A New Macrophenomenological Concept of Comminution in Ball Mills

A P Yovanovic[1] and H P Moura[2]

ABSTRACT

Grinding is probably the single unit operation that requires the largest capital investment and operating costs in an ore beneficiation plant. Because of that, a great deal of research has been concentrated in the theory, design and operation of grinding equipment, in particular ball mills.

To-date, the models developed to describe grinding in tubular mills are macrophenomenological and can be divided into two categories: energetic and kinetic models.

This paper presents a new approach to modelling ball mill grinding: The Operational Model. The theoretical aspects of the model as well as a comparison with existing ones are discussed. The concepts of the model, summarised in easy-to-operate PC programs, permit the calculation and design of new ball mills, and a critical evaluation and optimising of mills in operation.

THE STATE OF THE ART

The general structure of modelling ore dressing processes considers, essentially, the description of the main process phenomena, the unit operations, and the auxiliary phenomena involved (classification, transport, etc in the specific case of grinding).

The main phenomenon can be decribed by the following approaches:

- Microphenomenological models that represent the process either chemically or physically;
- Statistical models that consider only the results of the process, and are based on a regression of a series of experiments instead of describing the process itself; and
- Macrophenomenological models that describe macroscopically a certain physical aspect of the process. For example: kinetic laws, separation curves, residence time distribution, among others.

To-date, the existing models of grinding in tubular mills are of the macrophenomenological type (Broussaud, 1988), and can be divided into two categories: energetic and kinetic models.

Both categories consider the evaluation of a certain macrophenomenological property measured either in laboratory or in pilot plant experiments. The design of the industrial mill is then carried out assuming the scaling up of this property, while the same approach should be followed when attempting to optmise industrial mills.

The grinding process involves three complex and simultaneous phenomena: comminution itself, which is the main phenomenon, (impact, attrition and abrasion), the auxiliary phenomena of classification of particles, and the macromolecular transport.

Energetic approach

For a determined particle size range, the third theory of Bond (1961) establishes the relationship between the energy applied to the mill and the amount of energy the ore actually receives. For this same range, Bond presents a quantity known as the 'Work Index' (Wi, kWh/st). This constant, which can also be extrapolated to industrial mills, is an estimate of the net power consumption that shall be applied to the mill to reduce the actual ore sizing to an established size range.

This essential property, the 'Work Index', is obtained in laboratory mills working batchwise. This laboratory experiment was standardised by the simultaneous operation of an industrial mill used in the establishment of the Bond theory. Therefore, one may assume that in the design of a mill identical to the Bond reference mill (8.5 ft external diameter, 8.0 ft length, 70 per cent of critical speed, 37 per cent of ball load, pulp density of 77 per cent, etc) the laboratory experiments would lead to a very accurate prognostic of the ore behaviour in industrial scale.

Recent studies (Deister, 1987; Magdalinovic, 1989) have presented interesting suggestions to improve the reliability and scale-up of the parameters derived from Bond type experiments in laboratory.

Austin and coauthors (1982) point out four main disadvantages in the Bond method of calculation:

1. The specific power consumption differs from the calculated value if the reference conditions are changed: circulating load, ball charge, rheologic conditions, etc;
2. Bond clearly states that the specific grinding energy is not a function of the ball load, which has been proved inexact;
3. Bond's method uses only the F80 and P80 particle sizes, whereas mill capacity, in general, depends on the shape of the feed and product size distributions;
4. Under some conditions the Bond method does not clearly show the reasons for mill operational inefficiency.

All these shortcomings, except item 3, have been taken into account in the development of the Operational Model.

Kinetic approach

Batch experiments carried out in bench scale, within specific ranges of particle size, permit establishing the basic components of the kinetic approach (Broussaud, 1988):

B = Breakage function

where 'bij' represents the mass fraction of the particles originally in size interval 'j' and reduced to size interval 'i' after breakage. It can be seen that 'bij' is similar to a stoichiometric coefficient of a chemical reaction.

