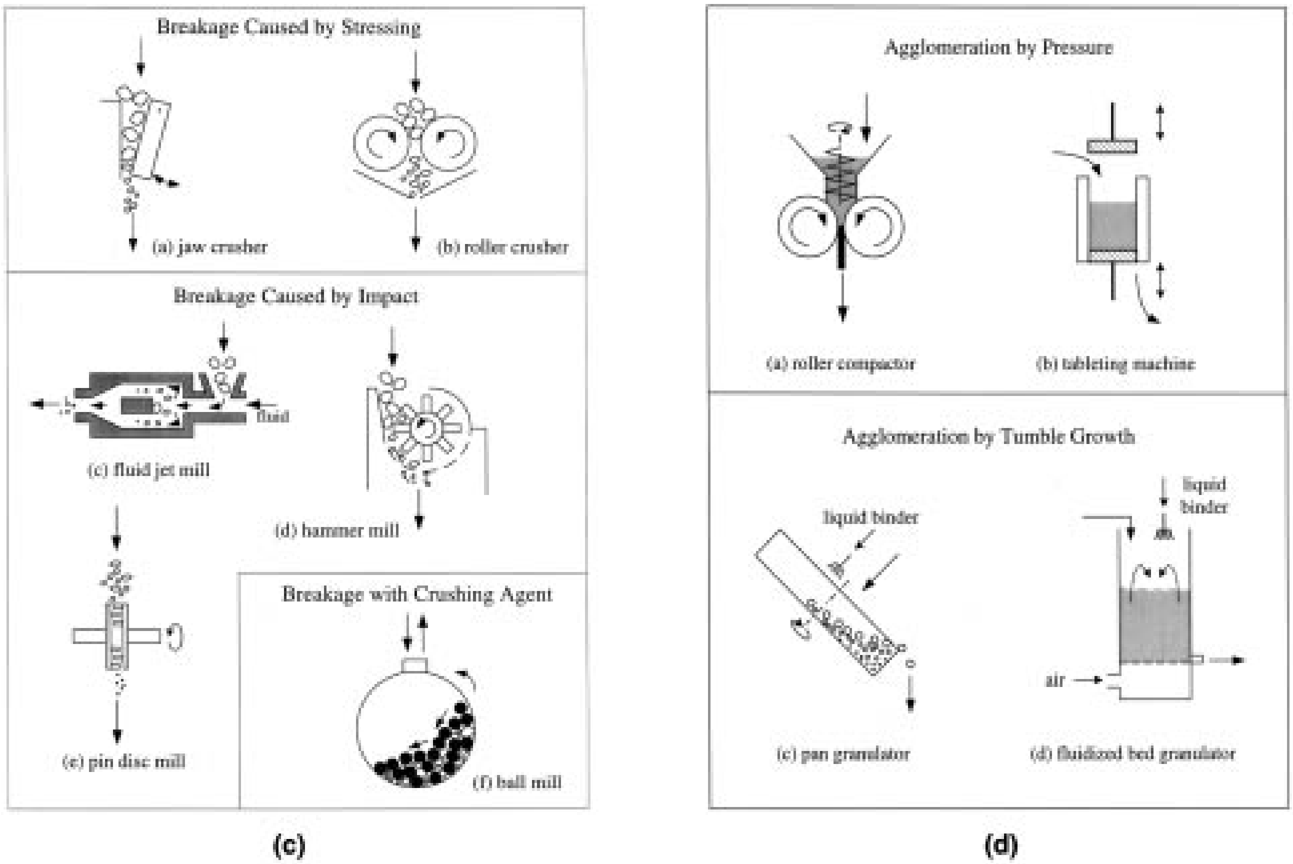


Figure 2. Selected types of (c) size reduction equipment and (d) size enlargement equipment.

ing equipment and material properties. Two most important factors to be considered are the flow property and possibility of segregation. Solid particles can either be free-flowing or cohesive. A rough measure of bulk solids flowability is the angle of repose, which is the angle formed by the limiting natural slope of the free surface of a heap of bulk solids with a horizontal plane (Woodcock and Mason, 1987). The angle of repose is not specific for a certain material, but depends on properties such as shape, size, porosity, and moisture content. For these reasons, this quantity by itself does not provide satisfactory information on flowability. Carr (1965) discussed a method for predicting flowability of solids, which is also based on measurable parameters such as compressibility and particle-size distribution (PSD). Cohesiveness can also be predicted using Geldart's classification of powders based on fluidization properties. Very fine powders (less than about 50 μm in size) are generally cohesive (Geldart, 1973).

In mixing free-flowing materials, homogeneity is often difficult to achieve due to the segregation tendency. Williams (1990) discussed possible mechanisms of segregation, all of which are basically caused by size or density difference. The width of a PSD can be represented using the coefficient of variation (CV) (dimensionless)

$$CV = \frac{d_{84} - d_{16}}{2\,d_{50}} \tag{1}$$

where d_{16}, d_{50}, and d_{84} are particle sizes corresponding to 16%, 50%, and 84% weight fractions on the cumulative PSD. Figure 3 gives an estimate of flowability based on the angle of repose, and indicates a region where segregation may be a potential problem. This segregation tendencies can be compensated by convective action of auxiliaries such as paddles or stirrers (van den Bergh, 1994). However, care should be taken because action of internal moving parts can introduce shear and stress on the particles, causing friable particles to break. Also, moving internal parts usually need lubrication fluids which can be a source of contamination to the product. Usually, excessive breakage is not found in machines with internal auxiliaries moving at a speed below 100 rpm, such as ribbon and Nauta mixers (Figure 2a, items c and d). The internal parts of such mixers also push the particles towards the exit point and keep them from sticking to the walls, so they can also handle slightly cohesive materials.

In mixing very cohesive materials, the important step is to break down the pockets formed by the cohesive particles. A shear and convective mixer with higher speeds, such as a Forberg mixer (Figure 2a, item e), is suitable for this purpose. Segregation is not a problem in a cohesive mixture. If a cohesive material is mixed with a free-flowing one, the cohesive particles will create a stable pocket resistant to segregation (Harwood et al., 1975). Breaking such pockets is more difficult if the cohesive material is present in a much smaller quantity than the other ingredient. Experimental data by Orr

Table 5. Rules for Equipment Selection

General Rules	
Rule 5	Avoid using equipment with internal parts moving faster than 100 rpm for easily breakable materials, unless particle breakage is allowed.
Rule 6	Use parallel units if the required capacity exceeds the typical maximum capacity of one single unit.
Rule 7	Use equipment with no internal moving parts if product contamination is to be strictly avoided.
Rules for Blenders	
Rule 8	Use tumbling mixers to handle free-flowing particles with relatively uniform particles (CV less than about 25% and no density difference).
Rule 9	Use equipment with internal parts such as stirrers or pedals to mix easily segregated materials.
Rule 10	Use shear or convective mixers to handle cohesive materials.
Rule 11	In the case of mixing cohesive materials, use two mixing stages if the proportion of one ingredient is less than 0.5%.
Rules for Size Classification	
Rule 12	Use a screen for size classification whenever possible.
Rule 13	Consider using air classifiers with fines removal if there is a potential dusting problem.
Rules for Size Reduction	
Rule 14	Use a proper size reduction ratio to avoid generation of too many fines (typical maximum values are: 8 for jaw crusher, 3 for roller crusher; 300 for fluid jet mill, 40 for hammer mill, 400 for pin disc mill, and 100 for ball mill).
Rule 15	Use several crushers in series if the desired reduction ratio is larger than the typical maximum value.
Rule 16	Use equipment with auxiliaries for continuous cooling, if the material is heat sensitive, has low melting point, or contains components that evaporate easily.
Rule 17	Use equipment with replaceable crushing agent if the material being handled is very hard or highly abrasive.
Rules for Size Enlargement	
Rule 18	Use pressure agglomeration if it is desired to have high-strength products.
Rule 19	Use pressure agglomeration if the wettability of the material is low or if the material should not be in contact with liquid.
Rule 20	Use pressure agglomeration if the product should be in specific shapes (such as tablets or briquettes).
Rule 21	Use tumble agglomeration if the material is too hard or abrasive, too brittle, or too elastic.
Rule 22	Use tumble agglomeration if it is desired to have a relatively high porosity, easy-to-dissolve product.

and Shotton (1973) show that the degree of homogeneity of a mixture decreases as the ratio of minor to major ingredient decreases. One way to achieve better mixing in such cases is to mix in two stages, which is recommended if the minor ingredient is less than 0.5% (van den Bergh, 1994). The minor ingredient is first distributed in a portion of the other constituent, and the resulting mixture is then mixed with the rest.

Solids classification

In this article we only consider classification processes based on size. The choice of equipment depends on the particle-size range the equipment handles. A convenient parameter to characterize this size range is the cut-size, which is the size of a particle having a 50/50 probability of ending up

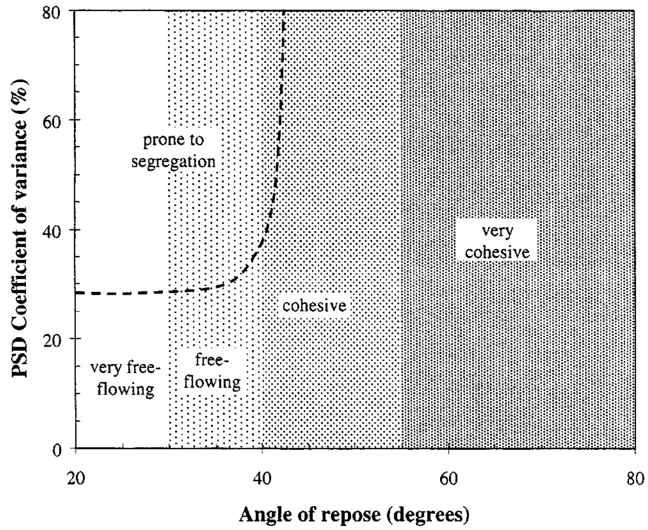


Figure 3. Estimated bulk solids flow properties based on angle of repose and difference in size.

in the underflow or overflow. Figure 4 shows the range of applicability of selected size classification equipment.

Screening is the cheapest and simplest method. When abrasive materials are involved, wear problems are inevitable. Even in this case, a screen is still the best choice because we only need to replace the deck instead of the whole equipment. Other equipment units such as air classifiers and cyclones are based on elutriation phenomena. They give less sharp separation, but are particularly useful when the desired cut-size is less than about 1 mm.

Size reduction

Selection of size reduction equipment strongly depends on the feed size. The lower triangle in Figure 5 shows the input-output particle size map for selected crushers. The typical size reduction ratio, defined as the ratio of initial to final size, is different for each equipment unit; in general it ranges between 4 and 400. An excessively large size reduction ratio can cause generation of very fine particles. Such particles can

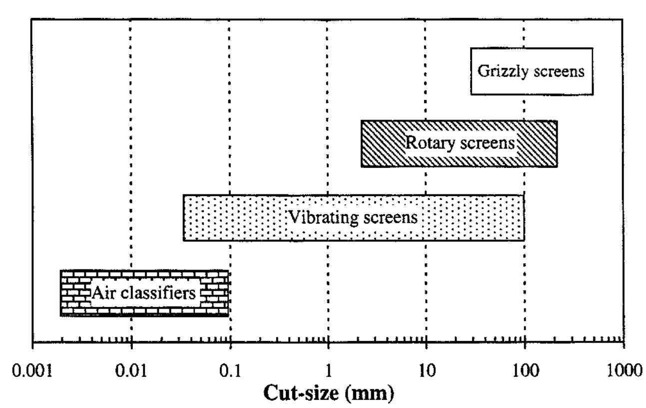


Figure 4. Range of applicability for selected size classification equipment.

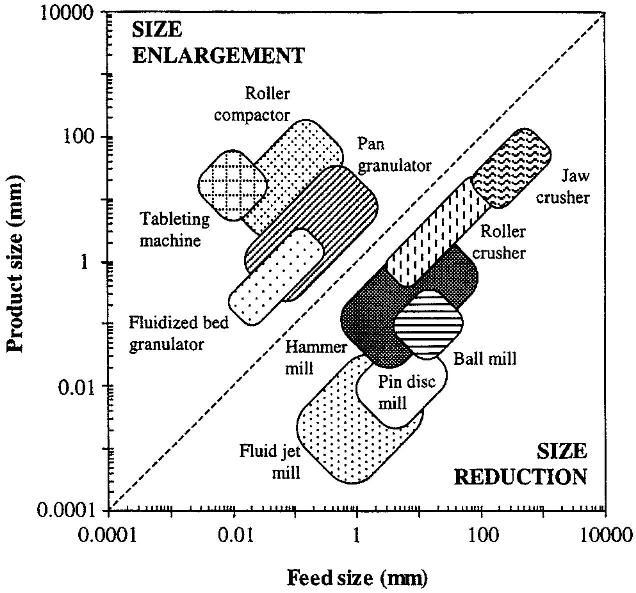


Figure 5. Feed and product size map for selected size change equipment.

create dusting problems, leading to health problems and severe explosions. Thus, if a reduction ratio larger than the typical value is desired, the size reduction should be performed in several steps in series. The size reduction ratio in an equipment unit can sometimes be adjusted with different settings. For example, the reduction ratio in a roller crusher depends on the rotational speed of the rolls and the gap between them.

Size reduction is a very energy intensive process with up to 99% of the supplied energy being converted to heat (Hixon et al., 1990). The liberated heat can cause melting and evaporation of some of the components, and cooling is sometimes needed. The energy requirement in kWh/tonne for size reduction depends strongly on particle size, as suggested by the well-known Bond's formula

$$E = 10 W_i \left(\frac{1}{\sqrt{X_P}} - \frac{1}{\sqrt{X_F}} \right) \qquad (2)$$

where W_i is the work index ($\text{kWh} \cdot \text{tonne}^{-1} \cdot \mu\text{m}^{0.5}$), the value of which depends on the material, and X_F and X_P are the feed and product size in μm, respectively. Work index for various materials can be found in Walas (1988) and Perry et al. (1997).

Very hard or abrasive materials can cause damage to equipment. In handling these materials, use replaceable crushing agents, such as the balls in a ball mill, for equipment parts which are prone to excessive wear. After these crushing agents wear out, we only need to replace the crushing agents instead of the whole unit.

Size enlargement

Basically there are two types of agglomeration: pressure agglomeration and tumble growth agglomeration (Sommer, 1988; Walas, 1988; Pietsch, 1996, 1997a,b). In pressure agglomeration, external forces are applied to particles to cause compaction and sometimes deformation of particles. However, pressure agglomeration may also cause breakage and hence is not suitable for materials with low tensile strength. It is also not suitable for abrasive materials since they can cause excessive wear to the equipment, as well as product contamination. In tumble growth agglomeration, a certain amount of binding liquid is added to cause colliding particles to stick to each other by capillary effects. This method works only if the binding liquid wets the surface of the material. The absence of pressure leads to the formation of agglomerates with higher porosity. Because of the high porosity, this product features a higher dissolution rate for the same solvent, as compared to a pressure-formed agglomerate. The tumbling action also tends to result in the formation of spherical granules.

The selection of size enlargement equipment should also be based on the desired particle size. For example, a pan granulator can only produce agglomerates no larger than about 50 mm (see Figure 5). In fluidized-bed granulation random and rapid movement of particles causes some of the new agglomerates to break, and larger particles are more difficult to fluidize. These facts limit the size of agglomerates produced via fluidization to about 2–3 mm at the most. On the other hand, fluidized beds may be favored if simultaneous drying is desired. Pietsch (1991) provided more detailed equipment selection criteria for size enlargement.

Step 2 for potash process

Having chosen the size enlargement structure, the next step is to choose an appropriate equipment unit. The given feed size allows the use of either pressure or tumbling method (Figure 5). At this point, we have no basis to favor one method over the other. However, since compaction has been widely used in many commercial plants (Middleton et al., 1983; Medemblik, 1983), it is considered for this example. The pan granulator and fluidized bed are kept as process alternatives.

Step 3: Generation of Flowsheet Configurations

For blending and classification structures, the only consideration is whether or not a pretreatment unit is necessary. For size reduction and size enlargement (Figures 6a and 6b, respectively), the flowsheet configurations can include pretreatment, post-treatment, blending, and classification. The size enlargement structure may also contain a size reduction unit. In both structures, there can be multiple input streams consisting of different components, which can enter the system at different places depending on their attributes. For example, if we have two input streams, and one needs size reduction but the other does not, then the latter should be fed after the size reduction unit. The boxes connected by dashed arrows represent optional units. Rules are available for determining which unit operations should be added to the functional structure (Table 6). This should also be accompanied by assigning an equipment unit to every operation appearing in the flowsheet, using the same equipment selection rules used in Step 2 (Table 5). All of these are discussed in detail below.

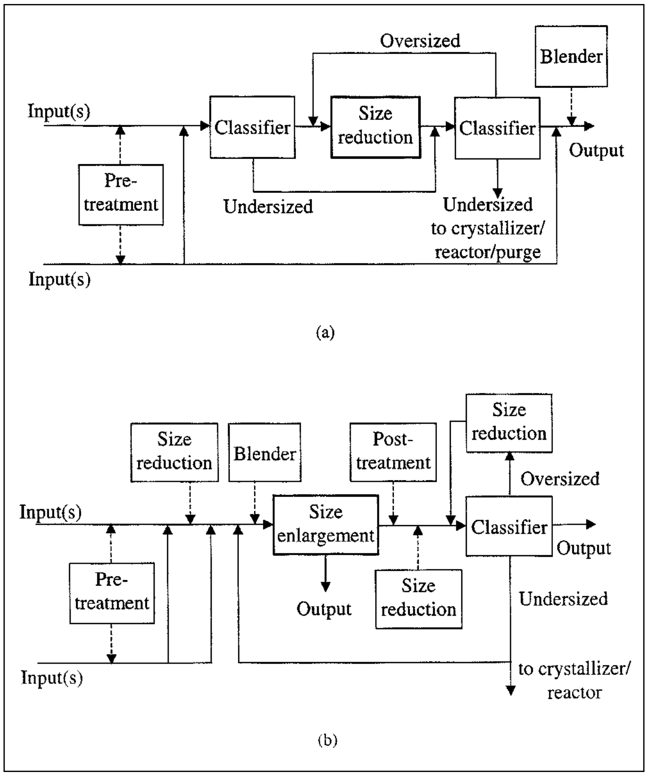


Figure 6. (a) Size reduction structure; (b) size enlargement structure.

Let us begin with pre- and post-treatment units (Rules 23 to 25). Only a few equipment units can handle wet systems. The presence of moisture will generate capillary forces that cause particles to agglomerate and stick to the wall. Normally, equipment for dry processes can only handle a stream with moisture content up to 2% (van den Bergh, 1994). If the feed contains more than this amount, we should use equipment suitable for wet processes. When this is not possible, we should remove the excessive moisture first by drying.

Furthermore, we seldom sell our products in a moist condition since agglomeration can lead to caking and "bag-set," that is, when the particles stick together to form a solid body of the shape of the bag. In post-treatment, any excessive moisture should be removed before packaging. Also, wetting of solid particles can sometimes be followed by dissolution, with the dissolved solute forming solid bridges connecting the individual particles in a drying process. Therefore, we should always be aware of larger particles having been formed in a dryer.

Similarly, rules can be developed for unit operations around a size enlargement unit (Rules 26 to 29). When a compactor is used, the product is normally in the form of a sheet so that a downstream size reduction unit is necessary. A wide PSD is required for pressure agglomeration, because more compact agglomerates can be formed by filling the void spaces between larger particles with smaller particles. Usually, good agglomerates can be achieved using a feed with a porosity around 8% (Klein and Popiel, 1983). This requirement can be met by adding smaller or larger particles, as necessary, into the feed stream of a compactor. The porosity of a solid stream can be estimated from the PSD using a model such as that of Ouchiyama and Tanaka (1984). However, the easiest way to obtain a reliable value is by measurement.

Table 6. Rules for Generating Flowsheet Configurations

Rules for Pre- and Post-Treatment Units	
Rule 23	Use an equipment unit that can accommodate wet processes if the moisture content of the input stream exceeds 2%. If it is not possible, install a dryer prior to the equipment to remove the excess moisture.
Rule 24	When drying is necessary, consider the possibility of performing it in a multifunctional unit such as a fluidized-bed granulator.
Rule 25	If a liquid additive is needed in a unit, add it directly into the unit. No additional mixing unit is required.
Rules for Unit Operations Around a Size Enlargement Unit	
Rule 26	Use a size reduction unit before size enlargement if the feed particles are too large to be fed to the selected size enlargement unit.
Rule 27	Use a blender before size enlargement if solid additives are needed or if particles of a different size should be added to provide the PSD required for proper operation of the size enlargement unit.
Rule 28	If the product has to be in the form of tablets or granules having a fixed composition, then the size of the feed particles to the agglomeration unit should be much smaller than the desired product size.
Rule 29	If it is difficult to maintain sufficiently constant feed rate to a tableting machine, use a preceding agglomeration process that yields large slugs, which are then crushed into larger particles than the original ones before going through the tableting machine.
Rules for Unit Operations Around a Size Reduction Unit	
Rule 30	If a large amount of fine particles are present in the feed to a size reduction unit, use a classification unit prior to size reduction.
Rule 31	If the output of a size reduction unit does not match the product requirement, it should be sent to a classification unit.
Rule 32	Consider combining two crushers having the same output destination, if their inputs have similar PSDs.
Rules for Unit Operations Around Blending and Solids Classification Units	
Rule 33	If a grinding (agglomeration) unit appears adjacent to a mixing unit, consider simultaneous grinding (agglomeration) and mixing.
Rule 34	If segregation may be a potential problem at some point after a blender, consider adding a small amount of liquid or relocating the blender downstream.
Rule 35	Use several screens in series if sharp-cut size separation is desired.
Rule 36	Use a sculping screen prior to the main screen if there are too many fines in the feed, to avoid low screen efficiency.
Rule 37	Oversized particles from a classifier should be sent to a crusher, and the product of this crusher should be recycled to the classification unit.
Rule 38	In the size enlargement structure, undersized particles from a classifier should be recycled to the size enlargement unit. Alternatively, if the particles are of a single pure component, they can be sent to an upstream crystallizer or reactor.
Rule 39	In the size reduction structure, if the classifier is located upstream of the size reduction unit, undersized particles should be combined with the output of the size reduction unit. If the classifier is located downstream and the undersized particles are of a single pure component, they should be sent to a crystallizer or reactor. Otherwise, they should be purged.