S = Selection function

1. Civil-Chemical Engineer, Senior Mineral Processing Engineer, LEME Engenharia Ltd, Belo Horizonte, Brazil.

2. Mechanical Engineer, Senior Mineral Projects Consultant, LEME Engenharia Ltd, Belo Horizonte, Brazil.

where 'Si' represents the fractional rate of breakage of the particles in size interval 'i'. 'Si' is a function of time, and corresponds to a kinetic constant of a first-order chemical reaction. This linear behaviour is mandatory in the kinetic approach.

Readers should note that these two functions bring together the concepts of chemical kinetics and the physical phenomenon of comminution (Franks, 1972; Lynch, 1977; Austin and coworkers, 1982; Austin, 1984; Herbst, 1987).

Restrictions to the kinetic approach are: linear behaviour of the selection functions (very difficult to occur in practice), direct scale-up of the kinetic specific constants from batch processes to steady state conditions, and direct extrapolation without any boundary conditions of the specific breakage function as related to the mill size.

THE NEW MACROPHENOMENOLOGICAL CONCEPT

A determined amount of contacts between the ore and the grinding media, per force unit, allows the comminution of the material up to the size established.

Depending on the residence time of the particles and the mill operating conditions, specifying the most probable amount of contacts between the grinding media and the ore and the intensity of these contacts, the Operational Model proposes that the phenomenon may be expressed by the following equation:

$$IC = \left(\frac{T}{TM}\right) * BL * RM * DI * FV, \ \left(\frac{\text{Power–hour}}{\text{ton}}\right) \tag{1}$$

where IC is the 'Comminution Index', which means the amount of contacts among the ore and the balls, per force unit, and:

- T - Average residence time of the ore (dry material) inside the mill, expressed in minutes (continuous process);
- TM - feed rate, expressed in metric tons per hour;
- BL - weight of the ball charge inside the mill, in metric tons;
- RM - mill speed, in rpm;
- DI - internal diameter of the mill, in ft;
- FV - critical speed ratio.

The mechanical approach that provides energy to the system can define the optimum mechanic-operating conditions to let the ore reach the 'Comminution Index' wanted.

It is possible to establish the Basic Comminution Index that is required to make an ore reach the P80 condition from an F80 condition considering Bond's experiments in laboratory, experiments in pilot plants or comparative experiments with industrial mills.

Figure 1 shows the relationship between the Comminution Index and P80, for F80=10700 micrometers, for different materials with different hardnesses.

Scale-up

Even without considering the auxiliary phenomena to comminution, the scale-up of the phenomenon to bigger-diameter mills has brought about serious difficulties up to now (Arbitrer and Harris, 1982; Lo, Herbst, Rajamani and Arbitrer, 1988; Rowland, 1988).

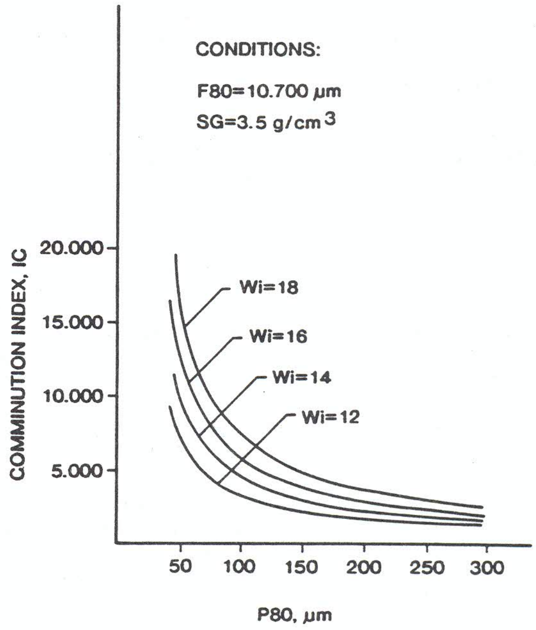

FIG 1 - Relationship between the comminution index and P80.

Bond's energetic approach of the phenomenon is improved by the application of certain correction factors, based on experiments as Mular and Bhappu (1980) have showed. Moreover, various authors (Austin, Klimpel, Luckie and Rogers, 1982; Herbst and Rajamani, 1982; Lo, Herbst, Rajamani and Arbitrer, 1988; Herbst and Lo, 1989) find that the kinetic approach considers a proportional relation between the specific selection function and the diameter of the mill, without limits.