Formation of granules or tablets with a fixed composition requires the feed to be present in considerably smaller sizes than the granule or tablet itself; otherwise, composition uniformity cannot be assured. In addition, it is often desired to have products with uniform weights as in the production of pharmaceutical tablets. When dealing with nonfree-flowing powders, such uniformity is difficult to achieve. In this case, agglomeration in two stages should be considered. The powder is fed to the first tableting machine to produce large tablets, often referred to as slugs, which do not necessarily have uniform weights (Carstensen, 1984). These slugs are crushed into particles with a size greater than the original powder, which are then compressed in the second tableting machine to yield the final product.

Next, rules for unit operations around a size reduction unit are developed (Rules 30 to 32). Breakage of undersized particles passing through a size reduction unit can give rise to dusting problems. Even if no further breakage occurs, they become an unnecessary additional load to the unit. Therefore, a classification unit is often used prior to a size reduction unit to remove the excess amount of small particles. Such a unit is also needed after a size reduction unit if its output contains too many oversized or undersized particles.

Rules for unit operations around blending and solids classification units have also been developed (Rules 33 to 38). Certain types of blenders naturally cause agglomeration and breakage to occur to some extent. On the other hand, some size change equipment such as a ball mill and a pan granulator also performs mixing. Hence, it is possible to perform mixing and size change in one single equipment unit. Clearly, there is a limit to the degree of mixing or size change that can be achieved in such a multifunctional unit.

Resegregation can occur after a well-mixed state has been achieved. Such a problem may be encountered upon discharge from a batch mixer. Carson et al. (1996) proposed that we could eliminate this problem by adding a small amount of liquid to make the material slightly cohesive. Furthermore, it is advisable to perform mixing operation close to the end of the process. Accordingly, if a blender and a size reduction unit are both necessary, we always put the blender downstream of the size reduction unit in which segregation may occur.

If a classifier based on size is to be used, side streams may be generated. Oversized particles can be sent to a dedicated crusher. If another crusher is already present upstream of the classification unit, we should consider sending the oversized particles there. However, this may not be possible if the size of these particles is very different from that of the upstream crusher's input. For example, if 1-mm particles were fed to a jaw crusher whose gap had been set to handle a feed of 10 cm in size, the small particles would just pass through without being crushed. The destination of undersized particles depends on whether a size enlargement unit is present, the location of the classifier, and the particle composition. Particles of a single pure component can be sent to an upstream crystallizer where they can serve as seed crystals. In the size reduction structure (Figure 6a), the objective of the classifier upstream of a crusher is to provide a bypass for particles which do not need further crushing. In the size enlargement structure (Figure 6b), the underflow of the classifier provides small particles to the feed of the size enlargement unit.

If screening is used, we should always keep in mind that a screen has a limited efficiency as a result of the probability distribution of particles with various sizes to pass through the aperture. The probability that a particle with size d approaching a screen having square openings of dimension b and wires of diameter c in a normal direction will pass through is expressed as (Luckie, 1997)

$$p = \left(\frac{b-d}{b+c} \right)^2 \tag{3}$$

In industrial operations, the particle will have multiple chances to pass through as it slides along the screen. For n opportunities, the overall probability can be expressed as

$$q = 1 - (1-p)^n \tag{4}$$

It follows that near-cut-size particles ($b \approx d$) have a very low probability of passing. Therefore, we should always use an aperture of a larger dimension, normally 10–20%, than the desired cut size. Consequently, sharp-cut separation cannot be achieved and a series of screens must be used to improve separation sharpness.

Segregation due to size or density difference can cause bypassing, which results in the loss of some undersized particles to the product stream. For example, some particles smaller than the screen apertures do not pass through the screen due to the presence of a large amount of finer particles moving down the apertures. Bypassing reduces screen efficiency; therefore, it is desirable to remove excessive fines early by using a sculping screen.

Step 3 for potash process

The size enlargement structure for the potash plant can be expanded based on Figure 6b. It contains several units that need to be specified using the rules. The resultant flowsheet is depicted in Figure 7a. Crystals with a dominant size of 200 microns (Table 3) are suitable for the compactor input which can accommodate feed particles of size under 800 μm (Figure 5). Therefore, no size reduction unit is necessary before the compactor (Rule 26). The moisture content is lower than 2% so that drying is not necessary (Rule 23). From the calculated value of the variance of the PSD, the feed does not meet the feed requirement for the compactor. Therefore, a blender should be installed before the compactor, with the intention of mixing in another stream to modify the feed PSD (Rule 27). Our choice of size enlargement unit (compactor) does not provide mixing, and a separate blender is necessary (Rule 33). At this point, an experiment using the candidate compactor should be conducted to obtain the product size. The product leaves the compactor as sheets with dimensions of several hundred millimeters, which are then broken into flakes with a maximum dimension of 150 mm (Middleton et al., 1983). A roller crusher can be used as the breaker (Break) (Figure 5). Since the desired product should consist of particles with a size of 2–4 mm, an additional crusher referred to as the primary crusher (Crush 1) should be installed. A hammer mill is suitable for this primary crusher.

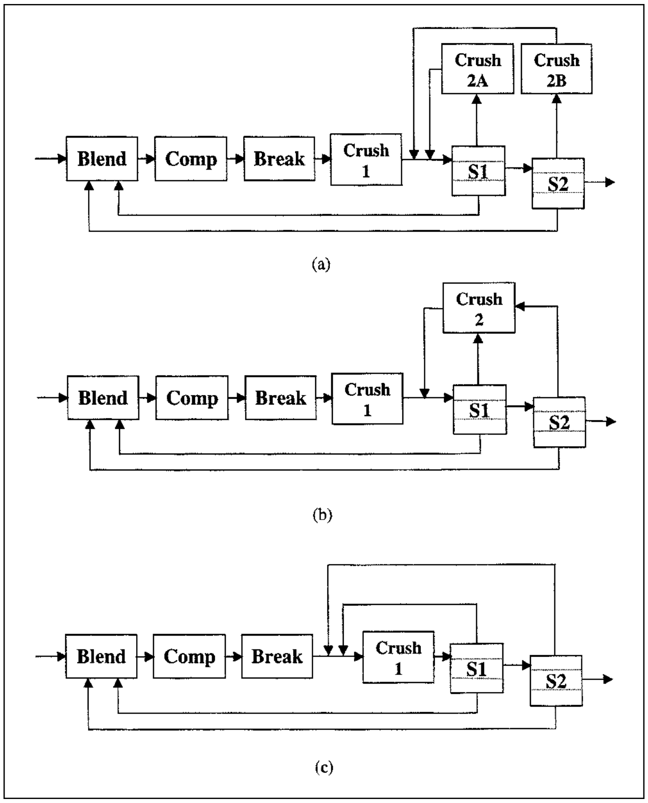


Figure 7. Step 3 for potash process.

(a) Base case flow sheet; (b) alternative generated by combining secondary crushers; (c) alternative generated by eliminating secondary crushers.

The output from Crush 1 is expected to consist of particles with sizes covering a wide distribution, so that a classifier must be installed. Since the desired cut-size is about 2–4 mm, a vibrating screen will be used (Figure 4, Rule 12). Since a sharp-cut is desired, we use two screens in series (Rule 35). The installation of the screens generates new streams that need to be assigned destinations. We send the oversize streams to secondary crushers (Crush 2A and Crush 2B) and their output stream back to screen S1 (Rule 37). Hammer mills can also be used for this task. The undersize streams should be recycled to the blender before the compactor (Rule 38). Because fines are mixed with relatively large particles, the PSD will be wide and we should use a blender with an internal part to minimize segregation (Rule 9). Since breakage is not a problem (Rule 5), we can choose the ribbon blender.

We can consider combining Crush 2A and 2B because their outputs go to the same screen (S1) (Rule 32) (Figure 7b). It is also possible to eliminate the secondary crusher (Crush 2) by feeding the recycle streams to the primary crusher (Figure 7c). Note that although the two crushers are of the same type, they might have different specifications. Further evaluation is needed to justify this decision of combining units.

Step 4: Selection of Equipment for Storage and Transportation

Extensive discussion of equipment for storage and transportation of solids can be found in Woodcock and Mason (1987). Rules for selection of solids handling equipment units are summarized in Table 7.

Table 7. Rules for Selection of Storage and Transportation Equipment

Rules for Storage	
Rule 40	Use intermediate storage bins for streams that have to be mixed in a certain proportion, or prior to a unit where starvation of feed can disrupt the process.
Rule 41	Use a dryer prior to an intermediate storage tank if the feed moisture content is too high.
Rules for Solids Transportation	
Rule 42	Use a belt conveyor to transport dry bulk solids horizontally, unless there are other special considerations.
Rule 43	Consider using a drag conveyor with a small overflow stream to transport the material to the feeder as an alternative to installing an intermediate bin, to avoid starvation problem.
Rule 44	Use a screw conveyor or weighing belt conveyor when the flow rate of the solids to be transported has to be controlled.
Rule 45	Use a bucket elevator to transport solids to a higher elevation.
Rule 46	Use a pneumatic conveyor if the dry particles to be transported may cause a dusting problem or for long distance transportation.

Storage units are necessary for the feed, intermediate product, and final product. They provide a continuous supply of raw materials and product in stock. Intermediate storage is sometimes necessary for various purposes. For example, if we have to mix two solid streams in a certain fixed proportion, it is important to anticipate disturbances that can cause the flow of either stream to fluctuate (Pietsch, 1997a). An example of such disturbances is the change in the PSD of a crusher output with time because of the probabilistic nature of breakage. If this stream is sent to a classifier, the flow rates of the classifier products will also fluctuate. Another disruption can be caused by starvation of the feed to an equipment unit. Maintaining a small overflow stream larger than the anticipated fluctuation in the feed point of such a unit can eliminate this problem (Pietsch, 1997a). Flow problems in storage units such as plugging and arching may also cause serious disturbances to the overall process. Therefore, efforts should be made to avoid poor flow properties, such as by controlling the hydrophobicity of the particles. Careful design of equipment also plays an important role in maintaining smooth operation (Carson and Marinelli, 1994).

Transportation of dry bulk solids is mainly accomplished mechanically using conveyors and elevators. Conveyors are generally used to transport materials to another place with approximately the same level. Belt conveyors find the most applications in solids processing plants, because they are the cheapest and simplest. Other conveyor types should be used for special reasons. Screw or weighing belt conveyors should be used if metering of solids flow is required (Middleton et al., 1983). Application ranges for various conveyors have been summarized by Reece (1995). Since gravity is often the driving force for solids flow in solids processing plants, equipment units are usually arranged from high to low elevation. If many units are involved in the process train, there is often a need to move solids to a higher evaluation. Also, a recycle flow to an upstream unit may give rise to the same need. In this case, elevators such as the commonly used bucket elevator should be considered.

Pneumatic conveying is an alternative to mechanical conveying, especially if dusting problem precludes the use of uncovered conveyors or elevators (Reece, 1995). Powders are transported in pipelines using a gas, normally air, as the carrying agent. The particles should be separated using a cyclone at the end of the line. Sometimes, it is necessary to use an inert gas such as nitrogen to prevent possible explosions. Geldart's group A, B, and D powders are suitable for pneumatic conveying.

Unwanted adhesion and agglomeration can be a serious problem in handling bulk solids. For example, slightly moist powder can cause a depository problem in a pneumatic conveyor line (Fayed and Otten, 1997). Moisture can also cause plugging in storage tanks because the particles stick to each other and block the discharge. To avoid these problems, the moisture content of the inputs to storage and transportation units should be carefully controlled (Rule 23).

Step 4 for potash process

In this step, the necessary solids handling equipment units are added. The final flowsheet after this step is shown in Figure 8. Maintaining a constant ratio of the feed to recycle stream is important because starvation or variation in size distribution in either stream will cause a change in porosity of the compactor feed (Pietsch, 1997a), which in turn affects the product strength. This ratio is kept constant by using storage bins for both recycle and feed streams (Rule 40) and weighing belt conveyors (WB) to control the flow rates (Rule 44). Belt conveyors (not shown in figure) are used to connect most of the units (Rule 42). A drag conveyor (DC) is used to transport the feed to the compactor to maintain a small overflow so that starvation can be avoided (Rule 43). Based on approximate elevation of the equipment units, two bucket elevators (BE 1 and BE 2) will be necessary (Rule 45). However, the final decision can only be made after a careful consideration of plant layout.

Step 5: Evaluation of Process Alternatives

The process alternatives are now compared to identify the best flowsheet. This evaluation can be guided by simulation using short-cut models and population balance calculations. Simulation of solids processing circuits involving a granulator and a crusher has been performed by Adetayo et al. (1995). They used semi-empirical models based on discretized population balance equation (PBE) for the granulator, crusher, and screen. Their models contain adjustable parameters that should be determined experimentally. Similar equipment models (Table 8) are used with equal-size discretized PSDs to solve the population balances.

The values of model parameters depend on the settings of certain operating variables such as those shown in Table 9. The relationship between model parameters and equipment operating variables is complex and, in practice, all model parameters have to be backed out using experimental data. The effect of parameters such as the residence time in crushers and agglomerators, and the opening size in screens on process performance can be determined using sensitivity analysis.

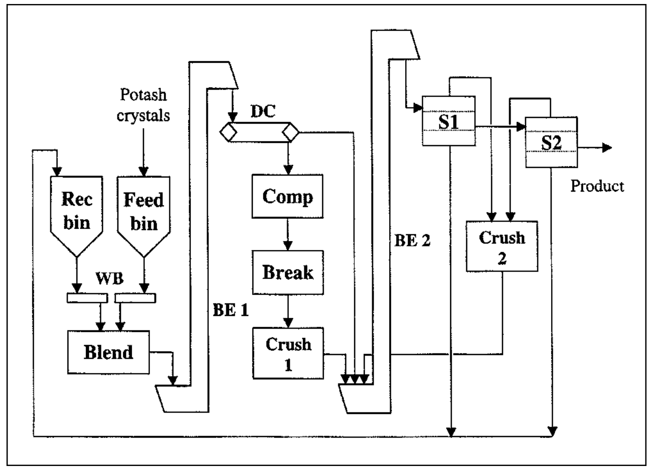


Figure 8. Final flowsheet for potash example.

Upon completion of the material balance calculations, we can determine the required equipment capacity, and, therefore, the related capital and energy costs. The screen capacity can be determined using the calculation procedure described by Osborne (1981), while the blender capacity can be found by multiplying the feed flow rate and the required mixing time. Available equipment costing models are generally based on capacity only and do not capture the effects of

Table 8. Equipment Models Based on Discretized Population Balance Equations

Continuous Blender	
$W_{f,k}$, $\omega_{fi,k}$ → [Blender] → W_p, ω_i; $k = 1, \ldots, m$	$\omega_i = \dfrac{\sum_{k=1}^{m} W_{f,k}\,\omega_{fi,k}}{\sum_{k=1}^{n} W_{f,k}}; \quad i = 1, \ldots, n$
Continuous Classifier	
W_f, ω_{fi} → [Classifier] → $W_{p1}, \omega_{i,1}$; ↓ $W_{p2}, \omega_{i,2}$	$\omega_{i,1} = \dfrac{W_f\,\omega_{fi}(1-P_i)}{\sum_i W_f\,\omega_{fi}(1-P_i)}; \quad i = 1, \ldots, n$ $\omega_{i,2} = \dfrac{W_f\,\omega_{fi}P_i}{\sum_i W_f\,\omega_{fi}P_i}; \quad i = 1, \ldots, n$
Continuous Crusher	
W_f, ω_{fi} → [Crusher] → W_p, ω_i	$\omega_i = \dfrac{\omega_{f,i} + \tau \sum_{j=i+1}^{n} b_{i,j}S_j\omega_j}{1 + S_i\tau}; \quad i = 1, \ldots, n$ (for binary breakage; $S_1 = 0$)
Continuous Agglomerator	
W_f, ω_{fi} → [Agglomerator] → W_p, ω_i	$\omega_i = \dfrac{\omega_{f,i} + \frac{1}{2}W_f\tau^2 \sum_{j=1}^{i-1} a_{j,i-j}\,\omega_j\,\omega_{i-j}}{1 + W_f\tau^2 \sum_{j=1}^{n} \frac{d_i^3}{d_i^3 + d_j^3} a_{i,j}\,\omega_j}; \quad i = 1, \ldots, n$ (for binary agglomeration)

Table 9. Operating Variables of Selected Bulk Solids Processing Equipment

Unit Operation	Equipment Type	Operating Variables	References
Blending	V-blender	Rotation rate Vessel geometry Filling level	Brone et al. (1998)
	Ribbon blender	Agitator speed Ribbon geometry Filling level	van den Bergh (1994)
	Nauta mixer	Screw revolutional speed Screw rotational speed Vessel geometry	van den Bergh et al. (1993)
Size Classification	Vibrating screen	Vibration frequency Vibration amplitude Screen length	Osborne (1981)
	Air classifier	Air-flow rate	Luckie (1997)
Size Reduction	Jaw crusher	Stroke speed Nip angle Gap between jaws	Bernotat and Schönert (1988)
	Roll crusher	Roll speed Gap between rolls	Hixon et al. (1990)
	Fluid jet mill	Fluid velocity Equipment geometry	Hixon et al. (1990)
	Hammer mill	Hammer rotational speed Discharge opening size	Ratcliffe (1972)
	Pin-disc mill	Disc rotational speed	Prior et al. (1990)
	Ball mill	Rotational speed Ball size Ball loading Filling level	Venkataraman and Narayanan (1998)
Size Enlargement	Pan granulator	Amount of binder Angle of inclination Rim height Rotational speed	Pietsch (1991)
	Fluidized-bed granulator	Amount of binder Air velocity	Pietsch (1997b)

changing operating conditions. Costs for solids storage and transportation are not accounted for in our cost estimation. These units are included in this procedure to make sure that the process works properly. However, process alternatives can be compared without them because they usually appear in all alternatives. This simulation is the last step of the proposed procedure, the entirety of which is depicted in Figure 9.

Step 5 for potash process

A computer code *BulkSolids* has been developed to simulate the potash process based on the basic flowsheet (Figure 7b). Equipment sizes and PSDs for each stream are calculated using the models, and the resulting PSDs are given in Figure 10. In this simulation, we make no attempt to model the process in the compactor and breaker, because little is known of the actual phenomena. The PSD for Stream 2 (Figure 10) is assumed for the breaker output. The primary crusher (Crush 1) and the secondary crusher (Crush 2) are assumed to be of different types with different parameter values. Parabolic breakage and a functional model for the specific rate of breakage are assumed (Hill and Ng, 1995). The complete list of model parameters used as a base case in this simulation is given in Table 10.

The total annualized cost (TAC) is defined as the sum of annualized capital and operating costs. The annualized

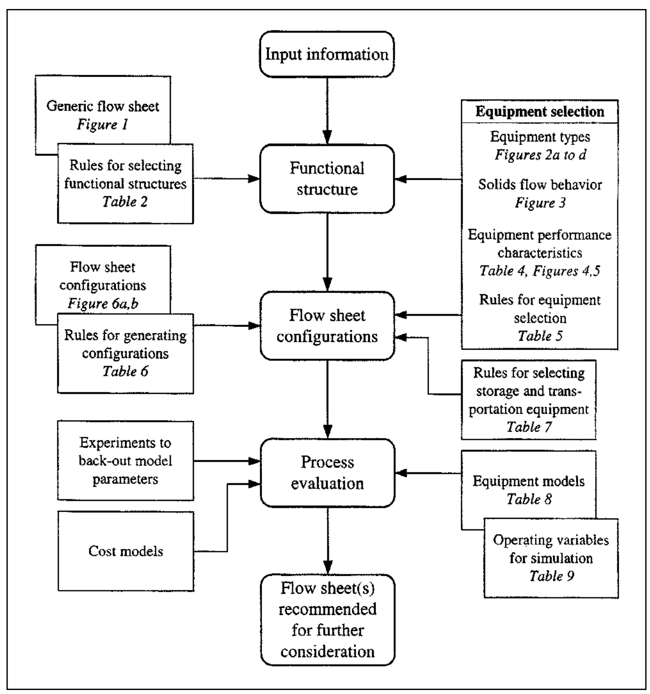


Figure 9. Procedure for synthesis of bulk solids processing systems.