Austin (1984) and Austin and coauthors (1984) define the limit breakage conception, or limit breakage force above which the fracture of a particle keeps constant. This is assumed as true by the Operational Model.

The Comminution Index, defined by Equation 1, considers a combination of factors where the difference between the laboratorial and industrial diameter is only a part of this system. It is also interdependent of energy (mechanical power), transportation and classification equations.

Although in a preliminary way, the model defines the maximum diameter of a mill in relation to the hardness of the material, from which it is noticed Austin's concept. An unpublishad report by one of the authors (Yovanovic, 1989) describes an experimental procedure required to define this limit diameter, in the laboratory.

Opposite to up-to-date practice, computer simulations using the percentage of critical speed suggested by Mular and Jergensen (1982) show that the energetic efficiency varies in an inverse proportion to the mill diameter (the power consumption increases), starting from the minimum diameter that allows top-size breakage. This condition reaches its important point when the maximum diameter established is surpassed, as shown in Figure 2.

A diameter smaller than the necessary one makes the top-size breakage difficult and requires more time to reach the established size (producing a smaller comminution effect per time unit, or more energy). The choice of the maximum and minimum diameters allows establishing an interval, in which the efficiency of the grinding process is almost independent of the mill diameter (indifferent energetic zone) to a determined process (Comminution Index defined to grind the material from an F80 pattern to a P80).

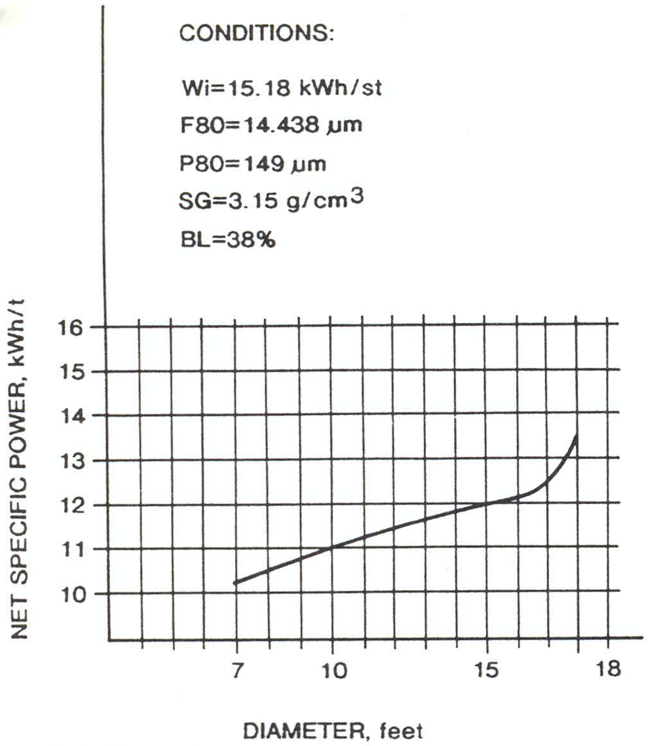

FIG 2 - Net specific power variation as a function of the mill diameter.

In industry, because of investment costs, it is generally specified a mill with a diameter as large as possible. Lack of knowledge of Austin's concept leads to mills with diameters bigger than necessary, as happened in the Bouganville plant (Arbitrer and Harris, 1982; Lo, Herbst, Rajamani and Arbitrer, 1984).

The Comminution Index, defined by Equation 1, measures the energy that the material receives or must receive. The ore, in any equipment or under any operating condition, requires a determined Comminution Index to be comminuted from 'F' size to 'P' size, where:

$$IC\ (basic) = IC\ (industrial) \quad (2)$$

assuming F80 and P80 as constant.

Steady state

Various authors (Rogovin, Casali and Herbst, 1988) insist in utilising chemical engineering concepts to obtain a residence time distribution inside the mill, taking as a basis the chemical tracer diffusion phenomenon, and not the physical thickening phenomenon inside the mill.