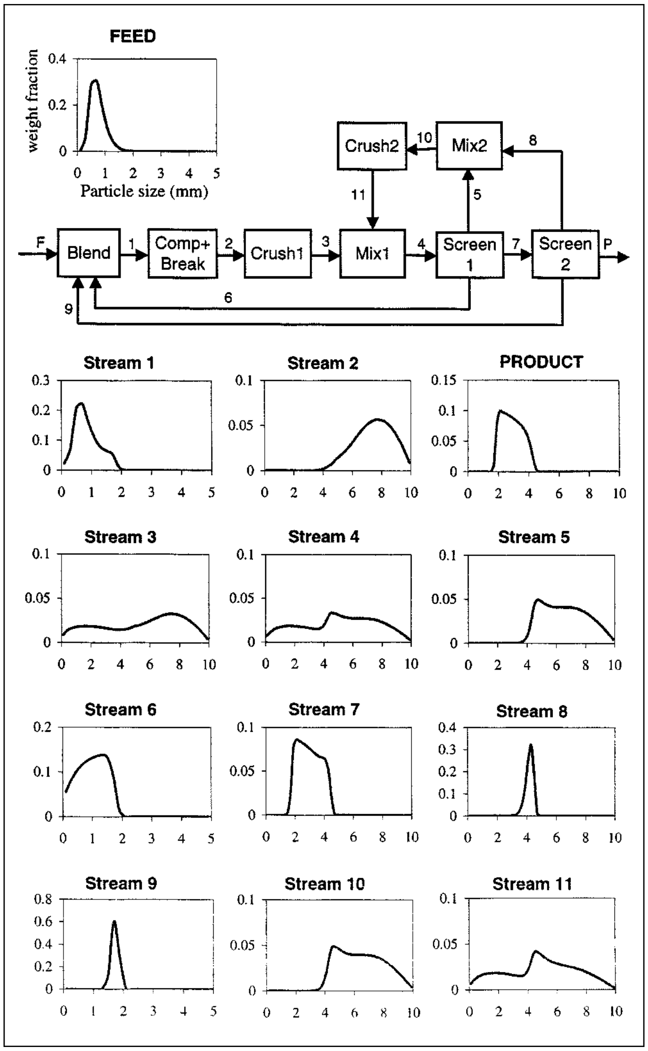


Figure 10. Simulation results for the PSD in each stream of the potash process base-case design.

equipment cost is calculated using a capital charge factor of 1/3 year, and includes the cost of the compactor, crushers, blender, and screens. Individual equipment costs are calculated using cost models available in the literature (Guthrie, 1969; Peters and Timmerhaus, 1980). The annual energy cost is the electric power requirement for the same equipment units. For crushing, the specific energy consumption is calculated using Eq. 2; while for other units, energy costs are estimated based on typical energy consumption per unit mass (Table 4). Figure 11 shows the contribution of each unit to the total equipment and energy costs for the base case.

As Table 9 indicates, the design variables for this process include the residence time in the crushers and size of screen openings. The impact of these variables on process economics can be studied by changing the variable from the base case values. It should be pointed out that some restrictions such as product size requirements would limit the range of permissible values. Simulation results indicate that when the fractions of oversized and undersized particles in the product are kept below 10% (see product requirements in Table 3),

Table 10. Model Parameters for Base-Case Potash Process

Blender	
Mixing time	10 min
Crushers	
Model for specific rate of breakage	$S_i = S_0 i^k$
Type of breakage	Parabolic $\left(b_{i,j} = \frac{6(ij - i^2)}{j^3 - j}\right)$
Material work index	$8.23\ \mathrm{kWh \cdot tonne^{-1} \cdot \mu m^{0.5}}$
Primary crusher	
Breakage rate constant, S_0	$0.012\ \mathrm{min}^{-1}$
Breakage order, k	0.86
Residence time, τ	3 min
Secondary crusher	
Breakage rate constant, S_0	$0.01\ \mathrm{min}^{-1}$
Breakage order, k	1
Residence time, τ	2 min
Size Classification	
Ratio c/b	1.3
Number of opportunities, n	80
First screen	
Upper deck openings	4.76 mm (4 mesh)
Lower deck openings	2.00 mm (10 mesh)
Second screen	
Upper deck openings	4.76 mm (4 mesh)
Lower deck openings	2.00 mm (10 mesh)

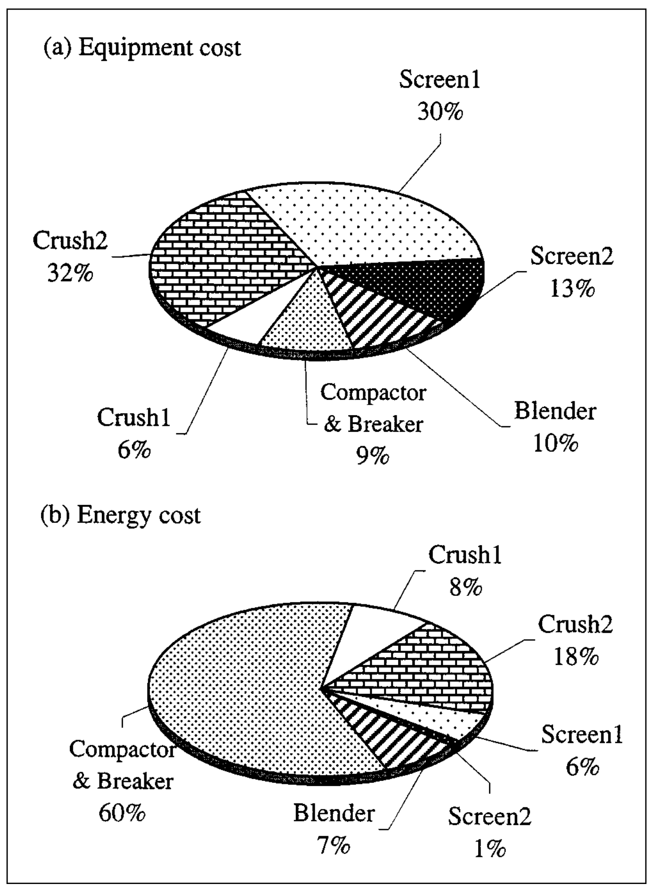


Figure 11. Cost distribution for the potash base-case design.

the d_{50} and CV of the product are about 2.8 mm and 28%, respectively. We also found that the feed to the compactor (stream 1) always has a wide PSD ($CV = 61-65\%$), which gives assurance that the requirement for pressure agglomeration is met.

Figure 12 illustrates the effect of changing primary crusher residence time. The longer this crushing time is, the smaller the amount of large particles in the output (stream 4). Using the same screen openings, the fraction of materials flowing into the overflows (streams 5 and 8) and the load to the secondary crusher will be reduced (Figure 12a). On the other hand, the recycle flow of smaller particles (streams 6 and 9) will increase, and, as a consequence, a larger blender, compactor, breaker, and primary crusher are needed (Figure 12a). The presence of a larger amount of smaller particles in streams 4 and 7 will also reduce the required screen areas (Figure 12b) and the total cost of the two screens (Figure 12c). Figure 12c shows that for residence times greater than 2 min, the energy cost increases as the primary crusher residence time increases. This is due to the greater energy requirement for crushing and the increase in compactor capacity. However, at low residence times, the energy consumption of the secondary crusher increases as the residence time decreases. This increase is responsible for the slope reversal in the total energy cost. Figure 12c also shows a local minimum in the TAC when the primary crusher residence time is about 2.5 min.

The effect of changing secondary crusher residence time is shown in Figure 13. For reasons similar to what has been described above, reducing this crushing time will increase the flow rates of streams 5 and 8, so that a larger secondary crusher is needed. At the same time, the flow passing the blender, compactor, and primary crusher becomes smaller, so again there is a trade-off leading to a minimum in the total cost. This minimum is achieved at a residence time of about 2 min. Simulation results also show that, at residence times below 2 min, the product would contain too many oversized particles, such that the coarse fraction product requirements given in Table 2 (maximum 5% larger than 4 mm) cannot be met. Meanwhile, at residence times above 3 min, the fine

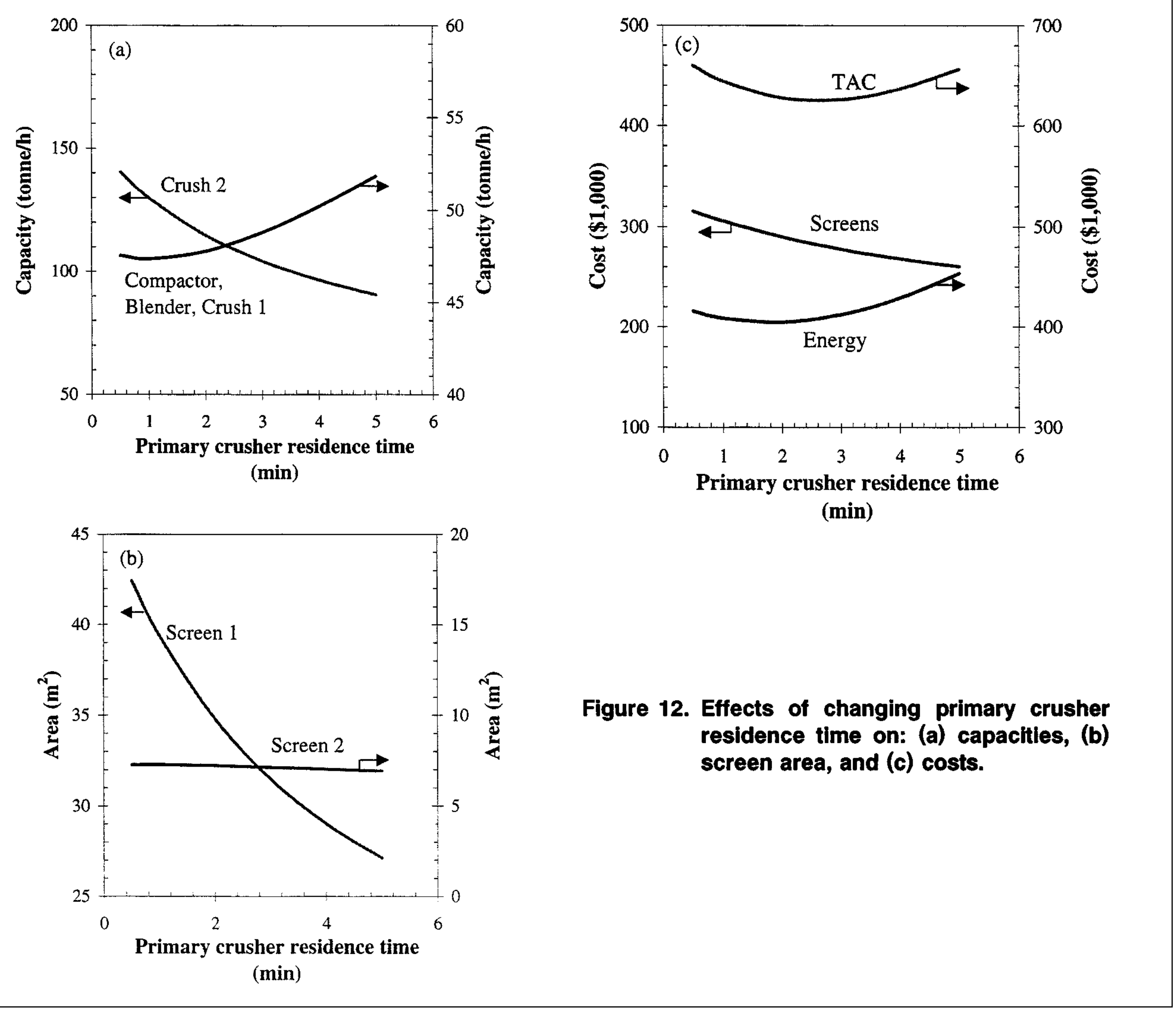


Figure 12. Effects of changing primary crusher residence time on: (a) capacities, (b) screen area, and (c) costs.