The Operational Model considers the existence of a hydraulic classification inside the mill, which is enough to operate, even in open circuits, for determined process conditions (relatively big-sized product, low level of grinding media, etc).

Experiments in Norway (Forsund, Norkyn, Sandvik and Winther, 1988) to grind iron ore, show that the internal classification works properly from a 22 per cent of ball charge on, producing the same product size than another mill operating in closed circuit, and still spending 16 per cent less net power.

In the operation of industrial mills, that fact can be clearly observed. When a system starts to operate, it takes some time before reaching the steady state (from 10 to 30 minutes, depending on the process). At the starting point of the overflow process, the density of the pulp in discharge is almost the same of the water and keeps increasing up to the same value of the feeding pulp (steady state).

That fact creates an increase of the pulp density inside the mill (which means an increase of solid concentration), as a result of decantation or hydraulic classification of the particles. It takes to a balanced point of the steady state that distributes the residence time.

Other experts (Klimpel, Austin and Hogg, 1989), trying to explain the effect of powder filling in continuous operations, had to change the original concept that attributed the rise of the pulp level inside the mill to the circulating load. They have verified, experimentally, that density inside the mill raises and pulp level keeps unchanged in mills with overflow unloading.

The Operational Model has developed some specific equations to describe this secondary phenomenon, using many variables which can improve or not the hydraulic classification. When the variables do not contribute to the improvement of the classification, the Model presents the limit conditions above which a closed system should be used to complement the classification process not satisfactorily completed by the equipment.

The Operational Model proposes an empirical equation to the calculation of the solid concentration in the pulp inside the mill, as shown below:

$$SI = f(SP,FE,W,GP,GM,P80) \quad (3)$$

where:

- SP - solid concentration ratio in the feeding;
- FE - thickness factor, in ton/kWh (this is a function of the solid concentration fed to the system and the thickness of the free layer passing through the mill);
- W - net power consumption, in kWh/ton (it considers indirectly the relative residence time of the comminuted particles);
- GP - specific gravity of the harder and heavier ore particles (in case of very heterogeneous ores);
- GM - average specific gravity of the whole material;
- P80 - size interval of the product, in micrometers.

So, the total average solid residence time can be calculated as follows:

$$T = f(VP,DP,SI,TM), \text{ in minutes} \quad (4)$$

where:

- T - average residence time of the material (dry);
- VP - pulp storage capacity inside the mill, in m^3;
- DP - pulp density inside the mill , in ton/m^3;
- TM - feeding rate, in dry ton/h.

It is important to remember that 'T' appears in the 'Comminution Index' equation (Equation 1).

When the hydraulic classification conditions are not effective, or when more than 20 per cent of the particles discharged by the mill has a size bigger than the required P80 interval, the model determines the fundamental equations of the auxiliary classification process.

Auxiliary classification

The Operational Model accepts the hydraulic classification as an auxiliary phenomenon that happens inside the mill. This phenomenon has been industrially proved good for ball charge levels not higher than 22 per cent of the mill internal volume. Above this level, the hydraulic classification error is mathematically estimated as a function of the considered ball charge level, the solid concentration, the size interval of the product and the thickness of the free pulp layer passing through the mill.

Using pratical varied industrial equipment experiments, the Operational Model estimates that, changed to P80 near 100 mesh, the absolute error of the internal classification of a high ball charged mill (40 - 45 per cent) should not be over 45 per cent. It means that in case of opening the circuit and reaching a new steady state, with the same production capacity, 45 per cent of the discharged particles can be bigger than the determined P80 size interval. On the other hand, it is known that the classification absolute error of a hydrocyclone, operating in open circuit, is near ten per cent (90 per cent of efficiency). Consider:

$$E = Em * Ec \tag{5}$$

where:

- E - total efficiency of the system theoretically operating in open circuit;
- Em - absolute mill efficiency;
- Ec - absolute classifier efficiency.

It has been mathematically demonstrated that when the circuit is closed, the circulating load in steady state is expressed by:

$$C = \frac{(1-E)}{E} * 100 \tag{6}$$

Assuming the efficiency values of Em = 0.55 and Ec = 0.9 applied to Equation 5, and substituting 'E' value in Equation 6, it is obvious that the circulating load should be near 100 per cent in the reference mesh, at worst.