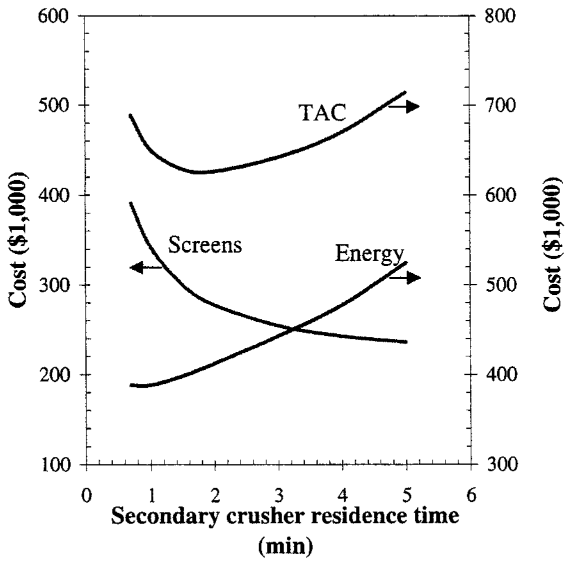


Figure 13. Effects of changing secondary crusher residence time on costs.

fraction product requirement (maximum 5% smaller than 2 mm) will be violated.

Table 11 shows the effect of using different screen openings on cost and PSD. When a larger lower-deck opening is used for the second screen (case 2), less fines are found in the product. However, an increase in the underflow (stream 9) leads to an increase in the blender, compactor, and primary crusher capacities. The TAC is increased by 13% to $710,100 in this case. Due to the presence of more fines in the recycle stream, the compactor feed has a wider PSD, as indicated by the higher value of *CV*. In case 3, a larger upper deck opening and a smaller lower deck opening are used for the first screen. As a result, the recycle stream (5) decreases and a smaller secondary crusher is needed. The underflow (stream 6) will also decrease and cause the blender, compactor, and primary crusher capacities to go down. Thus, a lower TAC is obtained. However, product requirements are not satisfied in Case 3. If screen openings could be varied continuously, we could have met the desired requirement by using slightly different screen openings for the two screens (Case 4) and still could have had a lower TAC compared to the base case.

Several process alternatives are generated during the flowsheet development. For example, in Step 3 we have chosen to use two screens in series because sharp-cut is desired. Simulations can be used to compare the base case (using two screens) with a case where only a single screen is used (Figure 14). The single screen has the same upper and lower deck openings as the two screens used in the base case, that is, 4.76 and 2.00 mm, respectively. Since the screen cost overwhelms other equipment costs, there is a substantial reduction in the total equipment cost by just using one screen instead of two. The TAC is reduced by about 13%, but we find that the product requirement is not met. This can, of course, be solved using smaller upper deck and larger lower deck openings (4.00 mm and 2.38 mm, respectively), but the TAC of this modified one-screen case is 32% higher than the base-case value.

Another process alternative considered in Step 3 is the combination of the primary and secondary crushers (Figure 7c). The secondary crusher is eliminated and both screen overflows are fed to the primary crusher. Calculation results reveal that this change implies a huge recycle flow through the crusher. This flow turns out to be so large that parallel screens are needed to handle it (Rule 6). Accordingly, both equipment and energy costs increase. As shown in Figure 15a, the TAC is found to increase by about 21% to $759,300. The largest contributor to this increase is the screen cost, as can be seen in Figure 15b. When longer crushing time (3.5 min) is applied and an upper deck opening of 5.66 mm is used for the first screen to avoid violating the product requirements, a lower TAC of $738,200 can be achieved. This modified case is still inferior to the base-case design (Figure 15a).

Table 11. Effects of Changing Screen Openings

	Case 1 (Base-case)	Case 2	Case 3	Case 4
Screen Openings (mm)				
Screen 1 upper deck	4.76	4.76	5.66	5.00
Screen 1 lower deck	2.00	2.00	1.68	1.95
Screen 2 upper deck	4.76	4.76	4.76	4.85
Screen 2 lower deck	2.00	2.38	2.00	2.00
Cost ($1,000)				
Equipment	641.8	707.5	610.0	620.0
Energy	412.3	474.3	380.8	393.3
TAC	626.3	710.1	584.2	600.0
Product Quality				
d_{50} (mm)	2.758	2.905	2.788	2.800
Coef. of variance (%)	28.3	24.1	30.8	29.7
% under 2 mm	9.1	1.5	11.5	9.6
% above 4 mm	6.7	7.7	10.5	9.8
Compactor Feed PSD				
Coef. of variance (%)	63.5	72.1	60.6	62.3

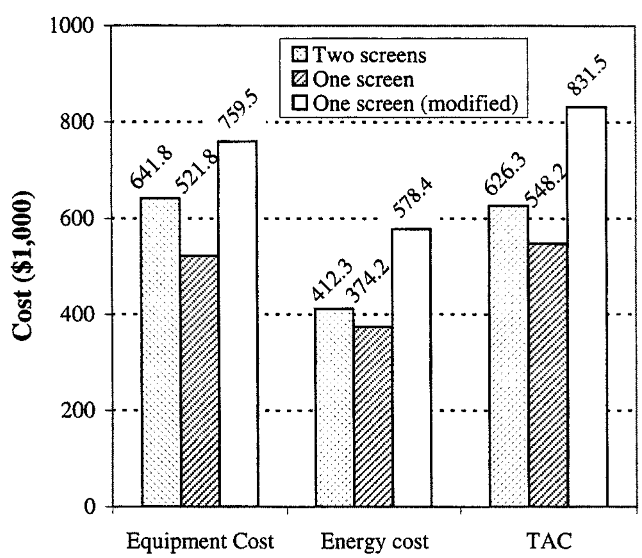


Figure 14. Comparison of costs between using one and two screens.

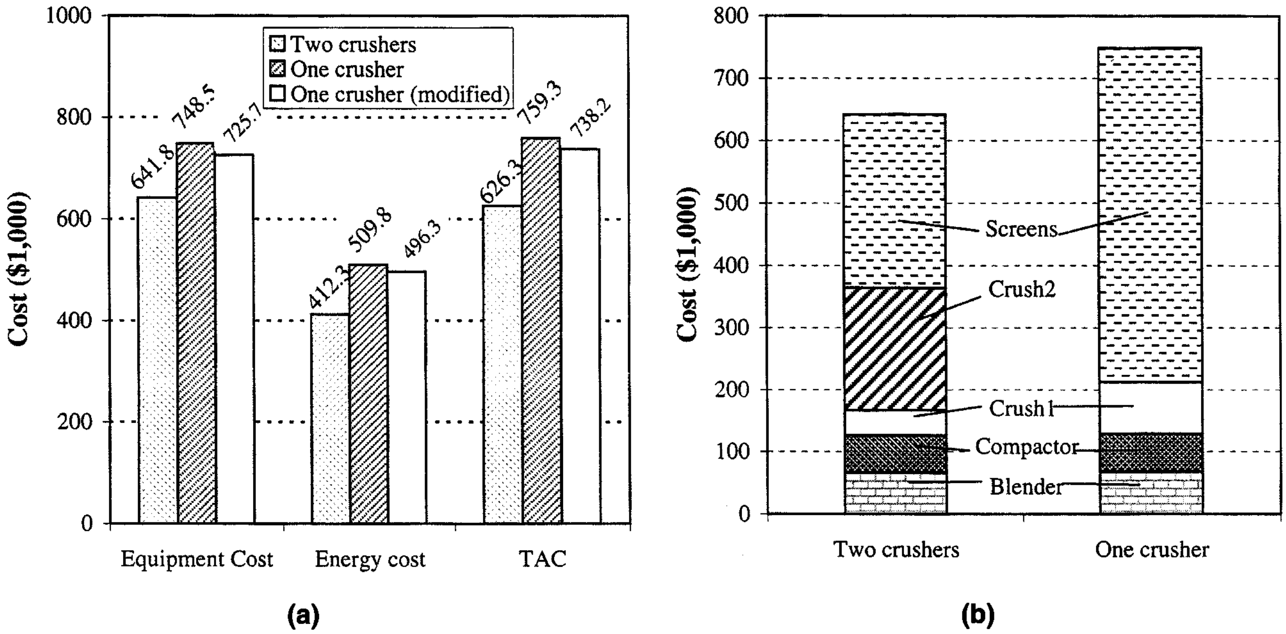


Figure 15. Comparison of costs between using one and two crushers.

(a) Total equipment cost, total energy cost, and TAC; (b) equipment cost distribution.

Sensitivity Analysis

Unlike gas and liquid systems for which simple to rigorous models have been developed, most of the models in solids processing are approximate in nature. There are two types of uncertainties in these models: structural and parametric. Structural uncertainty arises primarily because of the complexity of the underlying physics and chemistry. High resolution models with more parameters can be used to reduce structural uncertainty, for example, by replacing the parabolic breakage function with a multiple-adjustable-parameter function (Hill and Ng, 1996). Parametric uncertainty arises because of the difficulty in obtaining accurate model parameters for solids processing systems. One way to evaluate the effect of parametric uncertainty is the probabilistic collocation method, which provides the collective uncertainty in all the parameters under consideration (Tatang et al., 1997; Obrigkeit and McRae, 1998). In this work, due to the high structural uncertainty of solids processing models, a sensitivity analysis which examines one variable at a time is used.

Table 12 shows the change in TAC as compared to the TAC obtained when the residence times in the primary and secondary crusher are 2.5 and 1.8 min, respectively. These numbers reflect the relative impact of errors of $\pm 20\%$ in the value of a model parameter. The two crushers have been modeled using a functional form bearing two constants. It is evident from Table 12 that the change in S_0 for both crushers does not significantly affect the TAC. A change in the exponent k has a greater impact on the TAC. It is also clear that the TAC is more sensitive to changes in secondary crusher parameters. The number of opportunities of a particle passing a screen n does not appear to have a significant effect on overall economics. That is also the case with mixing time, which fixes the blender volume but does not affect material balance. This is not surprising since blender cost is only a small fraction of the overall equipment and energy costs (Figure 11). Bond's work index, on the other hand, has a more significant impact on the TAC.

The model used in the potash example does not account for parameters related to compaction. It has been assumed that reliable experimental data of compactor performance can be obtained. The cost distribution (Figure 11) indicates that the costs related to the compactor take up about one-half the total cost, which means that accuracy in determining compactor performance is essential.

Table 12. Sensitivity to Model Parameters

Model Parameter	Change in TAC -20%	Change in TAC $+20\%$
S_0 crusher 1 ($\min^{-1}$)	+0.38%	+0.13%
k crusher 1	+2.00%	+2.13%
S_0 crusher 2 ($\min^{-1}$)	+0.64%	+0.28%
k crusher 1	+6.30%	+2.61%
n in screen model	+0.92%	−0.64%
Mixing time (min)	−1.51%	+1.49%
W_i (kWh·tonne^{-1}·μm$^{0.5}$)	−4.33%	+4.33%

Additional Examples

Example 1: production of water dispersable granules (WDG) pesticides

The pesticide granules contain the active ingredient and an inert carrier, such as montmorillonite (McKay, 1996). Typically, the active ingredient crystals have a size around 1 mm and a specific gravity of about 2, while the carrier comes as granules of 1 to 2 mm in size and has a specific gravity of 1.2. The desired product is granules of 0.5 to 3 mm in size (Gerety, 1993) containing 10% of the active ingredient. Once diluted in water, WDG should be readily dispersed to form a suspension containing particles with a size range around 1–40 μm (McKay, 1996).

Step 1: Reducing the dosage of an active ingredient is one of the reasons to do size enlargement (Walas, 1988). To obtain granules with a certain composition, size enlargement structure should be chosen (Rule 4).

Step 2: Since it is desired to produce granules that readily disperse in water, the porosity should be low and the best option to achieve this goal is tumble agglomeration (Rule 22). Pan granulator is a good choice in this case. The product size requirement allows the use of a fluidized bed as an alternative.

Step 3: We should start with a size enlargement structure (Figure 6b). The flowsheet for this process at Step 3 is presented in Figure 16a. Since solid additives will be added, a blender is used before the pan granulator (Rule 27). Density difference between the components suggests that we use a convective blender such as a ribbon blender (Rule 9). Both components need to be crushed into particles with sizes substantially smaller than the desired granule size (Rule 28). In this case, the desired size of the particles in water suspension (1 to 40 μm) is taken as the target. Since particles in both feed streams are larger than this target, a size reduction unit would be necessary (Rule 26). The desired product size leads us to the pin disc mill for the crushing task (Figure 5). A fluid jet mill can also perform the crushing, but additional equipment, such as a cyclone, will be needed to separate the product from the carrier gas. The desired size reduction ratio (around 100–200) is still below the typical maximum value for a pin disc mill (Rule 14).

A small amount of water can be directly added to the pan granulator as a binder (Rule 25). Large particles exit from the pan granulator due to segregation, as illustrated in Figure 2d. The PSD of this exit stream should be measured. To obtain granules of a more uniform size, a classifier is needed. Since this stream would still be wet and may cause a blinding problem in the screen, a dryer should be installed after the pan granulator (Rule 23). Similar to the potash example, we also have an overflow from the classifier that should be sent to a crusher. Its overflow is then recycled to the screen (Rule 37). The screen underflow stream consisting of small particles should be recycled to the granulator (Rule 38).

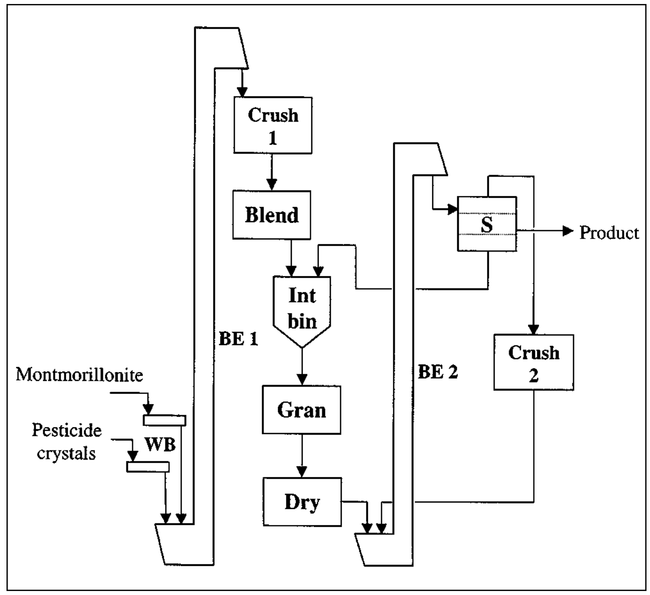


Figure 17. Final flowsheet for the WDG pesticide process.

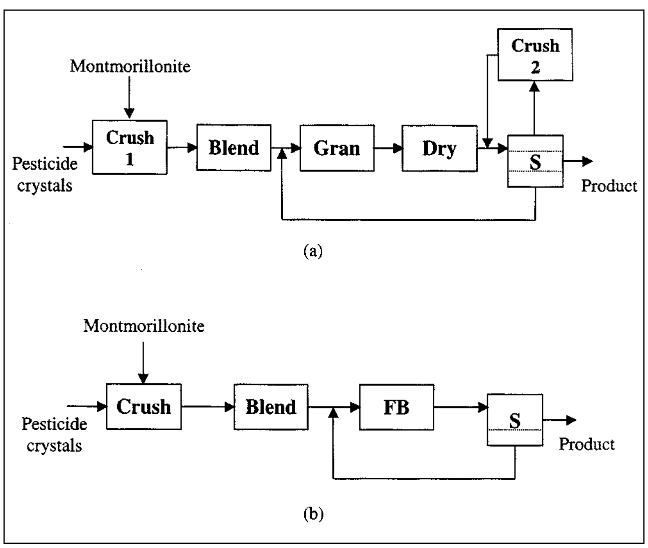


Figure 16. (a) Step 3 flowsheet for the WDG pesticide process; (b) process alternative using fluidized-bed granulator for the WDG pesticide process.

An interesting process alternative is replacing the pan granulator with a fluidized-bed granulator. The drying step after agglomeration will not be necessary in this case (Rule 24). Also, the size limitation of the granules produced via fluidization may eliminate the need for Crush 2. The flowsheet structure for this alternative is shown in Figure 16b.

Step 4: The final flowsheet for the pan granulation alternative is shown in Figure 17. An intermediate storage bin is needed for the feed to the pan granulator (Gran) to avoid starvation (Rule 40). Weighing belt conveyors (WB) are used for keeping a constant proportion between the two feed streams (Rule 44), and ordinary belt conveyors (not shown) are used for other streams (Rule 42). Bucket elevators (BE) are placed before the blender and the screen to enable gravity flow in other parts of the process (Rule 45).

Step 5: Further evaluation can be performed using simulations. Material balance calculations are performed for the two alternatives using parameters and input data as listed in Table 13. As shown in the results summarized in Table 14 and Figure 18, simulations indicate that the equipment costs are about the same for both cases. The low cost of a pan granulator is balanced by the elimination of the dryer and secondary crusher in the fluidized-bed option. The fluidized-bed option offers a slightly higher total cost, mainly due to high energy consumption for the blower. In both cases, the drying unit (drum dryer in pan granulator scheme and the fluid bed in the other scheme) dominates equipment cost. If

Table 13. Model Parameters for Water Dispersable Granule Pesticide

	Pan Granulator	Fluidized Bed
Blending		
Mixing time	10 min	10 min
Size Enlargement		
Model for specific rate of agglomeration	$a_{i,j} = a_0(ij)^l$	$a_{i,j} = a_0(ij)^l$
Breakage rate constant, a_0	5,000 kg^{-1} min^{-1}	20 kg^{-1} min^{-1}
Agglomeration order, l	-3	-3
Residence time, τ	5 min	30 min
Drying		
Type of dryer	Rotary drum	Fluid bed
Initial moisture content	15%	15%
Final moisture content	5%	5%
Drying air temperature	176°C	80°C
Blower		
Generated pressure	None	17 kPa (gauge)
Air flow rate		3.26 m^3/s
Size Classification		
Ratio c/b	1.3	1.3
Number of opportunities, n	80	80
Deck opening	Upper: 3.36 mm (6 mesh) Lower: 0.50 mm (35 mesh)	0.50 mm (35 mesh)
Secondary Crushing		
Model for specific rate of breakage	$S_i = S_0 i^k$	
Type of breakage	Parabolic $\left(b_{i,j} = \dfrac{6(ij - i^2)}{j^3 - j}\right)$	None
Material work index	8.23 $kWh \cdot tonne^{-1} \cdot \mu m^{0.5}$	
Breakage rate constant, S_0	0.012 min^{-1}	
Breakage order, k	0.86	
Residence time, τ	3 min	

the desired degree of drying cannot be accomplished in the fluid bed, the pan granulator is clearly a better option. Also, about 80% of the energy is consumed by the disc mill where micron-size particles are produced. We can also see from Table 14 that the product from the pan granulator has a considerably larger average size and a narrower PSD.

Example 2: aspirin process

Aspirin (acetylsalicylic acid) is a widely-used analgesic with a world consumption rate of about 35,000 tonne per annum (Thomas, 1994). It is manufactured by reacting salicylic acid with acetic anhydride (Porter, 1948; Edmunds, 1966). Aspirin crystals produced are filtered, washed, and finally dried to less than 0.5% moisture by weight (Thomas, 1994). The crystals are about 40 mesh (approximately 0.4 mm) in size and have a specific gravity of about 1.2. For easy consumption and fixed dosage, food starch and other ingredients are added to form 5-grain (32 mg) tablets containing about 81% aspirin. The starch (specific density = 1.53) is available as a powder of about 400 μm in size. The steps for generating the final flowsheet shown in Figure 19 are discussed below.

Step 1: The objective is to produce tablets, and the size enlargement structure should be chosen (Rule 4).

Step 2: Clearly, pressure agglomeration should be chosen to make tablets (Rule 20). A tableting machine is the only suitable equipment to perform this task.