The excess of circulating load observed in almost all the grinding industrial circuits, is due simply to the hydraulic error of the cyclone classification system, whose 'apex/vortex' combination should be hydraulically satisfied when the steady state is reached. Sometimes, it has been noticed that the classifier determines the mill performance and not the opposite, as would seem appropriate.

Increasing the ball charge to increase the performance capacity of the mill implies in closing the circuit from around 22 per cent of charge on (depending on many variables). To compensate the hydraulic classification error, the use of an auxiliary equipment is required (cyclone, hydroclassifier, screen, etc). To be used in Equation 6, the circulating load must increase as the ball charge level increases. Measuring the size interval of the product in an open circuit, for the same amount of ore processed in a closed circuit, it is possible to verify the absolute efficiency of the mill hydraulic classification (Em). And, using both Equations 5 and 6, the circulating load required by the system is recalculated.

Steady state in closed circuits

In closed grinding circuits, the classification is performed by an auxiliary equipment (cyclone, hydroclassifier, screen), and the mass transport inside the mill includes the circulating load of the classification system. This load, when increased, changes the time distribution pattern inside the mill, which gets closer to the so-called 'fully mixed system', where the pulp presents an almost homogeneous distribution.

In closed circuits, the average residence time of the ore (dry material) is given by:

$$T1 = T/FM \tag{7}$$

where FM = feed ratio, related to the new feeding, given by the expression:

$$FM = 1 + \frac{C}{100} \tag{8}$$

The total comminution efficiency of the system to produce and classify fine particles (below P80) can be expressed, in steady state, as follows:

$$Et = \frac{em * ec}{1 - em(1 - ec)} \tag{9}$$

where:

- em - ratio between the amount of particles below P80 (in ton/h) in the discharge and particles fed to the mill, including the circulating load;
- ec - ditto, between the overflow discharge of the classifier equipment and its feeding (dynamic efficiency).

It has been demonstrated that this Continuity Equation determines an almost constant 'Et' value for circulating loads ranging from 100 per cent to 500 per cent. It shows a kind of indifference of the process when operating in this interval, pointing out the hydraulic error of the pulp partition in the classifier.

The total efficiency of the system can be slightly improved when the circulating load is reduced to values next to the ones defined by Equation 6. It also allows a decrease in investment and operating costs in classification equipment.

On the other hand, the decrease in the circulating load causes an increase in the overflow pulp density, which reduces the thickening necessity, maximises the residence time in further unit operations and also reduces water consumption and pumping, along the whole process.

The mechanical equation

All conventional models consider the following mill mechanical equation:

$$HP = f(L,DI,FV,DC) \tag{10}$$

where:

- L - mill size;
- DI - internal diameter of the mill;
- FV - critical speed ratio;
- DC - average density of load inside the mill.

Normally, for an L/D ratio compatible with the mechanical resistence possibilities of the cylinder and, mainly, providing the transportation and hydraulic classification conditions according to the size intervals necessary to the process (eg L/D > 1 for fine grinding), the mill size appears as a consequence of the previous variables. Its mathematical expression is connected with the ore energetic necessities, as shown below:

$$HP(net) = 1.341 * W * TM \qquad (11)$$

where, W = net power consumption, in kWh/ton.

The load density is the calculation item that causes the greatest difficulty, and the known models consider as mill charge only the balls, which brings problems to the design of mills that operate with less than 30 per cent of ball charge.

The Operational Model considers as mill charge: the ball charge, the water and the material to be ground, including also the thickening effect inside the equipment. The Model permits the power calculation of the mill to every charge level (even the semi-autogenous).

Optimising of grinding operations

Summarising, the material needs certain amount of energy to pass from an 'F' condition to a 'P' condition, concerning size intervals. The energy is applied mechanically to the mill, which, in optimum conditions, must pass the maximum of this energy to the ore.

Despite the rheological approach, as various experts (Gottschalk and Husemann, 1989; Kawatra and Eisele, 1988; Kawatra, Eisele, Zang and Rusesky, 1989; Rule, Fergus and Daellenbach, 1985) showed, the Operational Model established an equation system that considers: the characteristics of the ore (hardness, size, etc), of the equipment (power, diameter, speed, etc) and, the most complicated, the way these two elements combine themselves in the mill to reach the process necessities (operation). This equation system tries to quantify how the ore is receiving the energy applied to the system.

There is an optimum point between the energy applied and the energy received, that leads to smaller operating costs and smaller net power consumptions. The auxiliary transport, rheology effects, and classification phenomena are part of this system, and contribute to improve the receptance of energy by the ore, as shown in Figure 3.

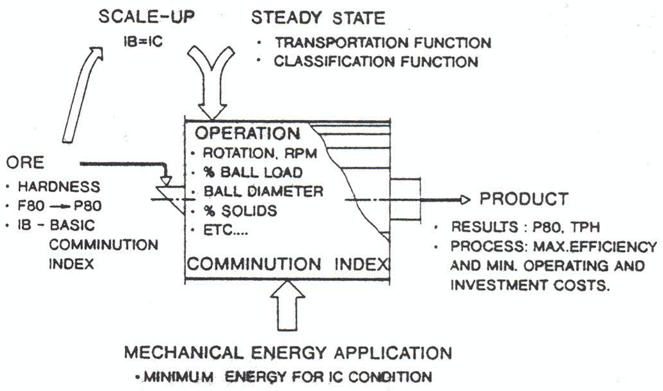

FIG 3 - Sketch of practical application of the grinding operational model.

APPLICATIONS

An important part of the model has been proved by resolution of its calculation steps using a personal computer. In order to do that, operational data of some known industrial mills were used.

Ball charge level

Depending on ore conditions and operational and mechanical characteristics of the mill, the Operational Model does show that there is an optimum ball charge level.

One specific industrial situation simulated in a computer is represented in Figure 4. Other simulations for different types of ore, operational conditions and mill sizes lead to the conclusion that the optimum ball charge level ranges normally between 16 per cent and 25 per cent.

Nomura and Tanaka (1989) concluded that the optimum value of ball charge is around 40 per cent of the maximum mill capacity.

Various authors (Austin, Klimpel, Luckie and Rogers, 1982; Austin, Klimpel and Luckie, 1984) point out, based in small-scale experiments, that despite the existence of the maximum grinding capacity at 40 - 45 per cent of ball charge, the net power consumption reaches its minimum in the 15 - 20 per cent range. This statement, based only on experimental observations, is mathematically proved by the Operational Model.

In Germany, the usual grinding process considers the ball charge level ranging from 22 per cent to 30 per cent, while in the United States this level is near 36 per cent.

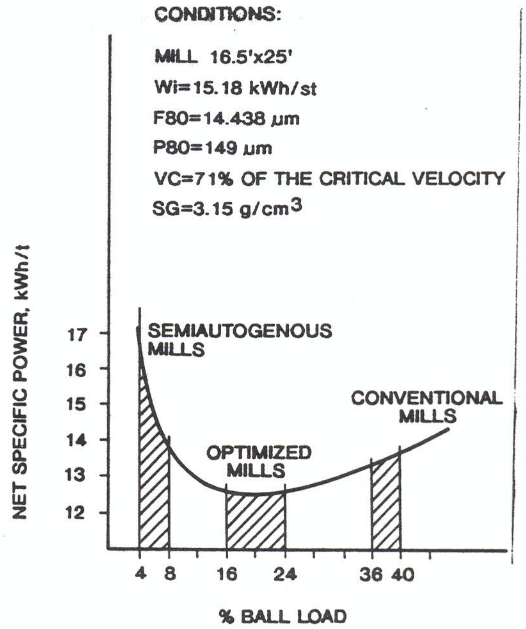

FIG 4 - Optimum ball load percentage.

The velocity of rotation

Figure 5 shows the results of the computer simulation of different conditions. It is clearly shown the existence of a critical point, from which the net power consumption per each new added ton of material fed to the system is almost doubled. The critical points are the same for different operational conditions.