Step 3: Since the crystals have a moisture content below 2%, additional drying before compaction is not necessary (Rule 23). A blender (Blend) is required because of the addition of starch (Rule 27). The density difference leads to the use of a ribbon blender (Rule 9), but the need to avoid contamination problems suggests that we use a tumbling mixer (Rule 7). In this case, the well-mixed state is essential, and we have to use a ribbon blender. Contamination is minimized by using stainless steel equipment. The final size of a 5-gr tablet is about 5 mm; therefore, the feed size to the tableting machine should be much smaller, say about several microns (Rule 28). Similar to Example 1, a pin disc mill is suitable. Very fine particles produced in the mill will be cohesive, especially in the presence of starch. Since cohesiveness may cause inconsistent flows, it is difficult to maintain a sufficiently constant feed rate to the tableting machine. Therefore, agglomeration in two stages is applied (Rule 29). Slugs of about 25 mm in size are produced in the first machine

Table 14. Comparison of Pan Granulator and Fluidized-Bed Systems (Simulation Results)

	Pan Granulator	Fluidized Bed
Cost ($1,000)		
Equipment	284.7	298.0
Energy	326.5	341.0
TAC	421.4	440.4
Product Quality		
d_{50} (mm)	1.792	1.117
Coef. of variance (%)	40.7	43.7
% under 0.5 mm	1.0	2.1
% above 3 mm	2.3	0.4

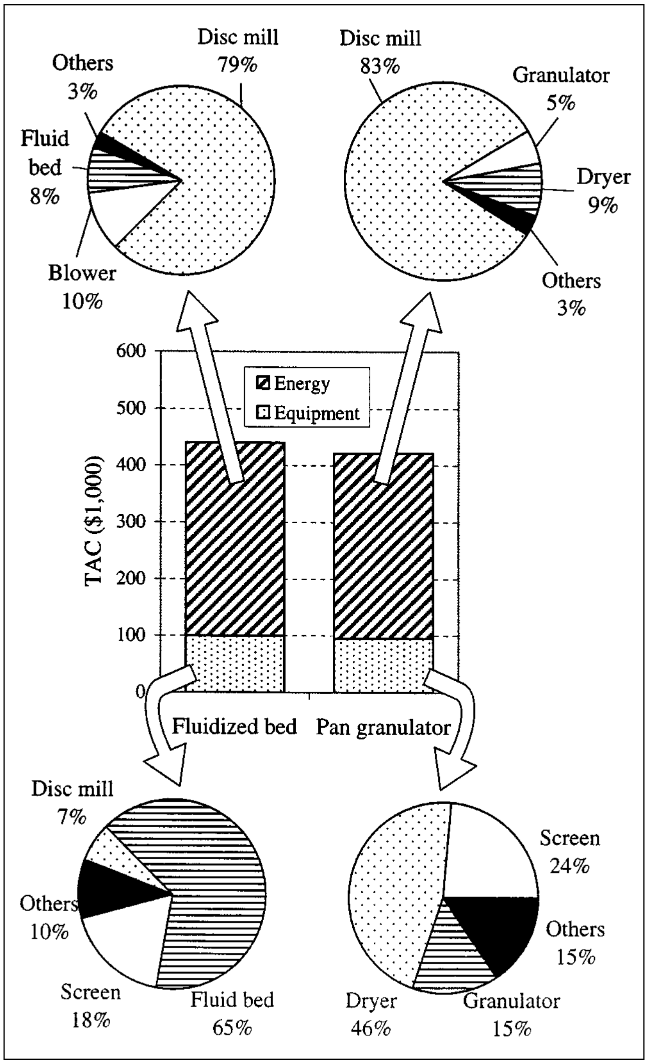


Figure 18. Cost comparison between pan granulator and fluidized-bed options for the WDG pesticide process.

(Tabl 1) (Carstensen, 1984). These slugs are then sent to a crusher (Crush 2), the effluent of which is fed to a second tableting machine (Tabl 2). A hammer mill is a suitable choice for the size reduction equipment (Figure 5).

Step 4: The input to the tableting machine should be carefully controlled to ensure production of uniform tablets, as well as to avoid any spill or starvation in the machines. For this reason, intermediate storage bins are used before the tableting machines (Rule 40). A screw conveyor (SC) is used to transport the material into the first tableting machine, because metering is necessary (Rule 44). The screw conveyor also prevents dusting problems since it is completely covered. Weighing belt conveyors (WB) are used to measure the feed flow rates. Bucket elevators may be necessary for this plant layout (Rule 45).

Step 5: Simulation can be performed if desired. In particular, we can consider the alternative of using a compactor to replace the first tableting machine, since the output must be broken anyway.

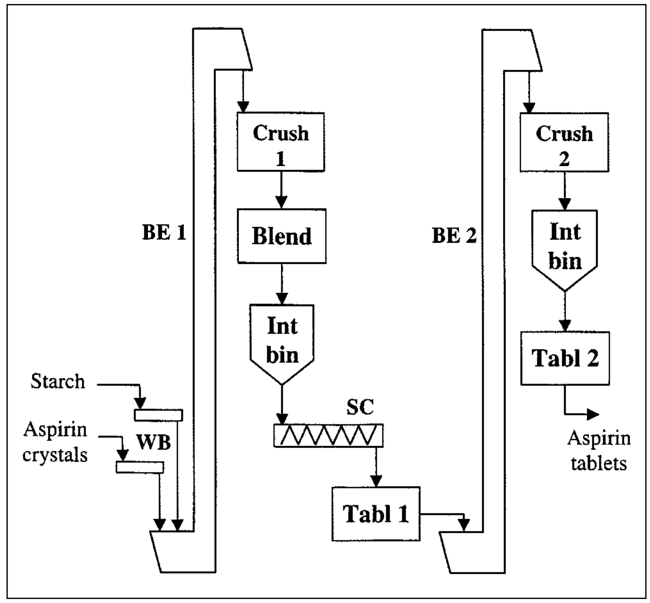


Figure 19. Final flowsheet for aspirin process.

Example 3: manufacture of azo dye

Azo dye pigments are the most versatile class of dyestuffs in industry (Guccione, 1963). The synthesis of azo dye pigments is basically a coupling reaction of diazonium salts. The slurry coming out from the reactor is filtered and dried to yield crystals of size about 0.1 to 1 mm. Several additives of approximately the same size must be added to lower and standardize the strength of the dye. It is desired to produce pigment powder about several microns in size for paint mixtures. Figure 20 shows the final flowsheet for this process, which is generated using the steps discussed below.

Step 1: Since the objective is to produce fine particles from crystals of larger sizes, the size reduction structure should be chosen (Rule 2).

Step 2: To produce micron-size particles, a fluid jet mill or a pin disc mill can be used (Figure 5). We choose a jet mill (Crush) for this process.

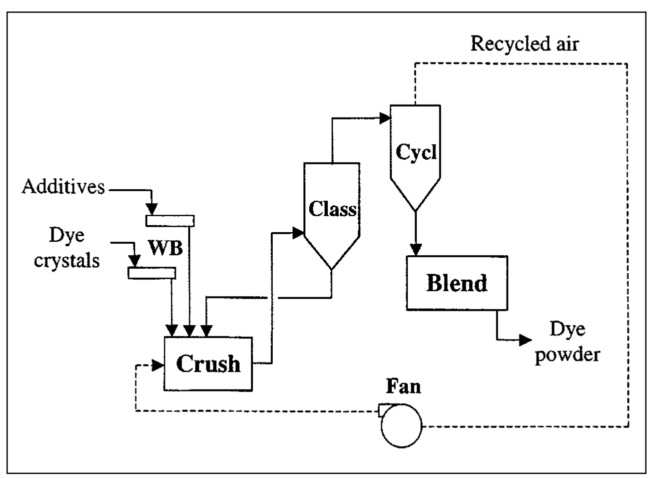


Figure 20. Final flowsheet for azo dye production.

Step 3: We now extend the structure by adding necessary unit operations around the mill. The input of the jet mill does not need to be homogeneous, and the ingredients in this process can be directly fed to the mill without being premixed. The output of the mill cannot be expected to be uniform in size, and a size classification unit must follow (Rule 31). Screens are not suitable for this particle-size range. An air classifier (Class), which has the capability of producing a cut-size as low as 2 μm (Figure 4), is used instead. The oversized particles are then sent back to the jet mill (Rule 37). The overflow stream contains particles of the desired size, which are collected in a cyclone (Cycl). Segregation in the cyclone is expected. This is undesirable because the product has to be relatively well-mixed. The problem can be resolved by putting a blender after the cyclone instead of at the beginning of the process (Rule 34). Due to the segregation tendency and cohesiveness of the particles, a ribbon blender is suitable (Rule 10).

Step 4: No intermediate storage is necessary for this process. Since we are dealing with fine particles, pneumatic conveying is suitable (Rule 46). This is in agreement with the fact that we are using air-operated crusher and classifier. For transporting the powder separated in the cyclone back to the mill, covered belt conveyors can be used (Rule 42).

Step 5: As in other examples, a simulation can be performed for further evaluations. Luckie (1997) describes a model for air classification, which can be used to predict the PSD of the overflow and underflow.

Conclusions

Solids processing systems have received relatively little attention in the chemical engineering process design literature. A five-step procedure supported by rules and simulations is presented for the synthesis of bulk solids processing systems. There are several limitations to this study. This work considers PSD as the only required solids attribute. In reality, other factors such as shape and impurity inclusions can also be important requirements. In the simulation, it is assumed that agglomeration and breakage do not occur to an appreciable extent in blenders, classifiers, and solids handling equipment; this may not be the case for many real systems. Nevertheless, the principal phenomena of bulk solids processing have been captured in our procedure, so that it can be used for comparing process alternatives. The proposed framework can readily accept more rigorous and detailed calculations to obtain more accurate estimates.

We did not consider changes that can be made in the upstream processes. However, bulk solids processing is only part of a complete solids processing system. This study needs to be extended to include other parts of the process, such as solids separation, reaction, and crystallization. This is important because changing the operating conditions in an upstream unit might have a significant impact on the bulk solids processing step. For example, controlling the PSD of the crystals from the crystallizer by applying certain operating policies may eliminate the need for a crusher or an agglomerator. An integrated procedure involving reaction, crystallization, solid-fluid separation, solid-solid separation, and bulk solids processing can lead to an even better design. Most of these efforts are now underway.

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Notation

$a_{i,j}$ = specific rate of agglomeration between a particle in size interval i and a particle in size interval j, $kg^{-1} \cdot min^{-1}$
b = dimension of square openings, mm
$b_{i,j}$ = mass fraction of material broken from size interval j which appears in size interval i
c = diameter of screen wires, mm
d = representative particle size or diameter, mm
E = energy requirement, $kWh \cdot tonne^{-1}$
k = breakage order, dimensionless
l = agglomeration order, dimensionless
m = total number of input streams in Table 8
n = total number of size intervals (equal-sized) in Table 8
n = number of opportunities given to a particle for passing through a screen (Eq. 4)
p = probability that a particle approaching a screen will pass through, dimensionless
P_i = probability that a particle in size interval i ends up in the underflow, dimensionless
q = probability that a particle approaching a screen will pass through after several opportunities, dimensionless
S_i = specific rate of breakage of particles in size interval i, min^{-1}
S_0 = breakage rate constant, min^{-1}
W_f = feed flow rate, $kg \cdot min^{-1}$
W_p = product flow rate, $kg \cdot min^{-1}$
τ = residence time, min
ω_{fi} = weight fraction of particles in size interval i in feed stream
ω_i = weight fraction of particles in size interval i in product stream

Subscripts

f = feed
i, j, k = indices
p = product

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