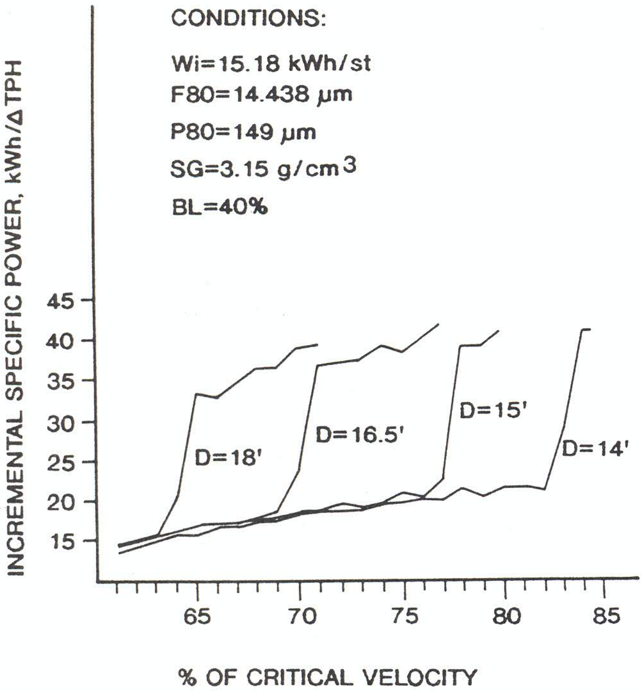

FIG 5 - Rotation velocity effect on the specific energy consumption.

CONCLUSIONS

By applying the concepts of the Operational Model, it is possible:

- To prove that there is an optimum ball charge to minimise the net power comsumption. The model demonstrates that the optimum ball charge lies in the 16 per cent to 22 per cent range of the mill volume, which is contrary to the levels prevailing in the industry today;
- That a reduction of the net power comsumption is not obtained by increasing the mill diameter, but rather is a function of the critical speed employed. For example: for a 16.5 ft diameter mill, it is demonstrated that the net power consumption increases markedly above a critical speed of 68 per cent, for a ball charge of 40 per cent;
- To minimise unnecessarily high circulating loads. The model recommends that circulating loads do not exceed 100 per cent of the feed to the mill.

In addition, the model considers the classification of particles inside a ball mill as an auxiliary phenomenon to comminution, while the operation in open circuit proves these interesting benefits.

REFERENCES

Arbitrer, N and Harris, C C, 1982. Scale-Up and Dynamics of Large Grinding Mills. A Case Study, *Design and Installation of Comminution Circuits*, Cap 26, (Society of Mining Engineers of the AIME: New York).

Austin, L G, 1984. Concepts in Process Design of Mills, *Mining Engineering*, p 628-35.

Austin, L G, Klimpel, R R, Luckie, P T and Rogers, R S C, 1982. Simulation of Grinding Circuits for Design, *Design and Installation of Comminution Circuits*, Cap 19, (Society of Mining Engineers of the AIME: New York).

Austin, L G, Klimpel, R R and Luckie, P T, 1984. Process Engineering of Size Reduction: Ball Milling, (Society of Mining Engineers of the AIME: New York).

Bond, F C, 1961. Crushing and Grinding Calculations, Processing Machinery Department, Allis Chalmers Manufacturing Company, Milwaukee, Wisconsin.

Broussaud, A, 1988. Advanced Computer Methods for Mineral Processing: Their Function and Potential Impact on Engineering Practices. XVI International Mineral Processing Congress. (Ed: E Forssberg), (Elsevier Science Publishers BV: Amsterdam).

Deister, R J, 1987. How to Determine the Bond Work Index Using Lab Ball Mill Grindability Tests, *Engineering & Mining Journal (E&MJ)*, pp 42-5.

Forsund, B, Norkyn, I, Sandvik, K L and Winther, K, 1988. Sydvarangers 6.5m diameter x 9.65 m Ball Mill. XVI International Mineral Processing Congress, (Ed: E Forssberg), (Elsevier Science Publishers BV: Amsterdam).

Franks, R G E, 1972. Modeling and Simulation in Chemical Engineering, (Chapter VIII, Staged Operations), (John Wiley & Sons: New York).

Gottschalk, J and Husemann, K, 1989. On Characterization of Conveying Processes in Ball Mills Rheological Parameters, *Powder Technology*, 58:131-6, (Elsevier Seguoia: Amsterdam).

Herbst, J A, 1987. Developments in the Use of Modeling and Simulation Techniques for the Improvement of Mineral Processing Operations, *AIME - Proceedings of the Fifth Extractive and Process Metallurgical Fall Meeting*. Mathematical Modelling of Materials Processing Operations, California, USA.

Herbst, J A and Rajamani, K, 1982. Developing a Simulator for Ball Mill Scale-Up. A case Study, *Design and Instalation of Comminution Circuits*, Cap 20, (Society of Mining Engineers of the AIME: New York).

Herbst, J A and Lo, Y C, 1989. Grinding Efficiency with Balls or Cones as Media, *International Journal of Mineral Processing*, pp 141-151, (Elsevier Science Publishers BV: Amsterdam)

Kawatra, S K and Eisele, T C, 1988. Rheology Effects in Grinding Circuits, *XVI International Mineral Processing Congress*. (Ed: E Forssberg), (Elsevier Publishers BV: Amsterdam).

Kwaatra, S K, Eiselle, T C, Zhang, D X and Rusesky, M T, 1989. Temperature Effect on Grinding Circuits Performance, *Minerals & Metallurgical Processing* pp 85-7.

Klimpel, R C, Austin, L G and Hogg, R, 1989. The Mass Transport of Slurry and Solid in a Laboratory Overflow Ball-Mill, *Minerals & Metallurgical Processing* pp 73-8.

Lynch, A J, 1977. Mineral Crushing and Grinding Circuits. Their Simulation, Optimization, Design and Control, *Developments in Mineral Processing*, (Ed: D W Fuerstenau), (Elsevier Scientific Publishing Co).

Lo, Y C, Herbst, J A, Rajamani K and Arbitrer, N, 1989. Design Considerations for Large Diameter Ball Mills, *International Journal of Mineral Processing*. Recent Advances In Comminution, (Eds: K Rajamani and J A Herbst), (Elsevier Science Publishers: Amsterdam).

Magdalinovic, N M, 1989. Calculation of Energy Required for Gringind in a Ball Mill, *International Journal of Mineral Processing*, 25:41-6 (Elsevier Science Publishers BV: Amsterdam).

Magdalinovic, N M, 1989. A Procedure for Rapid Determination of the Bond Work Index, *International Journal of Mineral Processing*, 27:125-32 (Elsevier Science Publishers BV: Amsterdam).

Mular, A L and Bhappu, 1980. SME - Mineral Processing Plant Design. 2nd Edition..

Mular, A L and Jergensen, G V, 1982. *Design and Instalation of Comminution Circuits*. (Society of Mining Engineers of the AIME: New York).

Nomura, S and Tanaka, T, 1989. Analysis of Mill Capacity Using a Theoretically Derived Selection Function Applied to Ball and Hardgrove Mills, *Powder Technology*, 58:117-24.

Rogovin, Z, Casali, A and Herbst, J A, 1988. Tracer Study of Mass Transport and Grinding in a Rod Mill, *International Journal of Mineral Processing*, 22 (Elsevier Science Publishers BV: Amsterdam).

Rowland, C A, 1988. Large Ball Mills - Length and Diameter, *XVI International Mineral Processing Congress*, (Ed: E Forssberg) (Elsevier Science Publishers BV: Amsterdam).

Rule, A R, Fergus, A J and Daellenbach, C B, 1985. Influence of an Organic Polymer in Ball-Mill Grinding of Quartz, Dolomite and Copper Ore, *Report of Investigations 8964*, Bureau of Mines, USA.

Yovanovic, A P, 1988. The Operational Model: A New Macrophenomenological Approach to Modelling Mineral Processing Unit Operations. (Unpublished paper), (Ed: B Horizonte), Brasil.

Yovanovic, A P, 1989. Construction of an Impactometer laboratory equipment for comminution test - Experimental procedure, (Unpublished paper), CVRD/SUTEC/DECEG.